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Identifying the Basic Conditions for Economic Generation of Public Electricity from Surplus Bagasse in Sugar Mills

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IDENTIFYING THE BASIC CONDITIONS FOR ECONOMIC GENERATION OF PUBLIC ELECTRICITY FROM SURPLUS BAGASSE IN SUGAR MILLS

A Study Prepared for the World Bank

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ABSTRACT

Bagasse, the fibrous cane residue from the process of sugar juice extraction, is a traditional energy source for the world's sugar mills, which burn it to generate process steam and power. Current annual sugar production worldwide is estimated to produce about 10 million tons of bagasse in excess of the sugar plants' normal requirements. This amount of fuel could substantially increase the electricity generated for public use. However, cane processing efficiency varies widely from mill to mill, largely depending on the type and age of equipment used, with the result that some mills today have substantial amounts of excess bagasse while others require supplementary fuel for their operation.

The present study identifies several ways, all using presently available technology, to greatly increase the overall energy efficiency of existing mills, produce surplus bagasse and generate electricity for sale to the grid. These include installing pre-evaporators to conserve steam, drying wet bagasse with flue gases to improve combustion efficiency, installing high-pressure boilers to increase steam generation efficiency, and pelletizing or compressing bagasse to enable it to be stored and used beyond the harvest season. Computer simulations were made to study the impact of these factory improvements under various base conditions. In most of the scenarios studied, the economic returns are well over 10 percent at an electricity selling price of US$.06/kWh.

The economics of co-generation using bagasse have changed dramatically as the costs of conventional electricity generation have increased. The guidelines developed by the study should be useful in identifying specific conditions under which production of public electricity from sugar mills is specially worth pursuing.

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IDENTIFY THE BASIC CONDITIONS FOR ECONOMIC GENERATION OF PUBLIC ELECTRICITY FROM SURPLUS BAGASSE IN SUGAR MILLS

Executive Summary

Cane Sugar is grown in 79 Countries within the tropical and sub-tropical belts with some incursions in the warmer areas of the temperate zones. Most of the countries where sugar cane is grown are underdeveloped and suffer from a lack of fossil fuels. The high cost of energy is a limiting factor to the development of these areas.

The Sugar Cane Industry has been able to survive in these countries because it is self sufficient in energy. Bagasse, the fibrous residue of the juice extraction process amounting to 25-30% of the cane weight, provides the fuel necessary for the production of the steam, which in turn produces the mechanical and electrical power needed for the process. In the majority of cases there is sufficient bagasse for that purpose.

The great majority of sugar cane factories were in operation decades before the 1973 oil crisis and were designed at a time when no attention was paid to alternate sources of energy. Hence, most sugar factories were designed to recover just enough energy from bagasse to meet their energy needs. In general the average sugar factory produces 12-15 kW/ton of cane processed and uses about the same amount of energy. This situation has changed radically since 1973 and today most nations are actively seeking to develop alternate sources of energy. Bagasse, due to its wide distribution and abundance in many under-developed countries, is a prime candidate for replacement of fossil fuels. A modern sugar factory designed to use the minimum amount of steam and recover the maximum amount of energy from its bagasse can produce about 50 kWh per ton of cane processed. In Hawaii there are some examples of factories producing as much as 70 kWh per ton of cane processed.

The main reasons for the low production of electrical power in the average sugar factory are:

--use of low steam pressure
--use of back pressure turbo-generators
--high steam consumption for processing

Modern factory designs, and improvement to old factories, make it possible to generate much more power by using higher steam pressures, condensing turbo generators and pre-evaporators for reducing steam usage.

The world bagasse production amounts to nearly 113,000,000 metric tons. It has been estimated that, of this amount, 10,240,000 metric tons are surplus. Converting the surplus of bagasse into power would produce 4.5 billion kWh of electrical energy. Much more could be produced by improving the thermal balance of the older factories, raising their steam pressure and replacing back-pressure turbo generators with condensing extraction units. Drying the bagasse using flue gases would further increase their potential by approximately 20%.

It is clear therefore that the energy potential of bagasse is indeed considerable and could help many developing countries to achieve a higher degree of industrialization. It would also improve their balance of payments by decreasing fossil fuel imports.

To maximize the exploitation of this renewable source of energy, it is necessary to aim at:

1. Increasing steam and power production from a given quantity of bagasse;

2. Decreasing the amount of steam needed for processing;

3. Enhancing the fuel value of bagasse through drying and densification.

The first objective can be achieved by generating steam at higher pressures and temperatures and replacing back-pressure turbo-generators with condensing-extraction units.

As an example, 25 tons of 50% moisture bagasse will produce just under 4000 kWh using a back-pressure turbo generator operating at 200 psig and 500^{0}F and exhausting at 10 psig. The same weight of bagasse will produce 9,[illegible]0 kWh if the steam pressure and temperature are raised to 600 psig and 750^{0}F, respectively, and a condensing turbo-generator is utilized instead of a back-pressure machine.

By using still higher pressures and temperatures, more energy could be generated but because of practical considerations relating to sugar factories, it is suggested that in general the pressure and temperature be limited to 700 psig and 800^{0}F, respectively, though by no means is it suggested that these limitations should apply to all cases.

The amount of process steam usage in raw cane sugar factories varies between 1400 lbs and 850 lbs per ton to cane processed. Though variations in cane quality affect steam usage to a certain extent, the most important factor lies in the design of evaporator and juice heater layouts. Process steam consumption can be decreased considerably by the judicious use of pre-evaporators and vapor bleeding techniques.

Finally, by using flue gases to dry the bagasse going to the boilers, combustion efficiency can be significantly increased resulting in more kWh being generated per unit of fiber burned. In general an increase of up to 10% can be expected in boiler efficiency.

When there is a surplus of bagasse, (there will almost always be a surplus of bagasse if the above steps are taken), the only practical method of storing this surplus is by drying and pelletizing. Though this process requires a certain amount of energy, there is nevertheless a net gain of energy through drying and pelletizing. We can show that 100 tons of 50% moisture bagasse will produce 56.8 tons of pellets at 12% moisture, and that the net gain in steam energy by burning pellets instead of 50% moisture bagasse amounts to approximately 0.2 kWh per pound of bagasse. Put in another way, about 20% less fiber (dry matter in bagasse) is required to produce the same amount of steam when burning pellets instead of bagasse at 50% moisture.

A comparative study of the economic aspects of power production from wet bagasse, dried bagasse and pelletized bagasse was done by means of computer simulations reflecting the impact of various factory improvements. The economic analysis and evaluation was done by the discounted cash flow rates of return (DCFRR) method.

These analyses point out that there is a clear benefit in drying bagasse and pelletizing the surplus, using to the maximum extent possible the heat available in flue gases. The positive influence of higher fiber content, higher grinding rates and use of pre-evaporators are also demonstrated. The simulations show that there are a great number of variables that interplay and that consequently it is not possible to make general predictions based solely, say on the grinding rate and fiber content of the cane, though these are undoubtedly important factors. Each case must be considered separately and the best solution chosen with the help of computer simulations.

Guidelines for Bagasse Energy Projects

General guidelines have been drawn for the pre-selection of applications for bagasse energy processes.

Factories that grind more than 150 tons cane per hour (TCH) are good candidates though factories grinding between 100-150 TCH can also be considered if the fibrous content of the cane is above 13.5% and particularly if the grinding season is more than six months in duration.

Boiler steam conditions and types of turbo generators are important considerations since it is generally not possible to produce significant amount of surplus power in factories that are equipped with low pressure boilers and back-pressure generators. In these cases new boilers and condensing extraction turbo-generators must be considered and in such cases a careful economic evaluation is necessary.

Other important considerations are:

-- The steam consumption of the "Boiling House" (the part of the factory processes the juice extracted from the cane). If the steam consumption is above 1000 lbs/Ton cane processed, there will be a negative impact on the amount of power that can be produced. Ideally steam consumption can be as low as 850 lbs/Ton cane in which case power production will be maximized. If above 1000 lbs/Ton cane it would be necessary to modify the steam distribution system to bring the steam consumption in line.

-- Regular cane supply and mechanical efficiency of the factory are important factors since frequent stoppages deplete bagasse stocks and may even require the use of other fuels. Factories that run on a seven-day week schedule and have a time efficiency of 90% and over are in a very favorable position.

-- The willingness of the Public Utility Company to buy energy at a reasonable rate or to buy pellets at a price near the oil equivalency rate. The best condition would be when the power company is willing to absorb all the energy produced by the sugar factory without dispatching restrictions.

-- Good cane preparation as a pre-existing condition helps the drying and pelletization processes. If the bagasse coming out of the last mill is too coarse the preparatory equipment will have to be improved before attempting pelletization.

SECTION I

Introduction

World centrifugal sugar production since 1976 seems to have reached a plateau of around 87 million tonnes per annum.

1976	86.9	million tonnes
1977	92.0	
1978	91.0	
1979	84.6	
1980	87.1	
1981	86.1	
Average	87.95	million tonnes

Of that amount about 60% is derived from sugar cane, or approximately 53 million tonnes.

Sugar cane is grown mostly in the topical and sub-tropical belts with some incursions in the warmer areas of the temperate regions, both of the Northern and Southern Hemisphere. (A list of developing countries where sugar is produced appears in Appendix I.)

In North America, and to a certain degree in Europe, corn derived sweeteners are gradually displacing sugar as industrial sweeteners. On the other hand, as the Third World Countries advance in their economic and industrial development, the need for more sugar arises and, since these countries are situated mostly in the areas where sugar cane can be grown, we can expect a gradual increase in cane sugar production from these countries.

The most serious obstacle to economic progress of undeveloped countries is the high cost of energy, since very few have sufficient energy resources. The great attraction of the cane sugar industry for these countries has always been and still is the fact that cane sugar factories are energy self-sufficient and therefore can be operated in areas without electricity, coal, gas or oil resources. By developing further this biomass resource these countries could decrease the import of fossil fuels, thus improving their balance of payments. The authors of this study have personal experience of villages where the only source of electric power was the sugar factory, and this only three decades ago!

Evolution of the Cane Sugar Factory as a Source of Electrical Energy

It is important that World Bank personnel who will be evaluating the merits of sugar factory energy projects understand the following dealing with the evolution of the cane sugar factory as a source of electrical power since they will find that around the world today there are factories at all stages of this evolution.

The processing of sugar cane into sugar requires two distinct processes: the extraction of juice from cane stalks and the "boiling" of the juice into sugar crystals. The extraction process requires mechanical power whilst the "boiling" process requires heat energy in the form of low pressure steam, since it is mainly an evaporation process.

Steam is obtained by burning the residue, called bagasse, from the extraction process. It fortunately so happens that in most cases enough bagasse is produced to provide the steam necessary for running the factory. In the old days, however, with low grinding rates, inefficient milling equipment and steam engines, bagasse had to be supplemented with wood and/or coal where it was readily and cheaply available. The old sugar factories were entirely steam driven and the senior author remembers working in a relatively large factory where the only electric generator had a capacity of only 50 KW, just enough to provide lighting for the factory and camp houses. There are probably many such factories still in existence today in remote parts of the world. At the other end of the spectrum are large, efficient factories capable of producing several times more power than their own requirements, as exemplified by a number of Hawaiian sugar factories. In between these extremes will be found the vast majority of the world sugar factories.

Whereas in factories of the first type every piece of equipment is steam driven, in a typical modern factory all drives are electrical except for the large prime movers driving the milling equipment and some of the larger boiler auxiliaries, which are usually turbine driven.

This means that these factories are equipped with sizeable turbo-generators, which in most cases are of the back pressure type. Condensing units and condensing units with pass-outs are common in Hawaii but rather the exception elsewhere. This type of machine, however, will be found in increasing numbers as the sugar factories of the world gear up to produce electrical energy in greater quantity.

The philosophy of the early designers was simply to produce enough steam at the lowest possible cost to provide motive power and enough heat to process the juice. These goals were achieved by using steam boilers operating at 100 to 150 psig. Steam was saturated, only rarely superheated. The boilers in most instances were of the fire-tube type suspended over "Dutch ovens," though in certain areas water-tube boilers were in use. The boilers were of small capacity; hence typically the steam generating station consisted of a battery of boilers, a situation which still exists today in many sugar factories. At these low steam pressures, steam turbines are very inefficient and most steam drives were of the reciprocating type.

Gradually as factories became larger and the need to modernize and improve efficiency arose, there was a shift towards higher pressure boilers of larger capacities and most of the fire-tube boilers were replaced by water-tube boilers at 175 to 250 psig pressures. With higher steam pressures and temperatures turbo-generators began to appear more frequently, most of them of the back pressure type. With higher live steam pressures, higher back pressure became possible, which in turn made it possible to obtain more evaporation per pound of process steam in the boiling house by using vapor cells and vapor bleeding, as will be explained further down.

In general, however, the sugar industry throughout the world was slow in adopting boiler pressures above 250 psig because of the ingrained philosophy that no more energy was necessary than what was required to satisfy internal needs, including irrigation.

The oil embargo of 1973 that triggered a worldwide energy crisis made many cane sugar producers realize that they had an under-used and under-valued resource in bagasse, particularly in those instances where their sugar factories had enough capacity to produce electrical power for export. Suddenly the philosophy changed from one of mere self-sufficiency to one of exploiting to the maximum a natural resource that was readily available. Unfortunately the cost of boilers and electrical generators is high and, since their useful lives range between 20 and 30 years, one can easily understand why the shift towards larger, more efficient boilers and turbo-generators at higher pressures has not taken place rapidly. Also in many instances sugar factories are not allowed by law to supply power in the local utility grid or the price they are offered for their energy is unattractive. In many instances there is not even a utility grid into which power could be supplied.

We must recognize, however, that attitudes of government, public utilities and producers are changing and today in many underdeveloped countries there seems to be a determined effort on the part of all parties to work together towards the common goal of energy self-sufficiency. Because of these changes in attitude and because in the last few years a significant number of large new factories have been erected in various countries, we are beginning to find steam generators with working pressures ranging between 450 and 1200 psig supplying steam to condensing-extraction turbo-generators ranging up to 20 MW, though the vast majority of boilers in sugar factories are still operating at pressures ranging between 175 and 250 psig. However, even at this lower range it is possible for a sugar factory of reasonable size and efficiency to generate a surplus of electrical energy, though not as much as would be possible at higher steam pressures and temperatures.

Another factor which is closely related to the production of surplus power in a factory is process steam consumption. For a given set of operating conditions, the less steam required to process the cane the more will be available for the production of power, assuming that the electrical generator has the extra capacity, as is sometimes the case.

If the extra generator capacity is not there, reduction in steam consumption will result in a surplus of bagasse. Until recently this was not really useful since it is difficult, expensive and dangerous to store bagasse in loose form. (Stored bagasse is prone to spontaneous combustion due to temperature rise within the mass caused by the fermentation of residual sugar.) Fortunately a recent technological advance makes it possible economically to dry and densify bagasse to a stable and dense form, making storage of large quantities feasible. Stored densified bagasse can be used to produce electrical energy beyond the harvest season (which in most countries lasts between 5 and 7 months per year), provided the factory is equipped with a condensing or condensing-extracting turbo-generator. Densified bagasse has the added advantage of causing the boiler efficiency to increase significantly, as will be discussed in another part of this study.

Global Overview of Energy Potential in the World Cane Sugar Industry

Yearly production of cane sugar = 53,000,000 metric tons

= 58,300,000 short tons.

Assuming average ton cane / ton sugar ratio of 8.5:

Weight of cane produced = 495,000,000 short tons.

Other assumptions (based on average data):

Average bagasse contains 50% moisture.

Average HCV of bagasse = 4,200 BTU/lb.

Average bagasse % cane = 25

Average fibre % cane = 12.5

Average steam production per lb. of bagasse = 2.2 lb.

Average boiler efficiency = 60%

Average steam requirement to manufacture raw sugar

= 0.5 tons/ton cane processed

$\therefore$ Steam requirement of whole industry = 0.5 x 495,550,000

= 247,775,000 tons

Bagasse needed to produce this amount of steam = $\frac{247,775,000}{2.2}$

= 112,625,000

= 1.12625×10^8 tons

But bagasse produced = 495,550,000 x 0.25

= 123,887,500

= 1.238875×10^8 tons

$\therefore$ Theoretical bagasse surplus = $1.2389 \times 10^8 - 1.1263 \times 10^8$

= 0.1126×10^8

= 1.126×10^7 tons

BTU contained in surplus bagasse = $1.126 \times 10^7 \times 4,200 \times 2,000$

= 9.4584×10^{13} BTU

By burning this surplus bagasse in boilers having 60% thermal efficiency, heat transferred to steam = $0.60 \times 9.4584 \times 10^{13}$ BTU

Assuming an efficiency of 27% in converting heat energy into electrical energy, energy produced from surplus bagasse = $5.6750 \times 10^{13} \times 0.27$

= 1.5322×10^{13} BTU

To convert BTU into KWH: $\times 2.93 \times 10^{-4}$ = 4.4893×10^9

= 4.49 billion KWH

This global overview points to a surplus of bagasse in the world today on the order of 1.126×10^7 tons assuming average values for factory efficiencies, fibre percent cane, etc. and a possible surplus of 4.5 billion KWH of electrical energy. Hence the potential for producing electrical power for other uses is very real. These figures do not take into consideration the improvements that can be made to the average cane sugar factory to reduce its steam consumption for processing needs, thus making more steam and more bagasse fuel available for power production, nor do they take into consideration the energy gains that would result from the drying of bagasse by making use of the boiler flue gases.

In his recent book By-Products of the Cane Sugar Industry, J. M. Paturau points out that on a world basis it would theoretically be possible to produce over 50 KWH of surplus electrical energy for every metric ton of cane processed, which would result in 22 billion KWH of extra energy per year whereas, with the present state of the industry, 4.5 billion KWH could probably be produced if all the surplus bagasse were to be utilized. We know that there already is a significant amount of surplus energy being produced by the cane sugar industry in several countries as exemplified by Hawaii and Mauritius, but we have no means of knowing how much surplus power is produced worldwide by sugar factories as these data are not usually readily available. It is clear, however, that by following the examples set by Hawaii and

Mauritius a very significant contribution to the energy shortage, especially in the underdeveloped countries, could and should be made by the cane sugar industry.

SECTION II

Exploitation of World Sugar Industry Energy Potential

In examining how the sugar industry could exploit more fully the vast potential of bagasse energy we must consider three avenues:

1) How to increase steam and power generation;

2) How to decrease the amount of steam needed to operate a sugar factory, thus making more steam and/or bagasse available for power generation;

3) How better to utilize bagasse by drying and densification.

Increasing Steam and Electrical Power Generation

The steam generators used in the sugar industry cover a broad spectrum of types, sizes, operating pressures and efficiencies, from the small fire-tube boiler at 100 psig and less than 50% efficiency burning bagasse on step grates using natural draft, to large boilers equipped with spreader-stokers operating at pressures as high as 1,200 to 1,500 psig at efficiencies of 67% and above. But as it was pointed out earlier in this report the vast majority of sugar factories still have unsophisticated water-tube boilers operating in the pressure range of 150 to 250 psig and on the average producing 2.2 pounds of steam for each pound of bagasse burned. For example, such a boiler producing steam at 200 psig and 500^{O}F temperature from boiler feed water at a temperature of 190^{O}F, burning bagasse at 50% moisture, would have an efficiency of 58.1% as the following calculation shows:

BTU in 1 ton of bagasse = 2,000 x 4,200

$= 8.4 \times 10^{6}$

Let efficiency be X%.

.˙. Heat transferred to steam by burning 1 ton of bagasse

$= \frac{X}{100} \times 8.4 \times 10^{6}$ BTU

Heat required to produce 1 lb. of steam at 200 psig and 500^{O}F from boiler feed water at 190^{O}F = 1,267.4 - 158

= 1,109.4 BTU/lb.

Steam produced per ton of bagasse = 2.2 x 2,000 lb.

$= 4.4 \times 10^{3}$ lb.

.˙. Heat required to produce 4.4×10^{3} lb. of steam from water at 190^{O}F $= 4.4 \times 10^{3} \times 1{,}109.4$

$= 4.8814 \times 10^{6}$ BTU

Equating we have: $\frac{X}{100} \times 8.4 \times 10^{6} = 4.8814 \times 10^{6}$

$$X = \frac{4.8814 \times 100}{8.4}$$

= 58.11%

However, modern bagasse boilers in the range of 100,000 to 300,000 lb./hr. can achieve efficiencies ranging between 60% and 65% when burning bagasse at 50% moisture. With 35% moisture, boiler efficiency increases to 70-75%; whilst burning bagasse pellets of 10% moisture, efficiencies would reach 78% to 82%.

Steam turbines become more efficient as steam pressures and temperatures increase. For example, a typical multistage turbine operating with steam at 200 psig and 500^{o}F, exhausting at 10 psig, would require 27.5 lb. of steam to produce 1 KWH and 12.0 lb. of steam if exhausting to a condenser at 2" Hg abs. On the other hand, at 600 psig and steam temperature 750^{o}F, the steam rate would drop to 16.6 lb./KWH when exhausting at 10 psig and 9.8 lb./KWH when exhausting to a condenser at 2" Hg abs. In these cases a thermodynamic efficiency of 0.70 has been assumed for the back pressure units and 0.72 for the condensing unit.

To illustrate the positive impact that an increase in steam pressure and temperature has on a steam turbine, we have worked out the following comparison. The boiler efficiencies in both cases have been assumed to be the same, though in fact if a new boiler were to replace an old boiler we would expect an increase in efficiency from 58% as quoted in this example to 65%.

	Case I	Case II
Boiler pressure	200 psig	600 psig
Steam temperature	500°F	750°F
Bagasse burned per hour	25 tons	25 tons
Boiler efficiency	58%	58%
HCV of bagasse	4,200 BTU/lb.	4,200 BTU/lb.
Temperature of feed water	190°F	190°F
Lb. of steam produced/hour	109,790	99,762
Theoretical steam rate of turbine exhausting at 10 psig	19.27 lb./KWH	11.64 lb./KWH
Assumed thermodynamic efficiency	0.70	0.72
Expected steam rate	27.52 lb./KWH	16.16 lb./KWH
KW output	3,989 KW	6,173 KW

If instead of a back pressure unit a condensing turbo-generator were used, the temperature of the condensate would drop to 101°F but much more power would be generated:

	Case I	Case II
Lb. of steam produced/hour	101,644	92,991
Theoretical steam rate of turbine exhausting at 2" Hg	9.32 lb./KWH	7.09 lb./KWH
Assumed thermodynamic efficiency	0.72	0.74
Expected steam rate	12.944 lb./KWH	9.50 lb./KWH
KW output	7,853	9,707

If the 200 psig condensing turbo-generator already exists and it is desired to keep it, approximately the same result can be obtained if a new boiler is installed at 600 psig and 750°F and a back pressure turbo-generator is installed admitting steam at the boiler pressure and exhausting at 200 psig. This "topping" turbo-generator will produce 1,915 KW. The exhaust from this turbine will be at a higher temperature than the theoretical temperature because of the low thermodynamic efficiencies of such topping turbines. In our example the temperature of the steam will be 550°F. By desuperheating the exhaust steam from 550°F to 500°F, additional steam will be generated. This steam can now be admitted to the existing 200 psig turbo-generator. In our example 2,413 lb. of steam per hour will be gained, and the low pressure turbo-generator will now yield 7,896 KW which added to the 1,915 KW from the topping unit will total 9,811 KW. (The calculations for this example will be found in Appendices II and III.)

It is clear therefore that the combination of high steam pressure and temperature is essential to recover the maximum amount of electrical energy from a given quantity of bagasse. The cost of boilers increases as we go from one range of working pressures to the next. Above 600-700 psig the cost increases rapidly and feed water treatment becomes increasingly critical. For these reasons we recommend that, unless warranted by special circumstances, sugar factory boilers should be kept within the 500-600 psig range with total steam temperatures below 775°F to avoid the need for special alloy steel tubes and piping.

The above is a hypothetical example, in which all the steam generated by burning 25 tons of bagasse per hour is used to provide electrical power. In practice, however, a sugar factory requires both electrical power for its own needs and steam to operate milling, power generation and processing equipment.

It can be assumed that 25 tons of bagasse is the result of crushing 100 tons of cane per hour. If the factory is electrically driven except for the larger prime movers driving milling equipment and power generation auxiliaries, we can guess that it will consume around 800 KW and require about 100,000 lb./hr. of steam for processing the juice extracted from the mill into raw sugar.

To continue our example, 110,000 lb. of steam is generated per hour at 200 psig and 500^{o}F. It will require about 22,000 lb. to produce 800 KW at the steam rate of 27.5 lb./KWH. Five 500 HP single stage turbines (which is the common practice in the industry) driving the cane knives and four mills running at 400 HP actual load would require approximately 2,000 x 35 = 70,000 lb./hr. of steam, leaving some 18,000 lb./hr. for the steam driven boiler auxiliaries. Such a factory could barely take care of its electrical, mechanical and process needs but, by raising the steam pressure and temperature to 600 psig and 750^{o}F respectively and adding a topping turbine, it would now be capable of generating some 1,900 KW of surplus power as indicated above. Also, due to the desuperheating of the steam exhausting from the topping turbine, there would be approximately 2,400 lb. more steam available. This factory could now more easily meet the demand of process steam on top of producing a significant amount of extra power.

It should be borne in mind when reading the above example that the purpose here is only to give the reader a rough picture based on some common assumptions and practical experience.

Decreasing Process Steam Requirements

The basic sugar factory uses process steam at 10 to 15 psig to heat the mixed juice coming from the mill to boiling point prior to the clarification process. Evaporation is usually carried out in a triple or quadruple effect evaporator with bodies of equal size, using 10-15 psig process steam for the first effect. Final concentration of the syrup is done in single effect vacuum pans also using process steam. There are also other minor needs for steam.

The process steam is provided by the exhaust from the steam driven prime movers. In designing a factory one usually attempts to balance out the need for live steam with the need for process steam. In practice this is not easy to achieve because the vacuum pan boiling process is an intermittent operation and therefore gives rise to a fluctuating demand for steam. If the designer has cut the balance too fine there will be times when the exhaust steam is in excess, in which case it is simply blown "over the roof," resulting in a loss of energy. On the other hand if there is not enough exhaust to meet peak demands some live steam has to be reduced to make up the amount of process steam required. Both instances are wasteful and should be avoided as much as possible.

The classical methods to reduce the demand of process steam are the following:

1) Use of vapor bleeding from the first and second bodies of the evaporator to heat the juice instead of using exhaust steam.
2) Adding a pre-evaporator or vapor cell, as it is sometimes called, ahead of the evaporator. If properly sized, the pre-evaporator should be able to absorb <u>all</u> the exhaust steam of the factory and provide vapor (steam) to the triple or quadruple effect evaporator, juice heater (for final juice heating), and vacuum pans. The net effect in this case is that the juice is evaporated in quadruple or quintuple effect and the pan boiling takes place in double effect. The savings in steam consumption are very significant. As a rule of thumb, bleeding vapor from the first body of an evaporator saves 1/4 of the steam that would have been required if exhaust had been

used instead; from the second body 1/2 of the steam is saved; from the third body 3/4 is saved, etc.

3) There are other more sophisticated methods such as the use of thermo-compression. These methods are rarely used in the sugar industry and will not be discussed here.

We have calculated that without any bleeding of vapor the boiling house of a basic factory would require 1,026 lb. steam / ton cane;

With bleeding of first vapor only 896 lb. steam / ton cane;

With bleeding of first and second vapor 846 lb. steam / ton cane.

In the case of a sugar factory of 150 tons cane / hour capacity this would translate into a saving of 27,000 pounds per hour. (All calculations that have led to these numbers will be found in Appendix IV.)

The importance of steam economy in the boiling house of a sugar factory cannot be overemphasized if power production is a consideration. It must be realized, however, that it would be pointless and as a matter of fact counterproductive to cut down the steam consumption of the boiling house below the level of exhaust steam generated by the prime movers. This is the reason why a pass-out turbo-generator becomes a key factor in reaching a good balance between exhaust steam and process steam requirements. With a pass-out turbo-generator, the moment exhaust steam tends to become overabundant it is diverted to the condensing side of the turbine where it produces electrical energy instead of being wasted by blowing off over the roof.

Drying and Pelletizing of Bagasse

General Considerations

The drying and pelletizing of bagasse as developed by TheoDavies Hamakua Sugar is now a patented and well proven technology. This technology uses the flue gases from the boiler as the drying medium. This by itself is not innovative as many attempts have been made to dry bagasse with flue gases and in fact there are several dryers on the market today which have met with varying degrees of success. As far as the authors of this report know, however, the only dryers which have achieved consistent results over a lengthy period of time are those installed at the Haina factory of TheoDavies Hamakua Sugar. These dryers were supplied by Rader Western, Inc. and the control system which is a key ingredient to their success has been developed by TheoDavies Hamakua Sugar.

It must be understood that a bagasse dryer using flue gases from bagasse boilers presents a special challenge because of the low level of heat energy contained in the flue gases, their high moisture content, large volume, and variability in terms of volume, temperature and moisture contents. Adding to the challenge are the variations in moisture, ash content and rate of production of bagasse experienced by any normal sugar factory. To complicate matters further, the exit temperature of the gases must not be allowed to fall below the dew point temperature since condensation of water vapor contained in flue gases would cause severe corrosion problems in the dryer, cyclones, duct work and I.D. fan. For these reasons we do not believe that a dryer can operate successfully on manual control, and the success of the drying operation and of the pelletizing operation which may or may not follow will depend to a large extent on the degree of automation developed to react instantly to the interplay of the numerous variables. In the case of the Haina factory, the controls have been computerized using an IBM Series/I computer, but we believe that a simpler system could be developed using microprocessors driven by a smaller and less expensive computer than the IBM Series/I.

It can be inferred from what precedes that it is impossible to offer drying design packages and that each drying system would have to be customized taking into consideration the numerous variables associated with each cane sugar factory.

Bagasse drying does not mean that pelletization or cubing must necessarily follow, but if densification of one form or another is desired, then drying is not a choice but a necessity. It will be found that in practice it is impossible completely to dry all the bagasse as there is generally not enough heat energy in the flue gases to do so; hence, choices have to be made. In most cases all the bagasse coming from the mills can probably be dried to a moisture content of 30% to 40% and in some instances to a much lower level, or part of the bagasse can be dried to a lower moisture content whilst the remainder will be dried to a higher moisture. The choice could also be made to dry only the surplus bagasse to be pelletized. There are now dryers coming out on the market which can effect differential drying of coarse and fine particles in a single unit. With the proper design the moisture content of the fine particles may be low enough to enable pelletizing after separation from the coarse particles, which are sent to the boiler.

As the bagasse fed to the boiler is dried, the characteristics of combustion change, resulting in higher flame temperatures, lower volume of flue gases, lesser amount of water vapor in flue gases, and lower velocities of gases through the boiler for a given amount of fibre burned. The net effect of these changes is a higher boiler efficiency, meaning that with the drier bagasse less fibre is required to produce the same amount of steam. By the same token there are now less flue gases, but on the other hand a change is taking place in the composition of the flue gases, which now contain a lesser percentage of water vapor which in turn will cause a lowering of the dew point temperature. Lowering of the dew point means that more heat can safely be extracted from the flue gases, thus enhancing the drying process.

When calculating a drying system all these factors have to be taken into account and the final design parameters can only be reached after numerous trial calculations necessary to reach the point of stable condition between the degree of drying and the composition and characteristics of the flue gases because of the number of variables involved. It would be a lengthy process to try to do these calculations by hand and consequently costly. Mr. Norland Suzor of TheoDavies Hamakua Sugar has fortunately developed a computer model which allows these calculations to be made very rapidly.

Pelletizing

In order to pelletize bagasse it is essential to do two things:

1) Reduce the particle size of the bagasse coming from the mills to a size that is acceptable by the pellet mill;
2) Reduce the moisture content of the fine particles to below 12%.

Particle Size

Bagasse coming out of a milling plant will exhibit great variation in particle size depending on a number of factors:

1) Percentage of fibre in the cane. The higher the fibre content for a given weight of cane, the more work has to be done to reduce this cane (and hence the bagasse) into fine particles, and obviously the more difficult this process is.
2) Type of preparation devices. The flow of cane entering the factory is normally subjected to the action of revolving

cane knives and/or a shredder (or fiberizer) the object of which is to reduce the cane into small pieces, at the same time attempting to separate the fibres into a fluffy mass. The more thorough this preparation, the finer will be the bagasse emerging from the milling tandem.

3) Type of milling equipment. The fibre loading of the mills, the number of units in the milling train, the type of mill roller groovings and the mechanical condition of the milling plant will determine to what degree more particle reduction takes place during the passage of the prepared cane through the mills.

Poor preparation, high fibre loading, poor milling plant design and condition may result in coarse bagasse having an insufficient portion of fine particles to enable successful pelletizing. A solution to this problem, albeit a costly one, would be to incorporate a bagasse shredder into the pellet plant. This practice, however, is most undesirable not only because of the extra cost for the equipment but also because it requires a considerable amount of energy to shred bagasse, energy that reduces the net energy gain of the drying and pelletizing process.

It is recommended therefore that, in the case where there is not enough fine bagasse for pelletizing, attention be given to the preparation of the cane prior to milling. In other words, if more shredding is necessary it should be done on the whole cane and not on the bagasse, since a better preparation prior to milling will increase significantly both the grinding capacity and the extraction efficiency of a milling plant. The additional sucrose extraction in most cases will more than justify the cost of the additional preparatory device, thus providing an additional benefit to the drying and pelletizing process at no cost.

The series of scenarios developed for the economic study of drying and pelletizing will indicate that at most 10% to 20% of the bagasse of a sugar factory can ever be converted into pellets economically, and again it will be restated that the only justification for pelletizing is the necessity to store bagasse fuel in a dense, stable form which is easily handled and can be utilized at any time, as during the off-season when the sugar mill is not operating.

Net Gain in Energy Through Drying and Pelletizing

Raw material: 100 tons bagasse at 50% moisture

Finished material: Pellets at 12% moisture

Weight relationship between bagasse and pellets:

100 tons bagasse contain 50 tons dry solids + 50 tons water.

100 tons pellets contain 88 tons dry solids + 12 tons water.

$\therefore$ 100 tons bagasse yield $\frac{100 \times 50}{88}$ = 56.81 tons pellets.

Water to be evaporated from 100 tons of bagasse to produce 56.81 tons pellets:

100 tons bagasse $\longrightarrow$ 56.81 tons pellets

56.81 tons pellets contain 6.817 tons water.

$\therefore$ Water evaporated = 50 - 6.817

= 43.18 tons.

Energy required to produce 1 lb. of pellets:

Power requirement for a 250 tons pellets/day plant:

Installed KW	= 535.2	(assuming boiler I.D. fan is sufficient)
I.D. fan	= 300	(if extra fan is required)
Total	835.2	

Case I: No extra I.D. fan necessary

Assume plant operates 22 hr./day

$\therefore$ Output of pellets $= \frac{250}{22}$

= 11.36 tons/hr.

$= 2 \times 10^3 \times 11.36$ lb./hr.

Assuming that load factor is 80%

Energy required per hour = 535.2 x .8

= 428.16 KWH

$\therefore$ Energy per 1 lb. pellets $= \frac{4.2816 \times 10^2}{2 \times 11.36 \times 10^3}$

$= 0.1884 \times 10^{-1}$

= 0.01884 KWH/lb. of pellets

Case II: Extra I.D. fan is required

Pellet output is the same at $2 \times 11.36 \times 10^3$ lb./hr.

Total KW installed is now 835.2 KW.

Assuming load factor of 80%

Energy required per hour = 668.16 KWH

$\therefore$ Energy required to produce 1 lb. of pellets

$$= \frac{6.6816 \times 10^2}{2 \times 11.36 \times 10^3}$$

= 0.02940 KWH/lb. of pellets

The net energy gain is obtained from the increased boiler efficiency.

Efficiency with bagasse at 50% moisture = 60%

Efficiency with pellets at 12% moisture = 78%

Calorific value of 1 lb. of bagasse = 4,200 BTU/lb.

Calorific value of 0.5681 lb. pellets = 4,200 BTU/lb.

In the case where all the bagasse surplus is converted to pellets at 12% moisture, the net gain in energy from this surplus will be as follows.

At 60% boiler efficiency, heat transferred to steam

= 0.6 x 4,200 BTU.

At 78% boiler efficiency, heat transferred to steam

= 0.78 x 4,200 BTU.

$\therefore$ Gain in energy transferred = $(3.28 - 2.52) \times 10^3$

= 0.76×10^3 BTU/lb. of bagasse

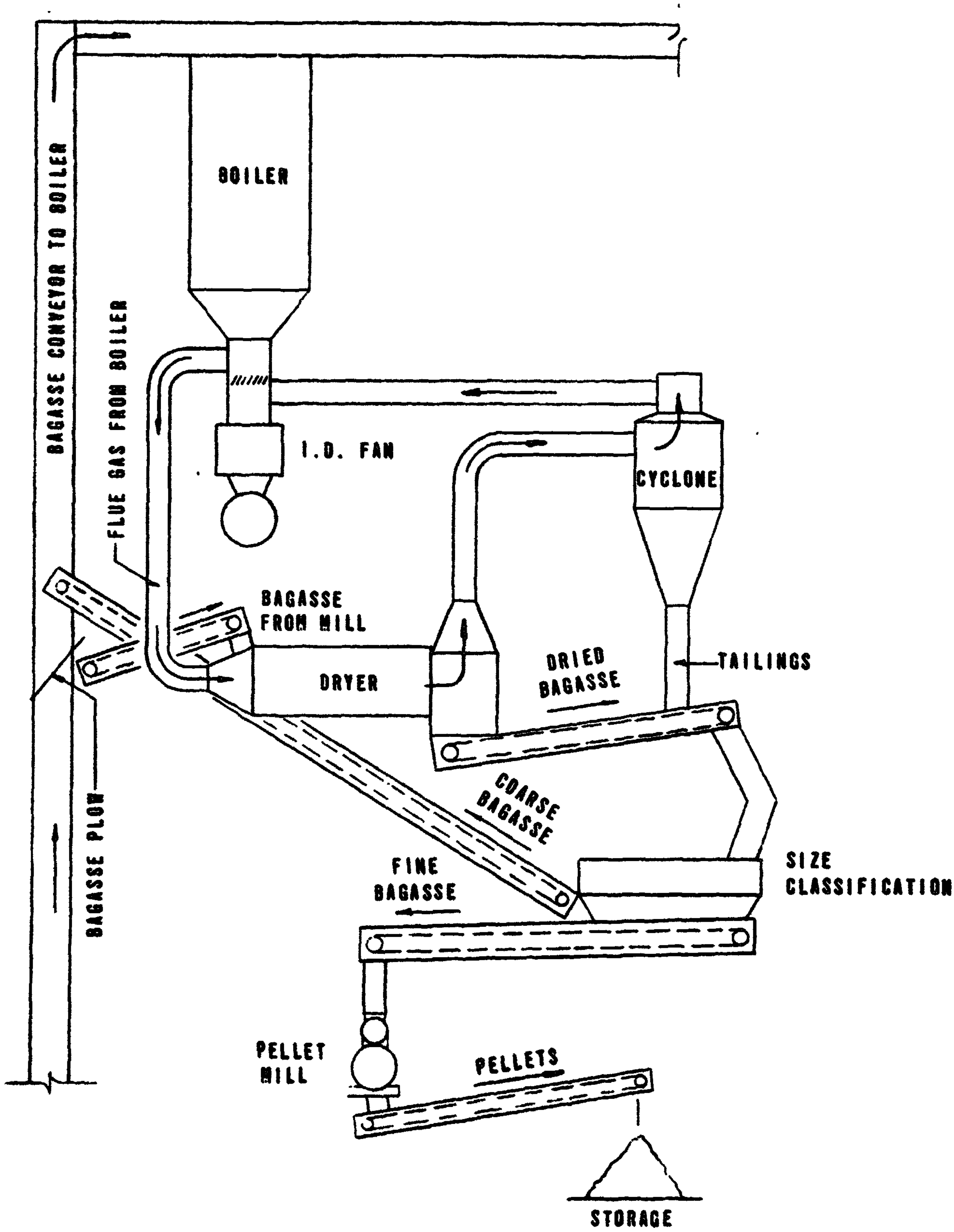
= 0.22268 KWH/lb. of bagasse

Net energy gain = energy gained in steam during combustion of pellets less energy required to make pellets

= 0.22268 - 0.01884 = 0.20384 Case I

= 0.22268 - 0.02940 = 0.19238 Case II

Put in another way, about 20% less fibre is required to produce the same amount of steam when burning pellets instead of bagasse at 50% moisture.



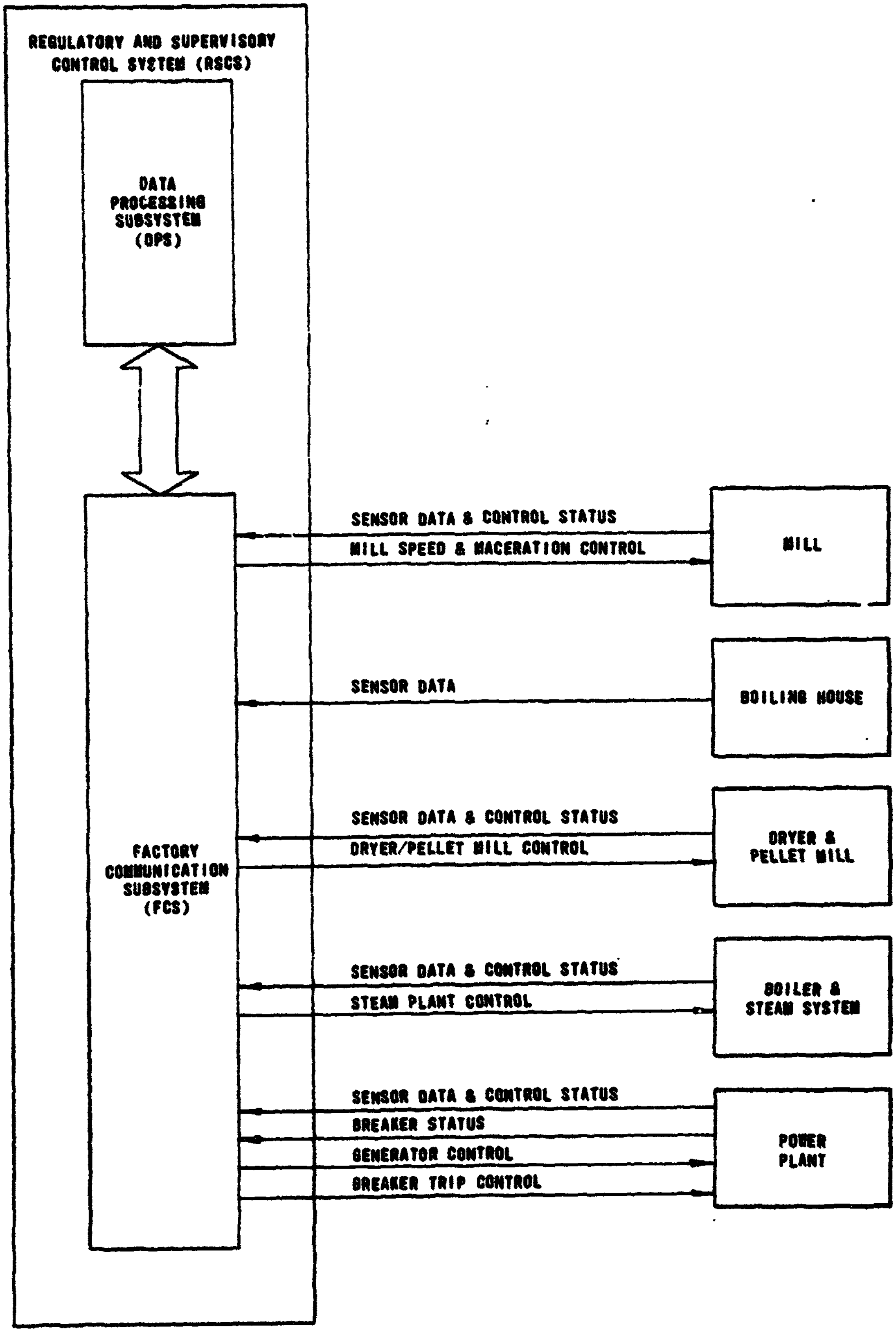
SCHEMATIC OF DRYING AND PELLETIZING

Drying and Pelletization Control System

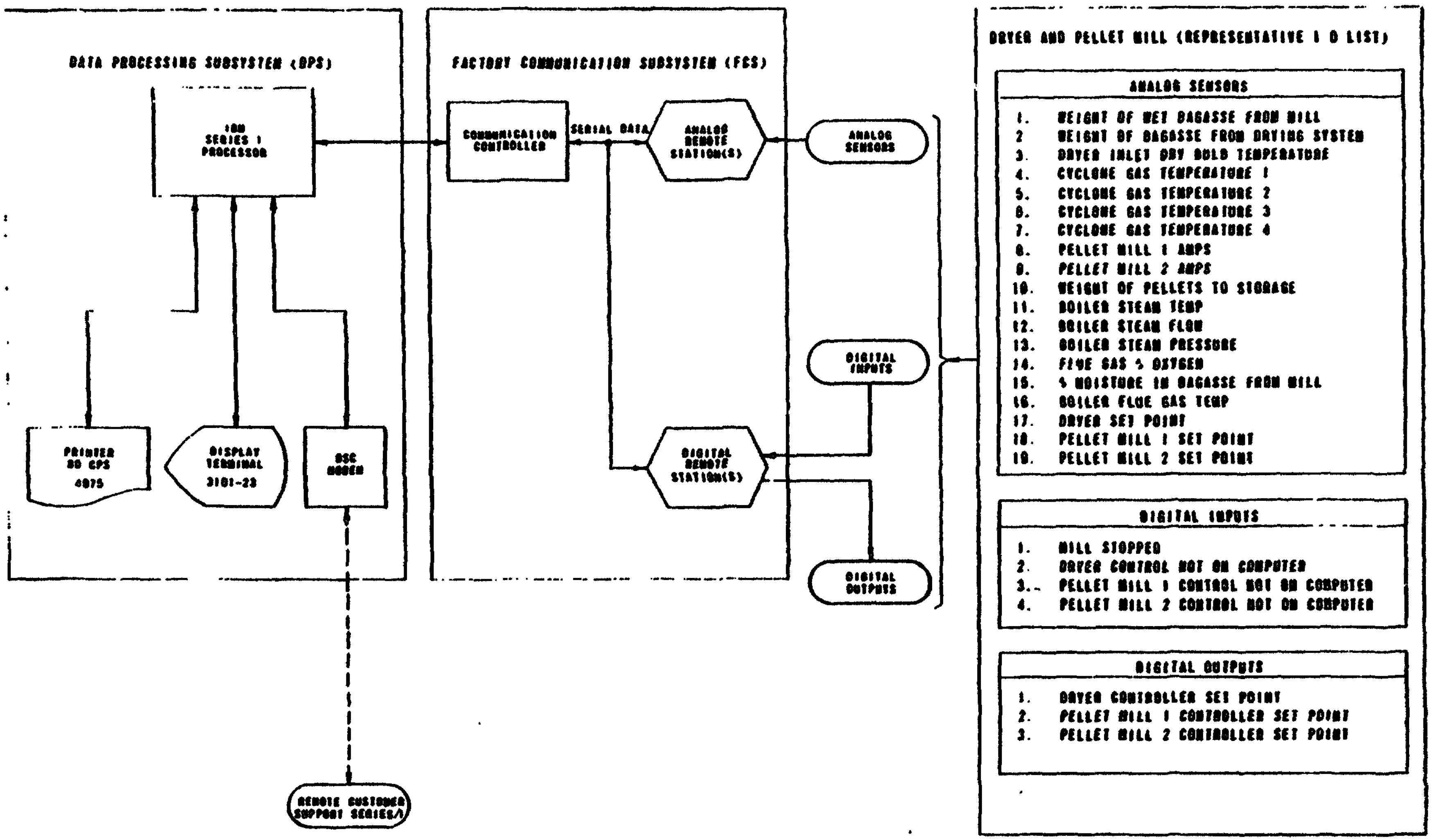
The system is defined as a Closed Loop Regulatory and Supervisory Control System (RSCS).

The RSCS Configuration diagram on the following page illustrates the essential elements of the control system as installed at the Haina factory of TheoDavies Hamakua Sugar. This system is now well proven and we recommend that it be adopted as the standard for other installations. Though the system depicted uses an IBM Series/I computer, other types of processors could be adapted to provide the same functions.

The RSCS Factory Relationship diagram which follows is presented for information only. This is the complete factory control system which operates at the Haina and Ookala factories of TheoDavies Hamakua Sugar, and the diagram shows the integration of the drying and pelletizing control as part of the overall control system of the factory.



RSCS FACTORY RELATIONSHIP



RSCS CONFIGURATION

SECTION III

Economic Aspects of Power Production From Wet Bagasse, Dried Bagasse and Pelletized Bagasse

We have already seen that in the processing of sugar cane into sugar a large quantity of low pressure steam (process steam) is required to evaporate water contained in the juice. Since steam is generally produced at a much higher pressure than process steam pressure, the difference in pressure is used to drive the equipment necessary for extracting the juice, boiler auxiliaries and electric power generator. The cane sugar factory is therefore a typical example of co-generation where power generated, mechanical or electrical, can be considered a by-product of the sugar manufacturing process.

The fuel used, bagasse, is itself a by-product of the extraction process and in order to maximize the use of this biomass it is necessary to decrease to a minimum the amount of process steam needed to produce a given weight of sugar and to increase to the maximum the conversion of bagasse fuel energy into steam energy. These objectives, which can be accomplished in a variety of ways, must nevertheless remain within acceptable economic limits.

As it was pointed out earlier in the study, the typical sugar factory was not designed to maximize the energy potential of the bagasse it produces, but significant improvements can be made towards that end, as we will demonstrate. Starting with a "basic" factory we will develop a series of scenarios that will show the impact, on the production of electrical power and pelletized fuel and on the ROI's, of the following factory improvements:

Increasing boiler steam pressure and temperature;

Decreasing consumption of process steam;

Adding a topping turbo-generator;

Adding a condensing turbo-generator;

Drying the bagasse using flue gases;

Pelletizing the surplus bagasse.

The basic sugar factory selected has the following parameters:

Grinding rate	150 tons cane/hour	
Fibre % bagasse	12.5	
Moisture % bagasse	50.0	
Bagasse % cane	25.0	
Steam generation:	Boiler working pressure	200 psig
	Steam temperature	500°F
	Flue gases temperature	520°F
	Exhaust pressure	15 psig

Boiling house steam requirement = 1,026 lb., to which must be added 60 lb. for losses and miscellaneous use, thus giving 1,086 lb. steam per ton cane. (See Appendix IV.)

N.B. The boiling house has a quadruple effect evaporator with vessels of equal heating surface. All juice heating and vacuum pan boiling are done with exhaust steam at 15 psig.

From what has been discussed so far it is apparent that the economical generation of surplus power by a sugar factory is dependent on a host of factors that are of both an internal and external nature and that the economics of power production are inextricably linked to these factors. The variability of so many factors over a wide range of conditions makes it impossible for this report to conclude categorically that surplus power can and should be produced at each sugar factory. What it will do, however, is to point out what steps can be taken in order to develop a power generation program and the probable economic consequences.

To demonstrate the effect of each of these steps we start with a "basic" factory representing average conditions, as described on the preceding page. The ROI results obtained from this exercise can be applied only to this particular case and cannot be generalized. It must also be borne in mind that the somewhat massive capital expenditures undertaken to turn this medium size, basically inefficient factory into an efficient unit capable of producing surplus power are all charged against the cost of the power program, discounting the benefits that would accrue to the sugar operation itself. The sole exception is in the case of the boiler, in which <u>only</u> the differential cost between a low pressure and a high pressure boiler is taken into account, the rationale being that unless a factory needs a replacement boiler the cost of a high pressure boiler could probably not be justified if it is to be installed solely for the purpose of increasing power production and be of no benefit to the sugar operation.

Our approach is therefore somewhat biased and unrealistic and tends to decrease the attractiveness of the production of surplus power. In real life it will be found that, in those cases where a large, efficient factory especially equipped with a condensing/extracting turbo-generator and with a low steam consumption already exists, the economics of power production would be extremely favorable, whereas in the case typified by our version of the "basic" factory it will be on the marginal side. Hence in practice the viability of each project will have to be determined by a study relating to the specific conditions prevailing at each factory.

Computer Simulations Reflecting Impact of Various Factory Improvements

In the example chosen, to illustrate the impact of each of the steps mentioned above on both the technological and economic planes we have used our computer model to develop a series of simulations consisting of six scenarios, each scenario showing three conditions:

Condition 1 Factory operation without bagasse drying and pelletizing;

Condition 2 Factory operation with drying only;

Condition 3 Factory operation with drying and pelletizing.

In the scenarios the additional steps taken to improve the basic factory other than drying and pelletizing are as follows:

Scenarios 1 & 2	Operation of basic factory without any alteration.
Scenario 3	Install a 3,000 KW turbo-generator to convert the excess process steam, generated in Scenario 2, into additional power.
Scenario 4	Same as Scenario 3 but with the addition of an 8,000 sq.ft. pre-evaporator so as to decrease the process steam requirement from 1,026 to 896 lb./ton cane.
Scenario 5	The basic factory of Scenario 2 is modified by the addition of a new boiler to upgrade the steam from 200 psig/500^{o}F to 600 psig/750^{o}F. Further, a topping turbo-generator is installed to reduce the high pressure and temperature steam to the previous operating level of 200 psig/500^{o}F necessary for the existing prime movers while generating additional power.
Scenario 6	As Scenario 5 but with the addition of a 1,000 KW condensing turbo-generator and an 8,000 sq.ft. pre-evaporator. Note: In this case one could combine the topping turbo-generator and the 1,000 KW condensing unit into a single 4,500 KW condensing/extracting turbo-generator.

Condition 1 of Scenario 1 determines whether there is a surplus of bagasse under the normal operating conditions of the basic factory. Since the accumulation of surplus wet bagasse is counter-productive, in all other scenarios Condition 1 mandates the burning of all the bagasse and as a result the boiler output under this condition is established. Maintaining this same boiler output expressed in total BTU in steam per hour for Conditions 2 and 3 of all other scenarios makes it possible to demonstrate the net effect of drying and pelletizing, which results in a surplus of dried bagasse or pellets. These surpluses will also remain constant throughout all scenarios (except number 5).

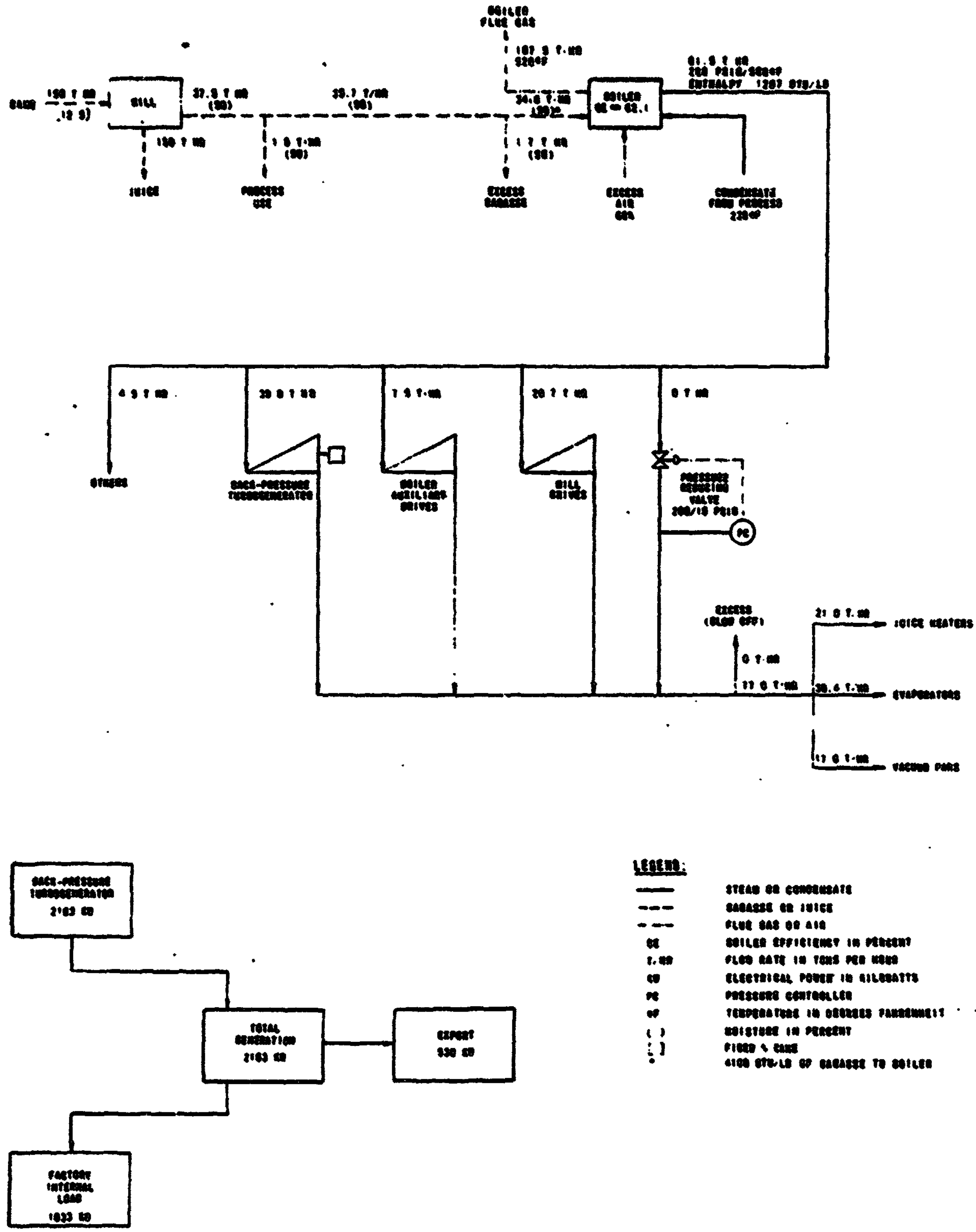
The amount of power generated will remain constant for all the conditions within each scenario since the only change between Conditions 1, 2 and 3 is the inclusion of drying and pelletizing equipment, but the amount of energy available for export will be reduced by the amount required for the drying and pelletizing.

The change in generation from scenario to scenario is the result of the implementation of one or more of the following steps: higher steam pressure and temperature; use of topping and condensing turbo-generators; decrease in process steam consumption. These steps can now be assessed individually or in combination and independently of the drying and pelletizing operations.

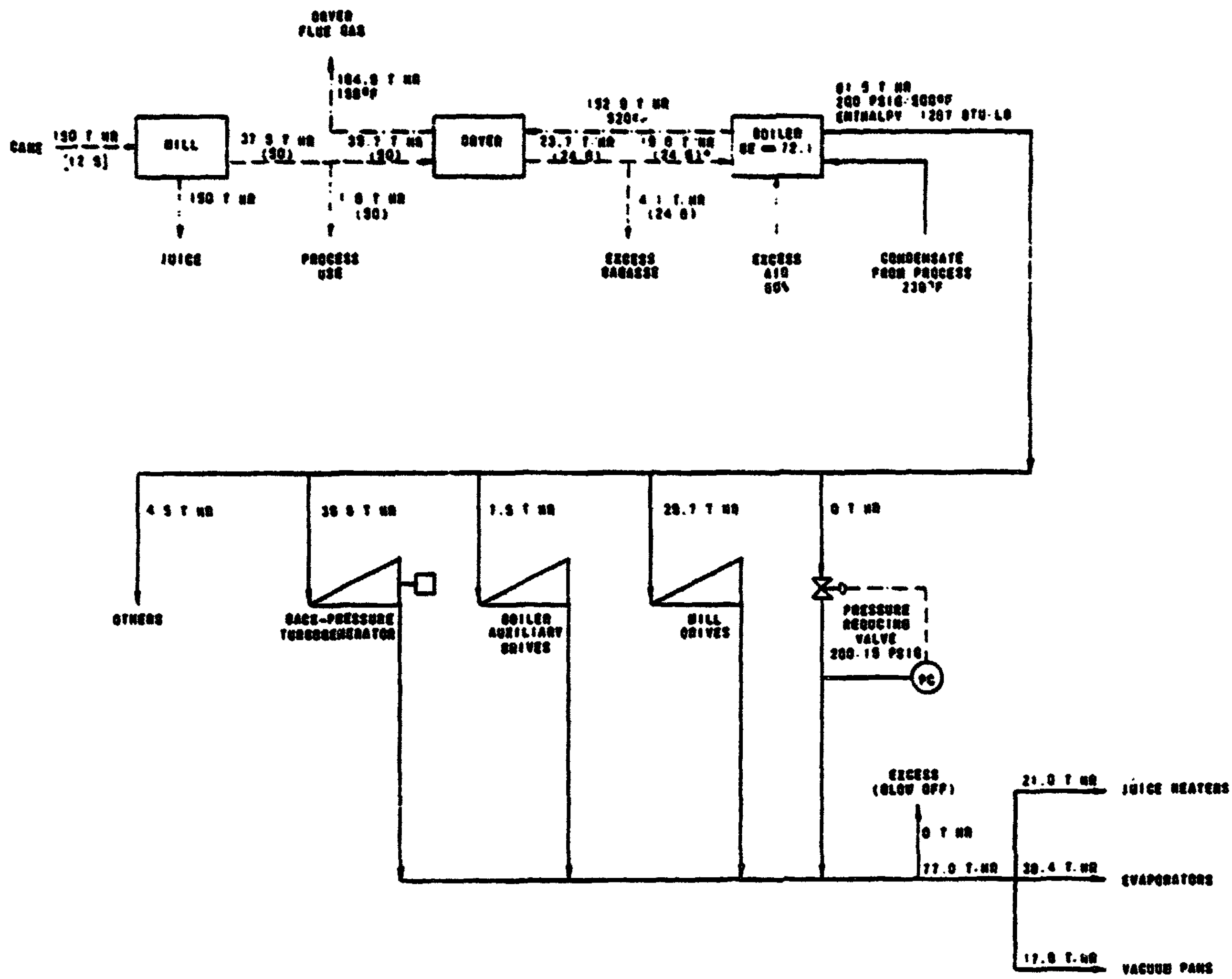
Note: In the case of Scenario 5, because the process steam requirement is still high but due to the higher enthalpy of steam at the higher pressure and temperature, less steam is produced with the available bagasse in Condition 1 and the process steam requirement can no longer be met, hence the necessity to burn oil (or other fuel) to generate the extra steam. In this case the equivalency of oil in terms of bagasse at 50% moisture is shown between brackets. Condition 2 in this case eliminates the need for oil burning and produces a positive supply of dried bagasse and/or pellets.

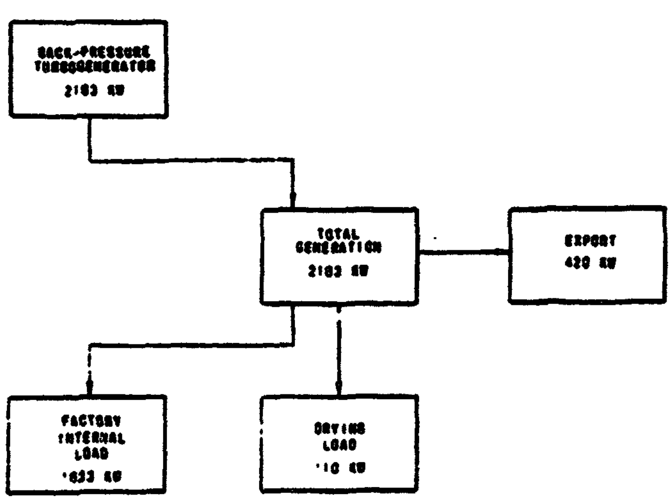
It can be observed from the printouts of the simulations in Appendix X what input is required from the factory to be studied. Given these inputs, the computer will calculate the boiler efficiency, the surplus or deficit of bagasse fuel, the steam and energy balance as well as other pertinent information. If drying is carried out, the computer will give the new boiler efficiency and bagasse moisture (after equilibrium is reached), the amount of surplus bagasse, the power required for drying and pelletizing, and the amount of power generated under the new conditions.

The scenarios are presented in graphic form as follows.



SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 1, CONDITION 1

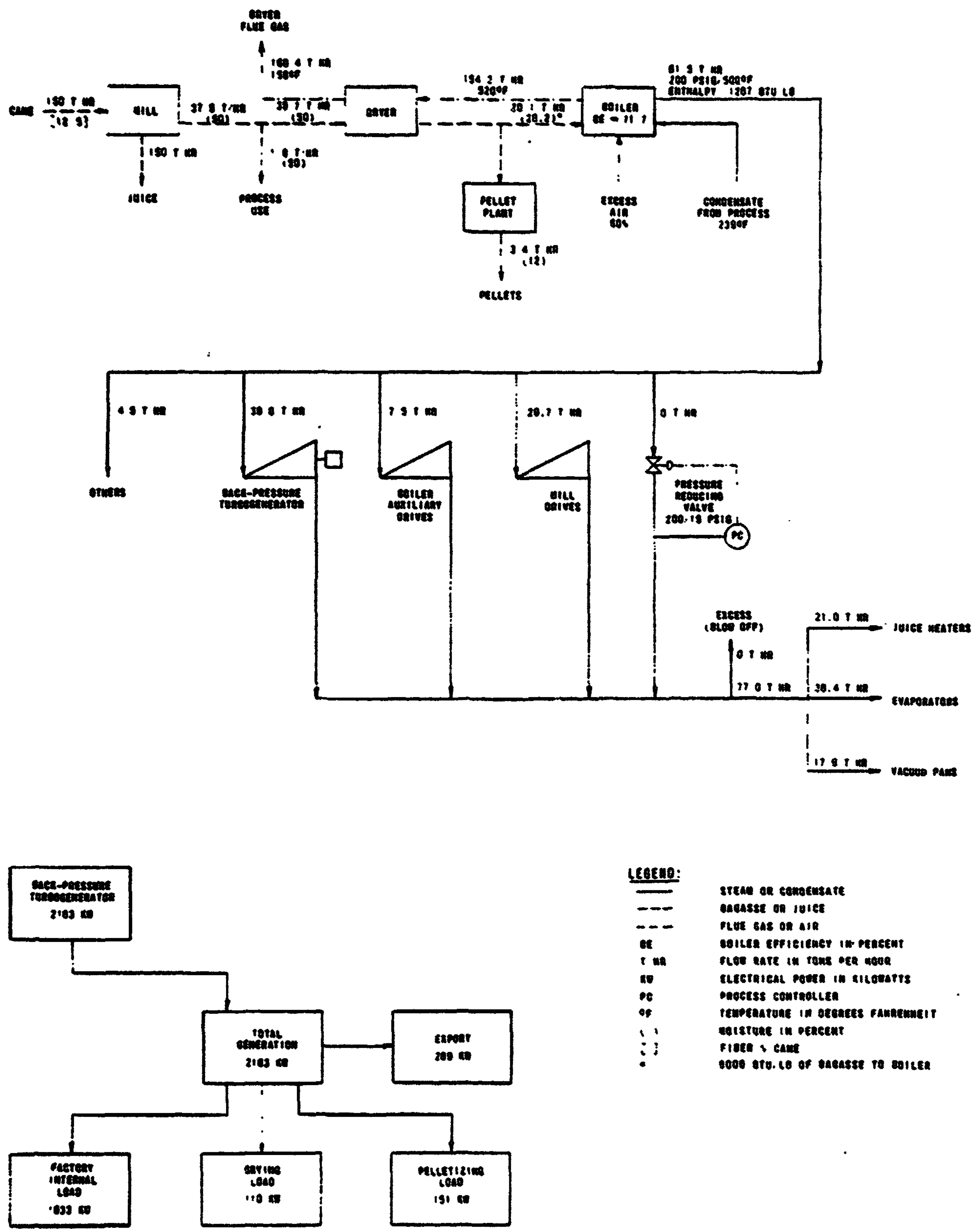




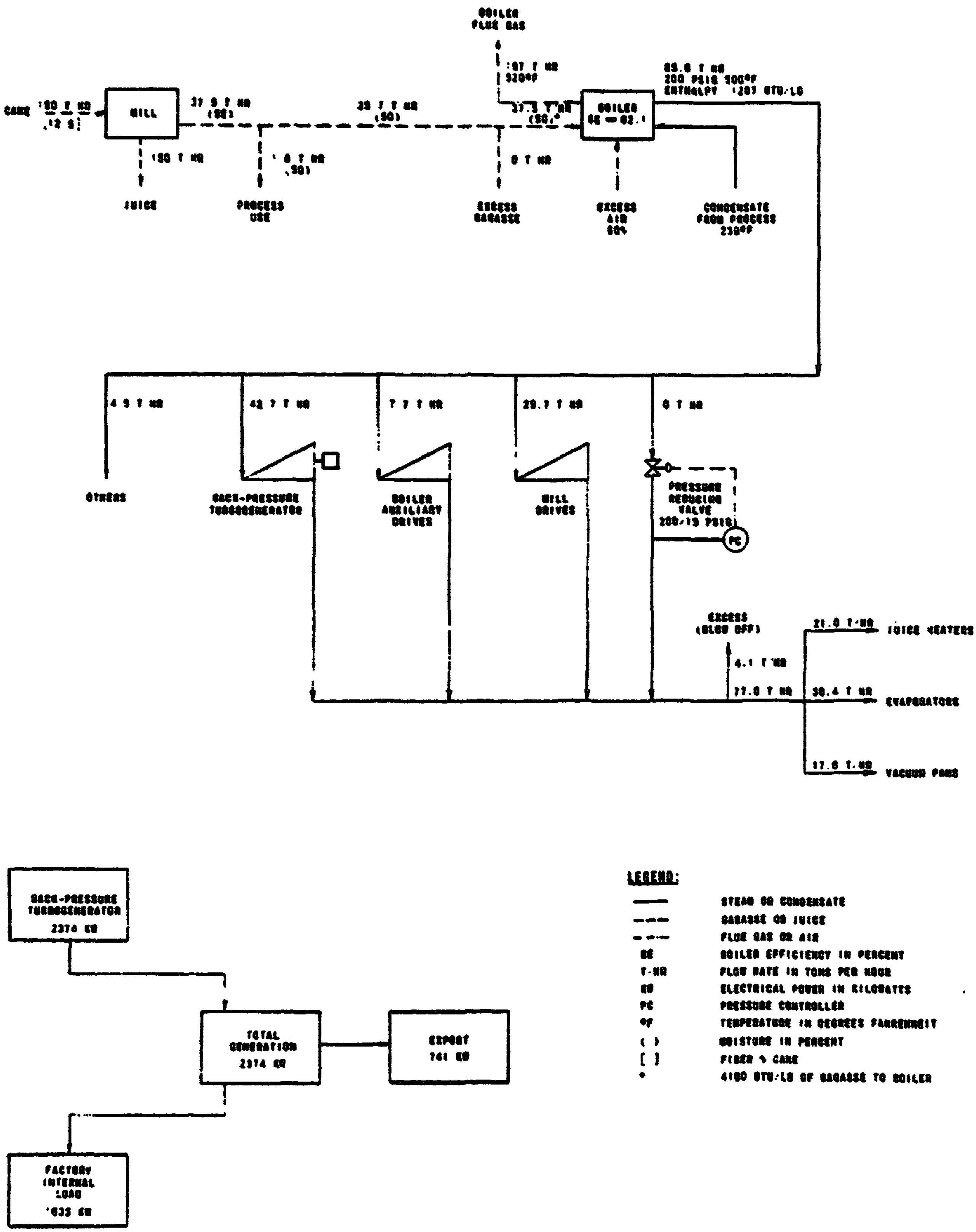
LEGEND:

———	STEAM OR CONDENSATE
— — —	BAGASSE OR JUICE
— - —	FLUE GAS OR AIR
BE	BOILER EFFICIENCY IN PERCENT
T/HR	FLOW RATE IN TONS PER HOUR
KW	ELECTRICAL POWER IN KILOWATTS
PC	PRESSURE CONTROLLER
°F	TEMPERATURE IN DEGREES FAHRENHEIT
( )	MOISTURE IN PERCENT
[ ]	FIBER % CANE
*	8130 BTU/LB OF BAGASSE TO BOILER

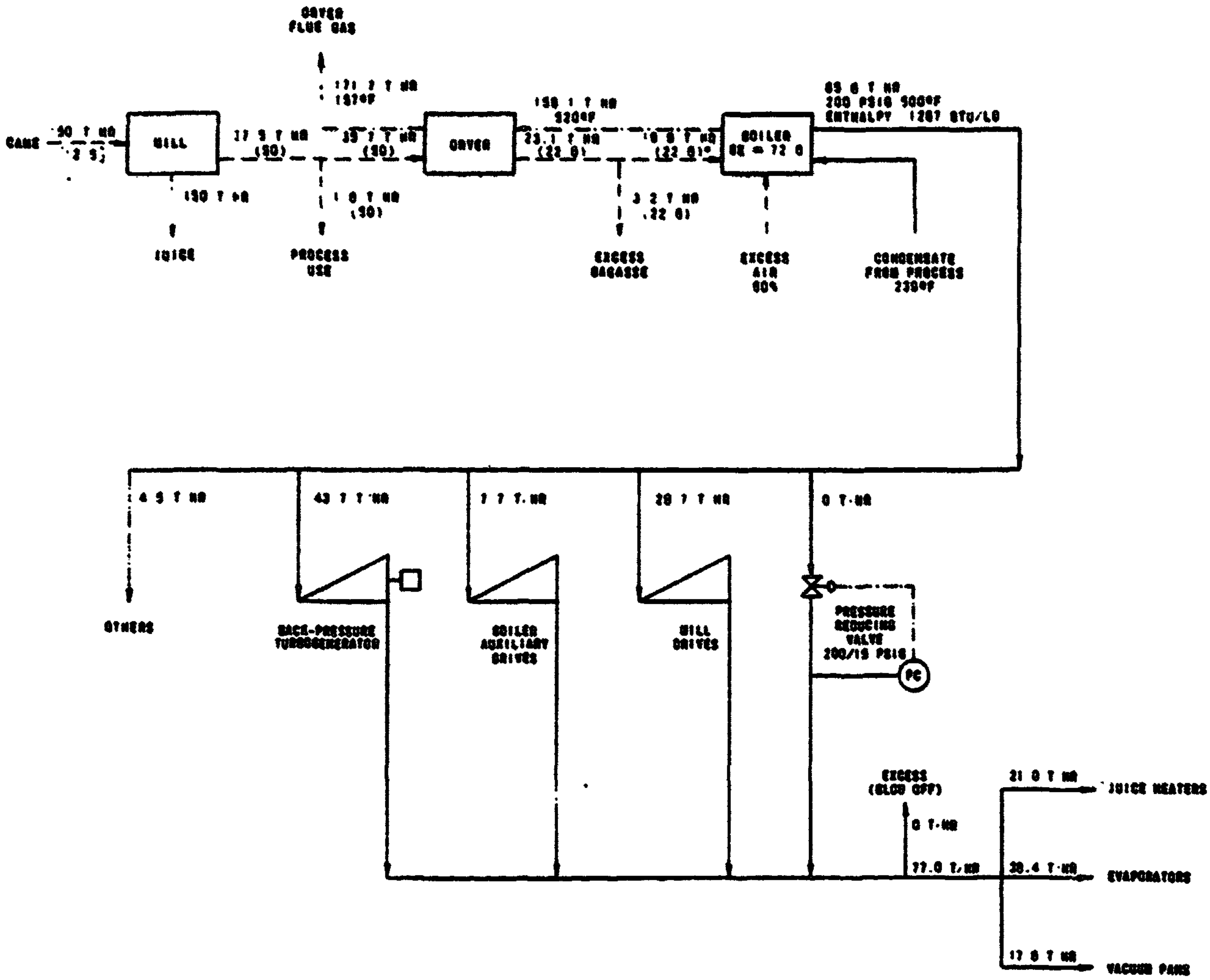
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 1. CONDITION 2

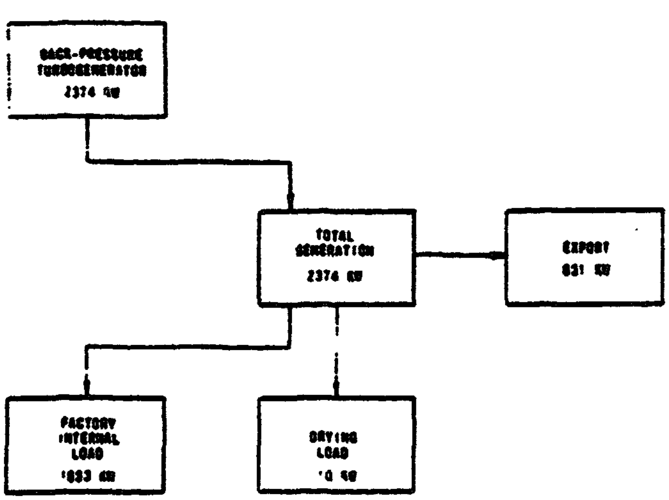


SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO I CONDITION 3



SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 2. CONDITION 1

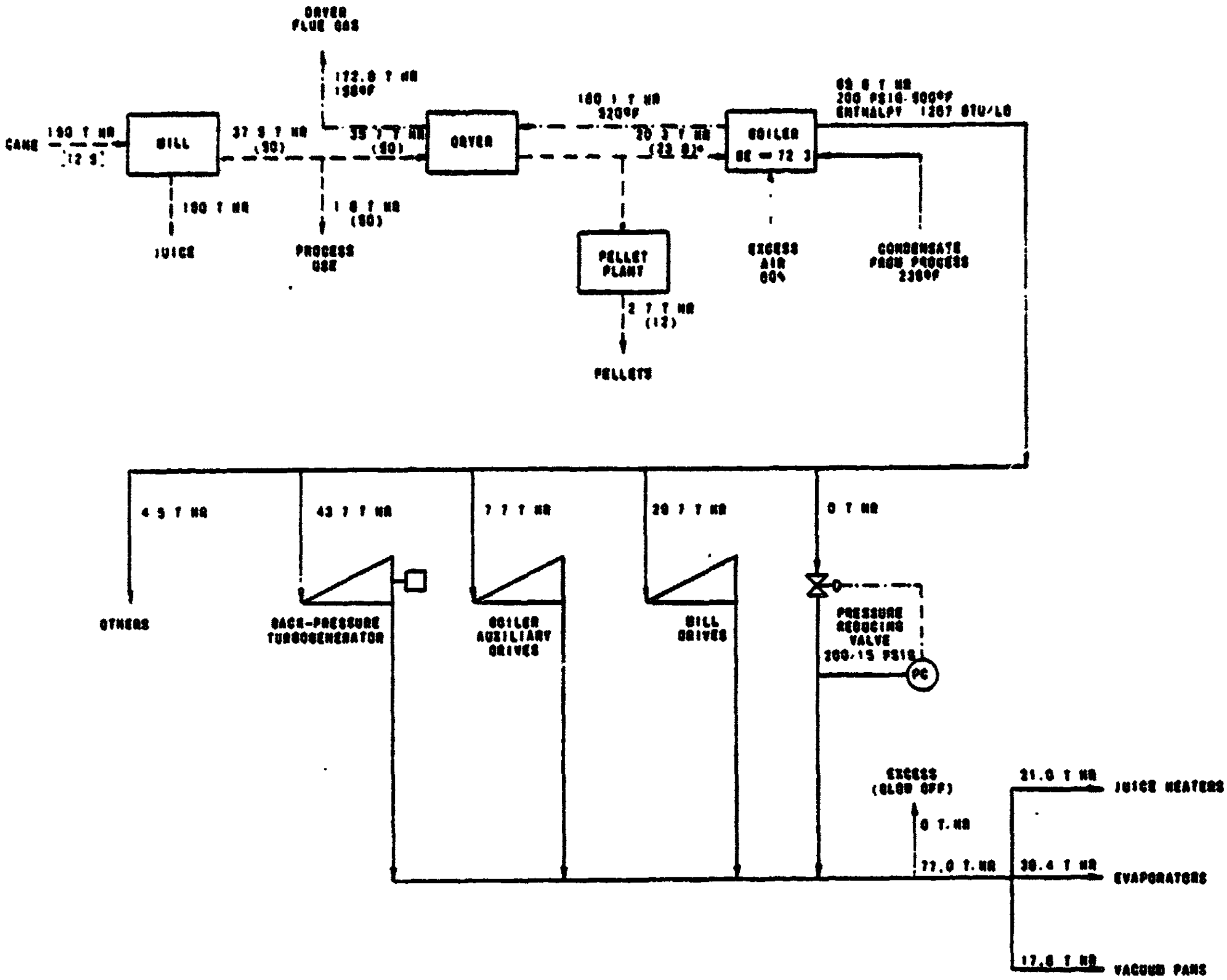


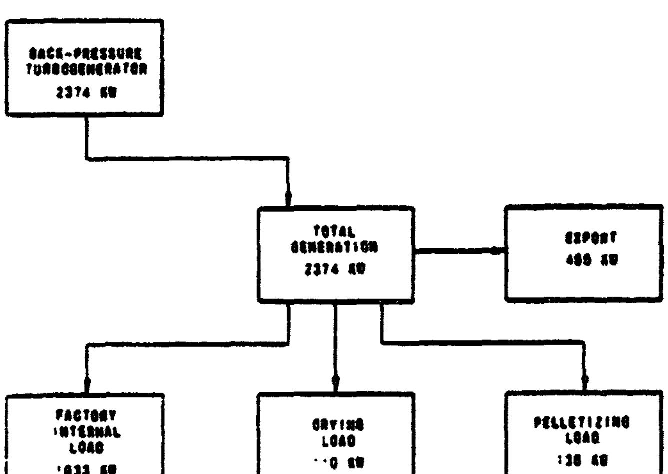


LEGEND:

- STEAM OR CONDENSATE
- BAGASSE OR JUICE
- FLUE GAS OR AIR
- BE — BOILER EFFICIENCY IN PERCENT
- T.HR — FLOW RATE IN TONS PER HOUR
- KW — ELECTRICAL POWER IN KILOWATTS
- PC — PRESSURE CONTROLLER
- °F — TEMPERATURE IN DEGREES FAHRENHEIT
- ( ) — MOISTURE IN PERCENT
- [ ] — FIBER % CANE
- * — 8288 BTU/LB OF BAGASSE TO BOILER

SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 2. CONDITION 2

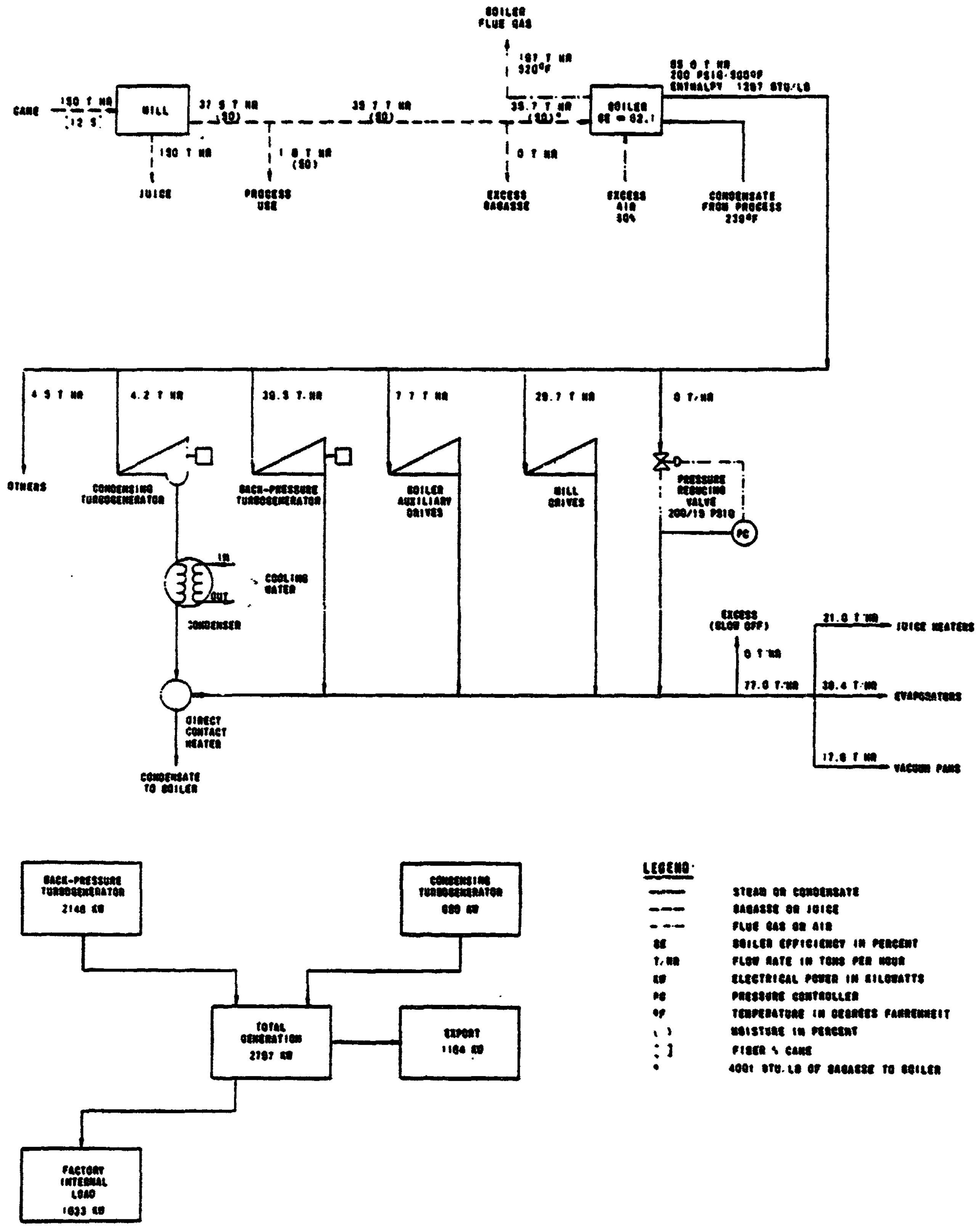




LEGEND:

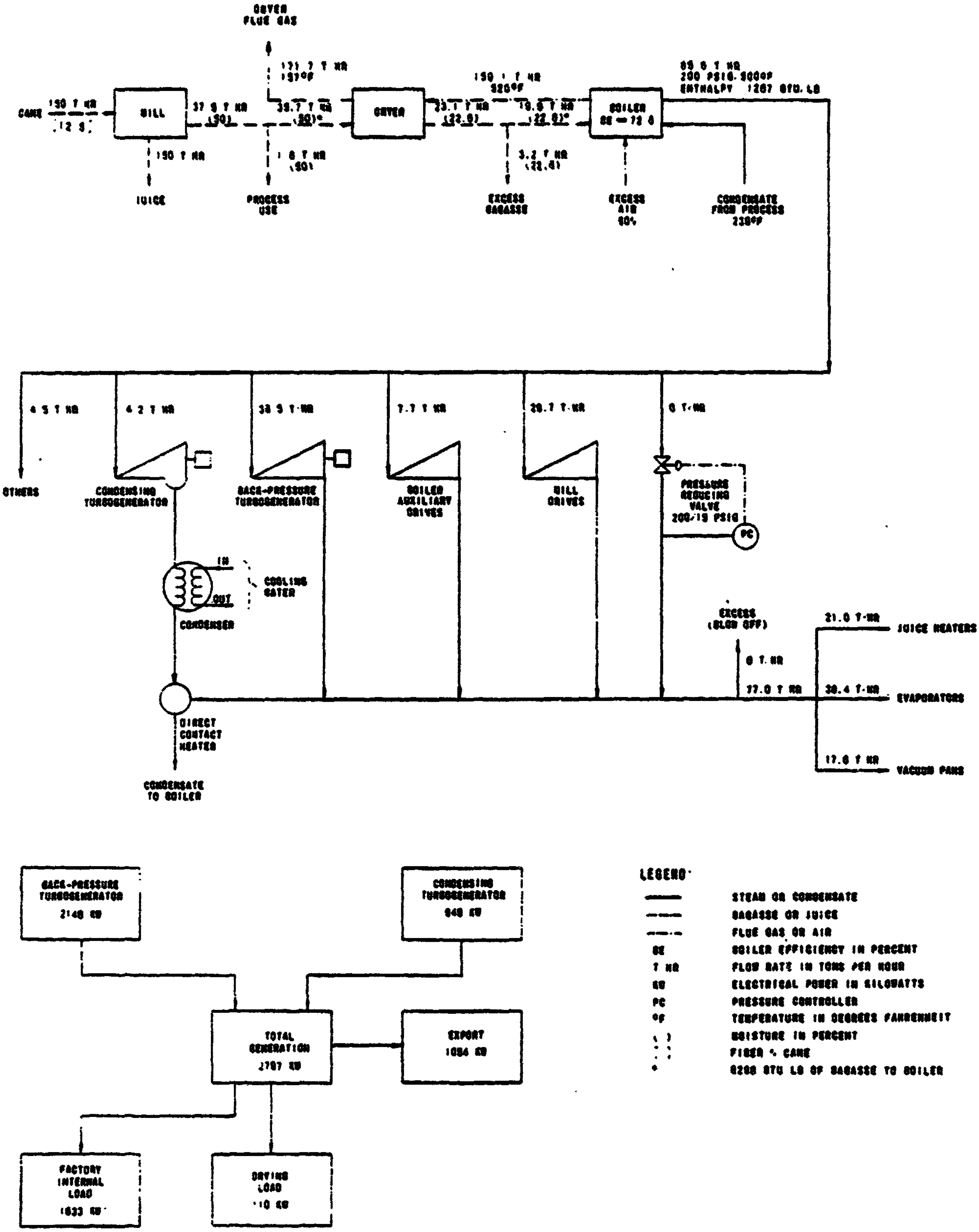
———	STEAM OR CONDENSATE
— — —	BAGASSE OR JUICE
—·—·	FLUE GAS OR AIR
BE	BOILER EFFICIENCY IN PERCENT
T HR	FLOW RATE IN TONS PER HOUR
KW	ELECTRICAL POWER IN KILOWATTS
PC	PRESSURE CONTROLLER
°F	TEMPERATURE IN DEGREES FAHRENHEIT
( )	MOISTURE IN PERCENT
[ ]	FIBER % CANE
*	8202 BTU/LB OF BAGASSE TO BOILER

SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 2. CONDITION 3

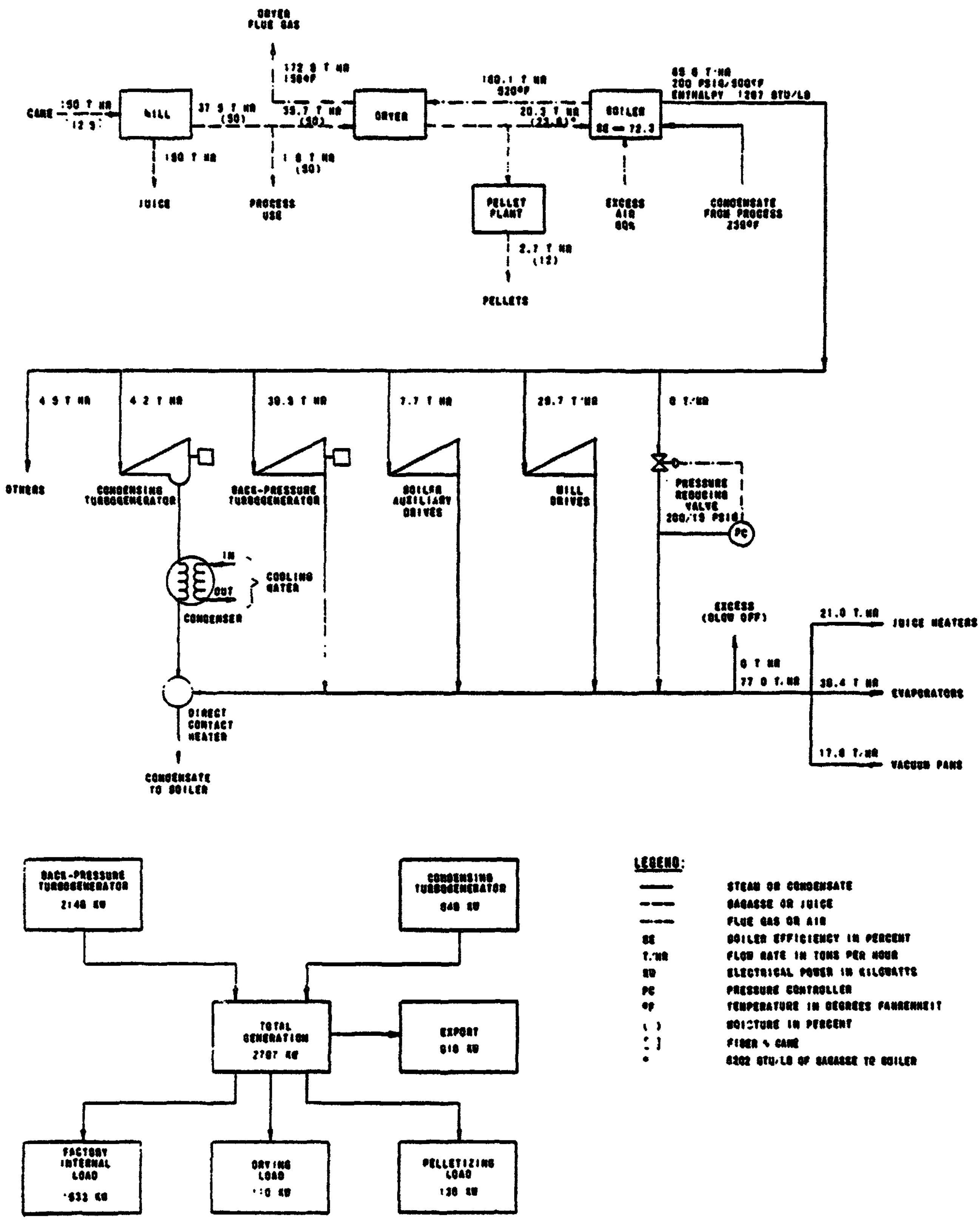


SIMULATION OF FACTORY THERMAL BALANCE

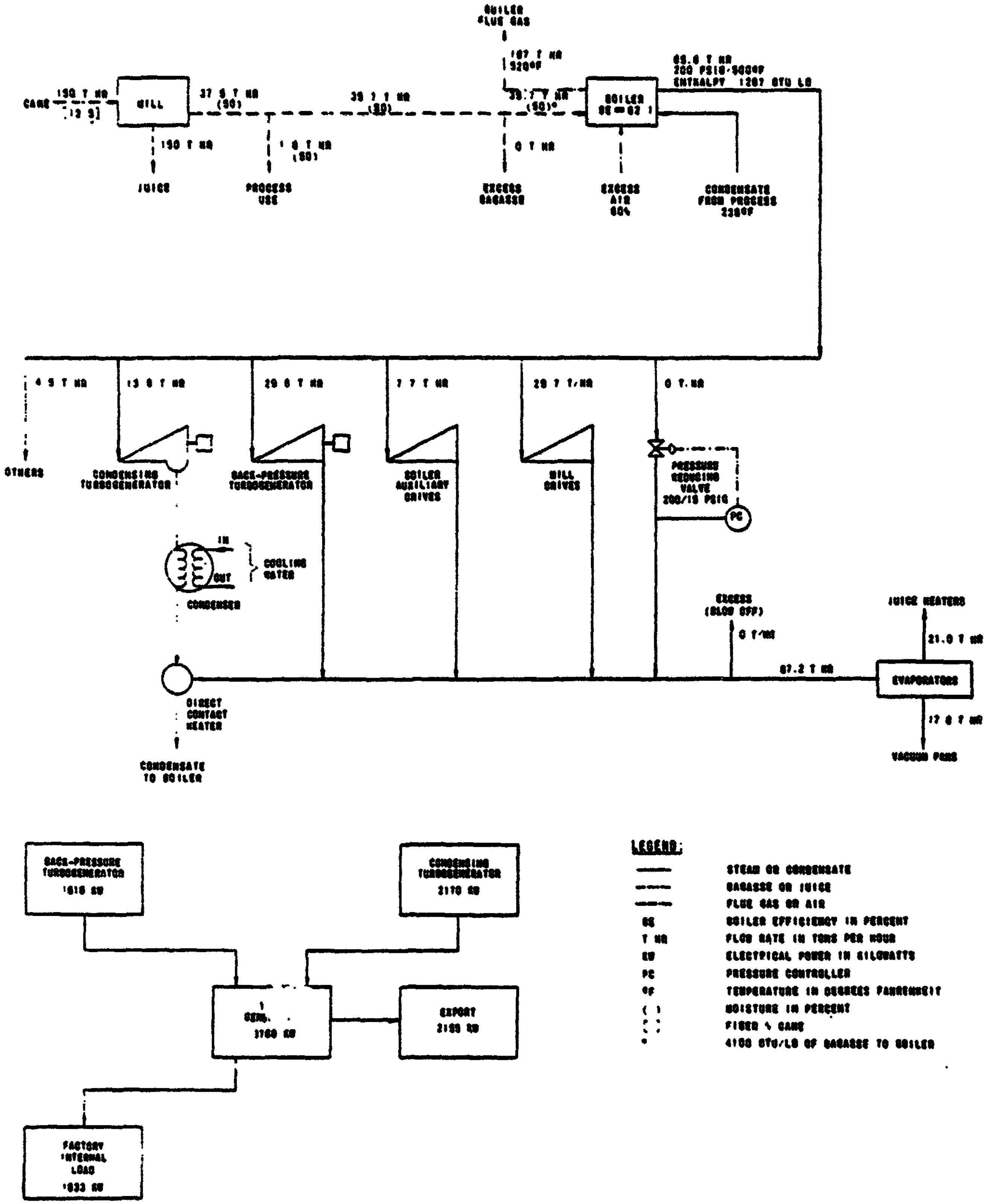
SCENARIO 3. CONDITION 1



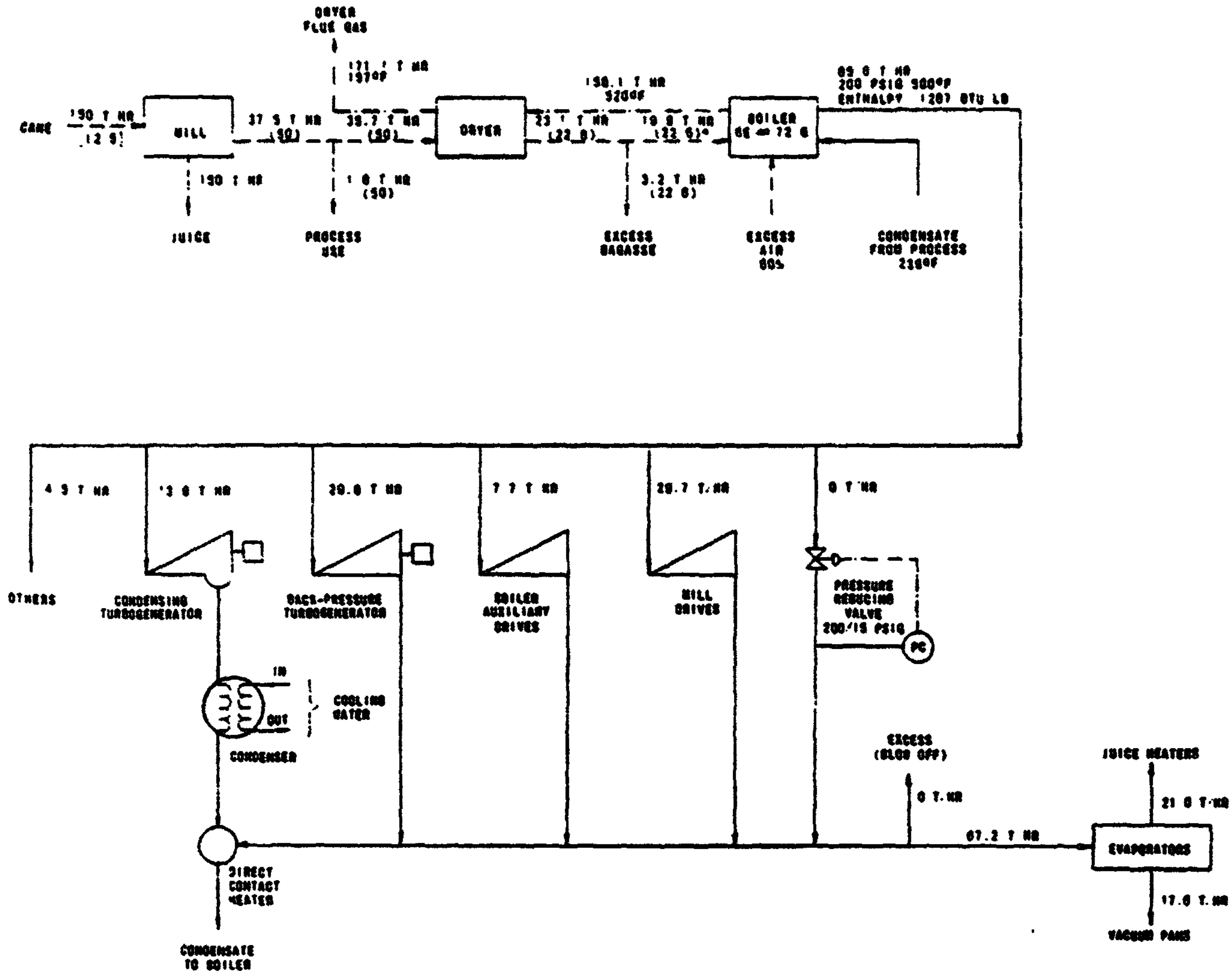
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 3. CONDITION 2

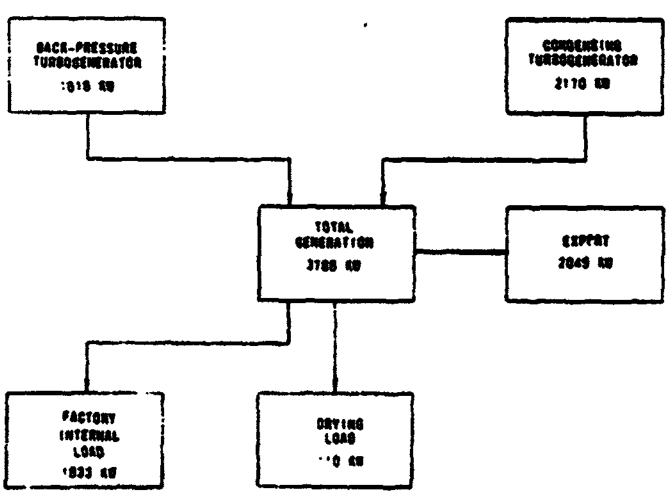


SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 3, CONDITION 3



SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 4. CONDITION 1

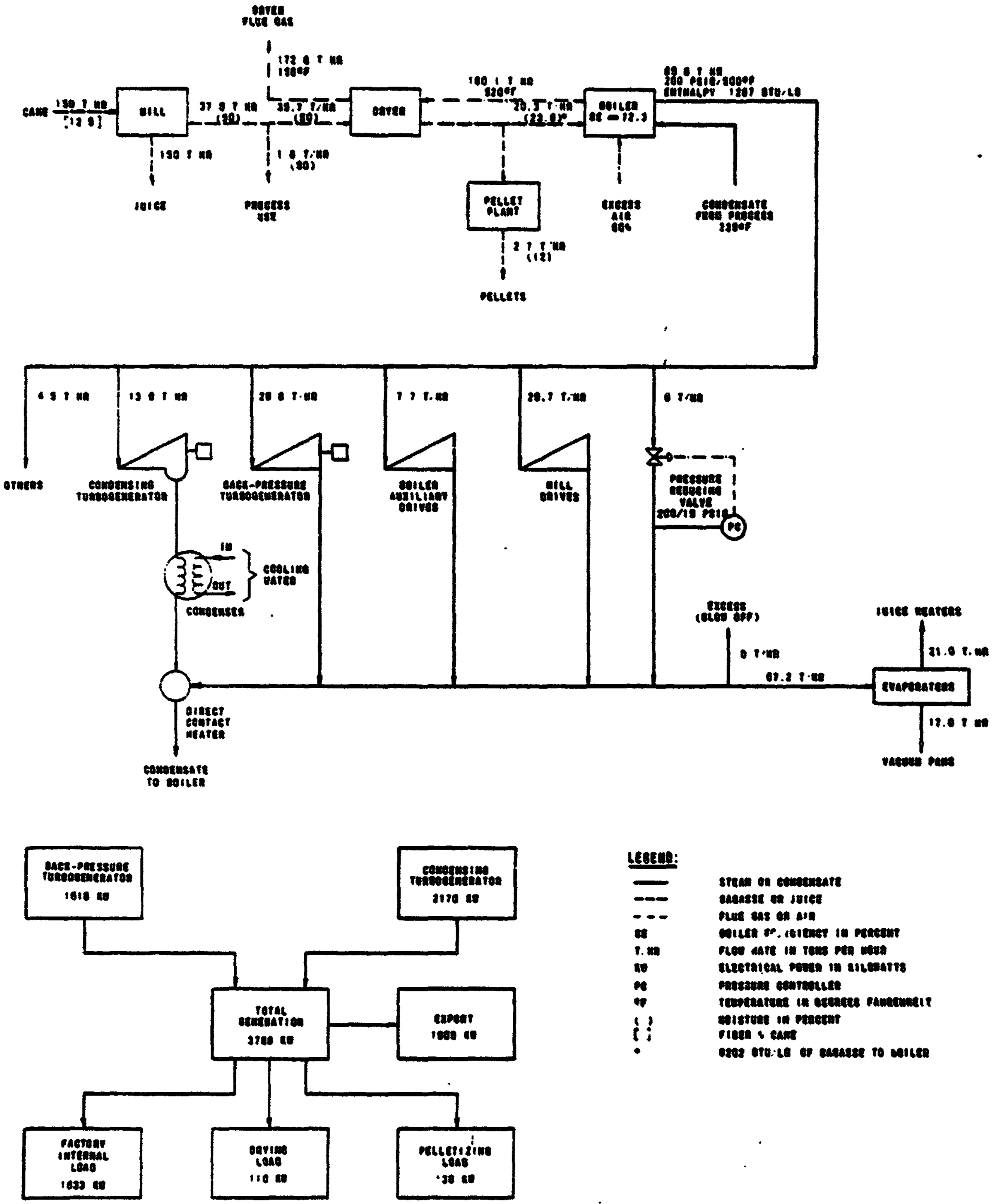




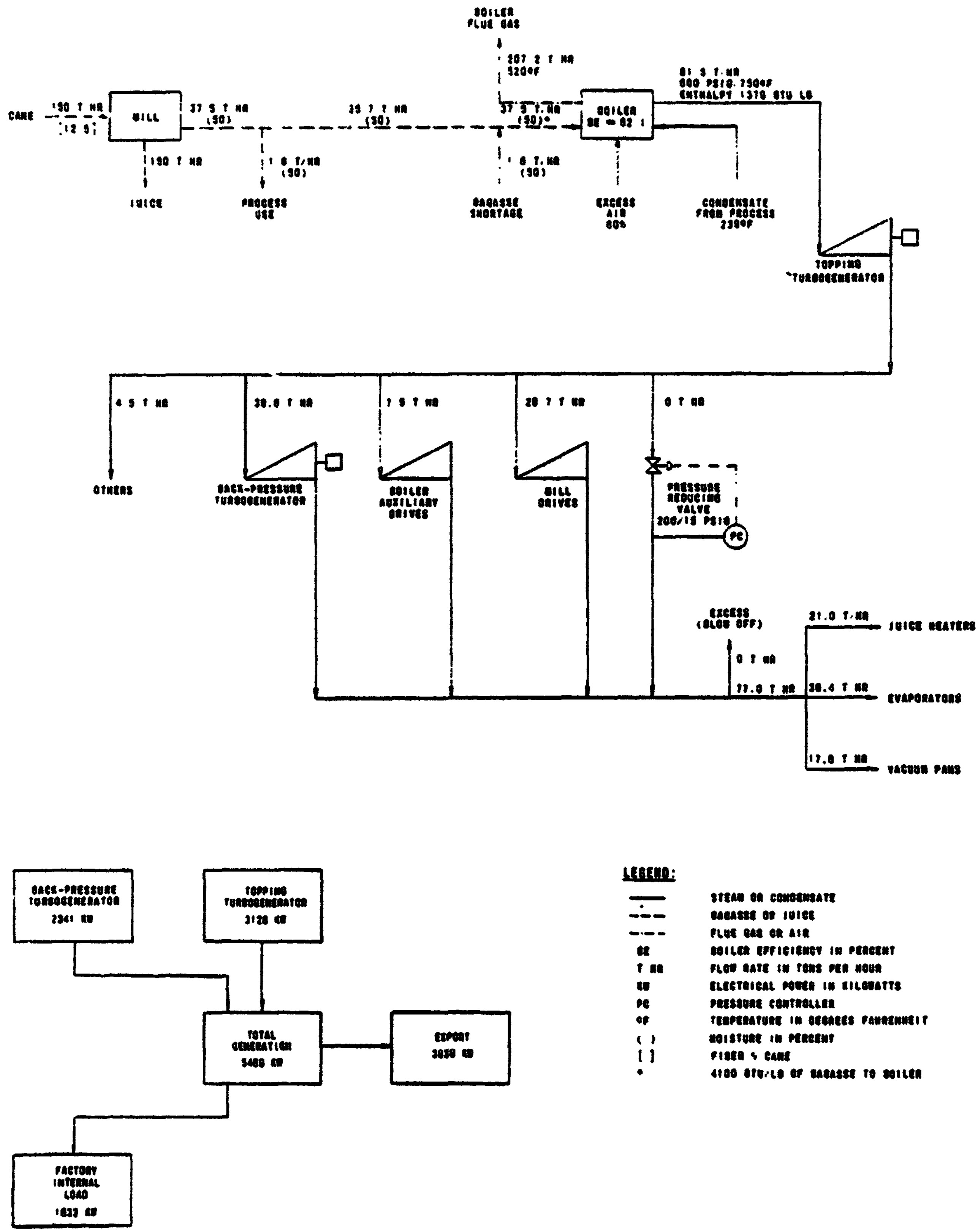
LEGEND:

———	STEAM OR CONDENSATE
— — —	BAGASSE OR JUICE
—·—·—	FLUE GAS OR AIR
BE	BOILER EFFICIENCY IN PERCENT
T HR	FLOW RATE IN TONS PER HOUR
KW	ELECTRICAL POWER IN KILOWATTS
PC	PRESSURE CONTROLLER
°F	TEMPERATURE IN DEGREES FAHRENHEIT
( )	MOISTURE IN PERCENT
: :	FIBER % CANE
*	8288 BTU/LB OF BAGASSE TO BOILER

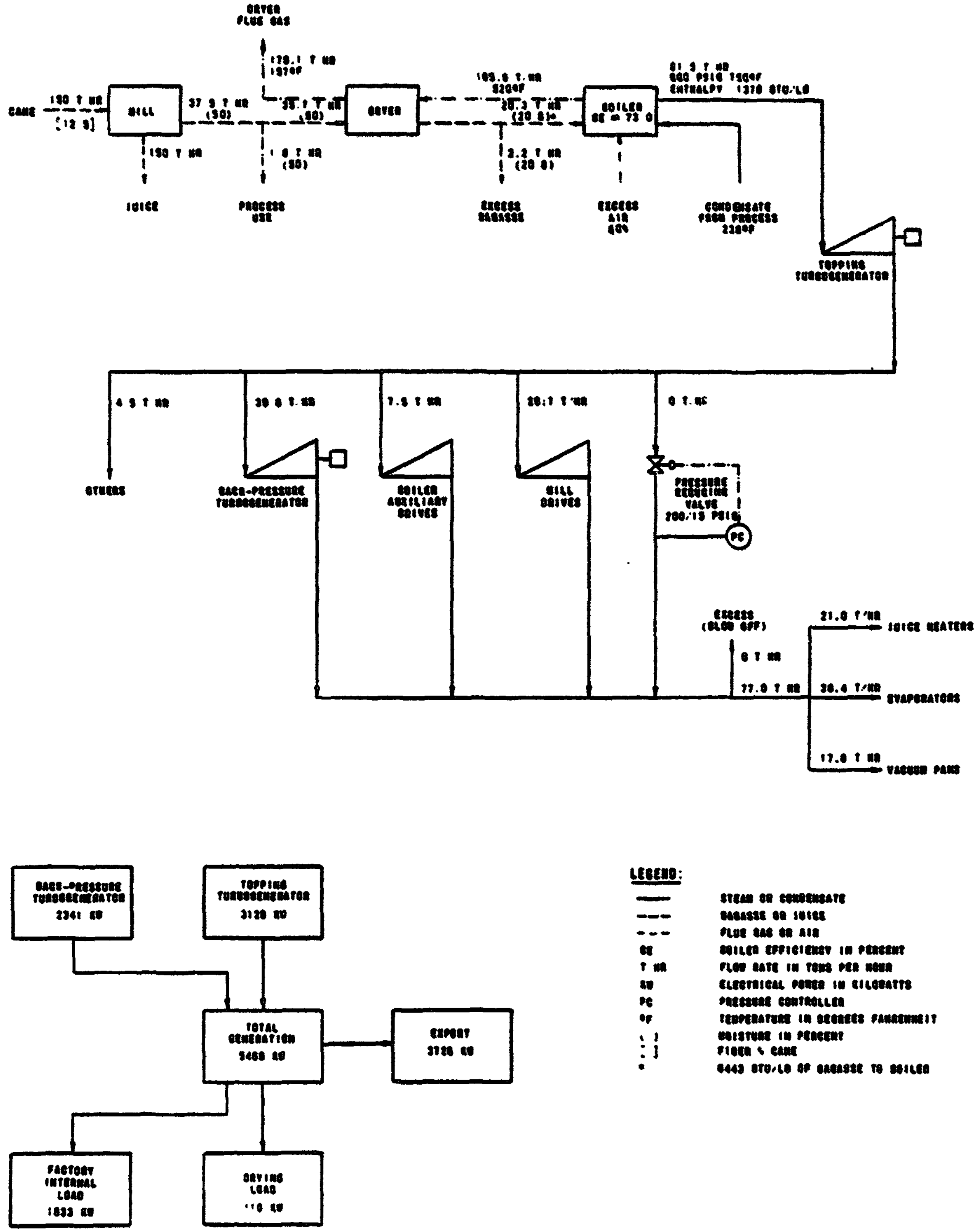
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 4, CONDITION 2



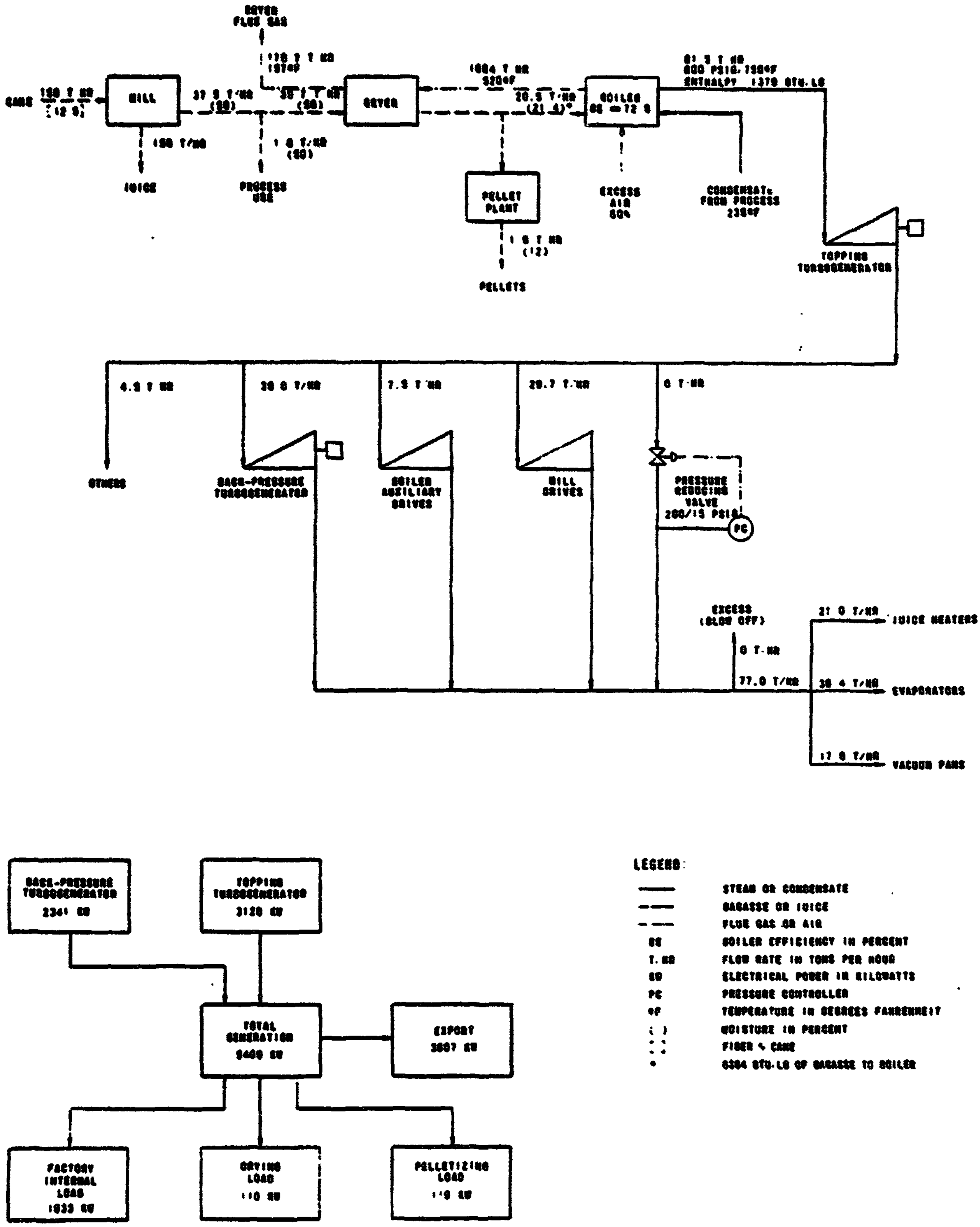
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 4, CONDITION 3



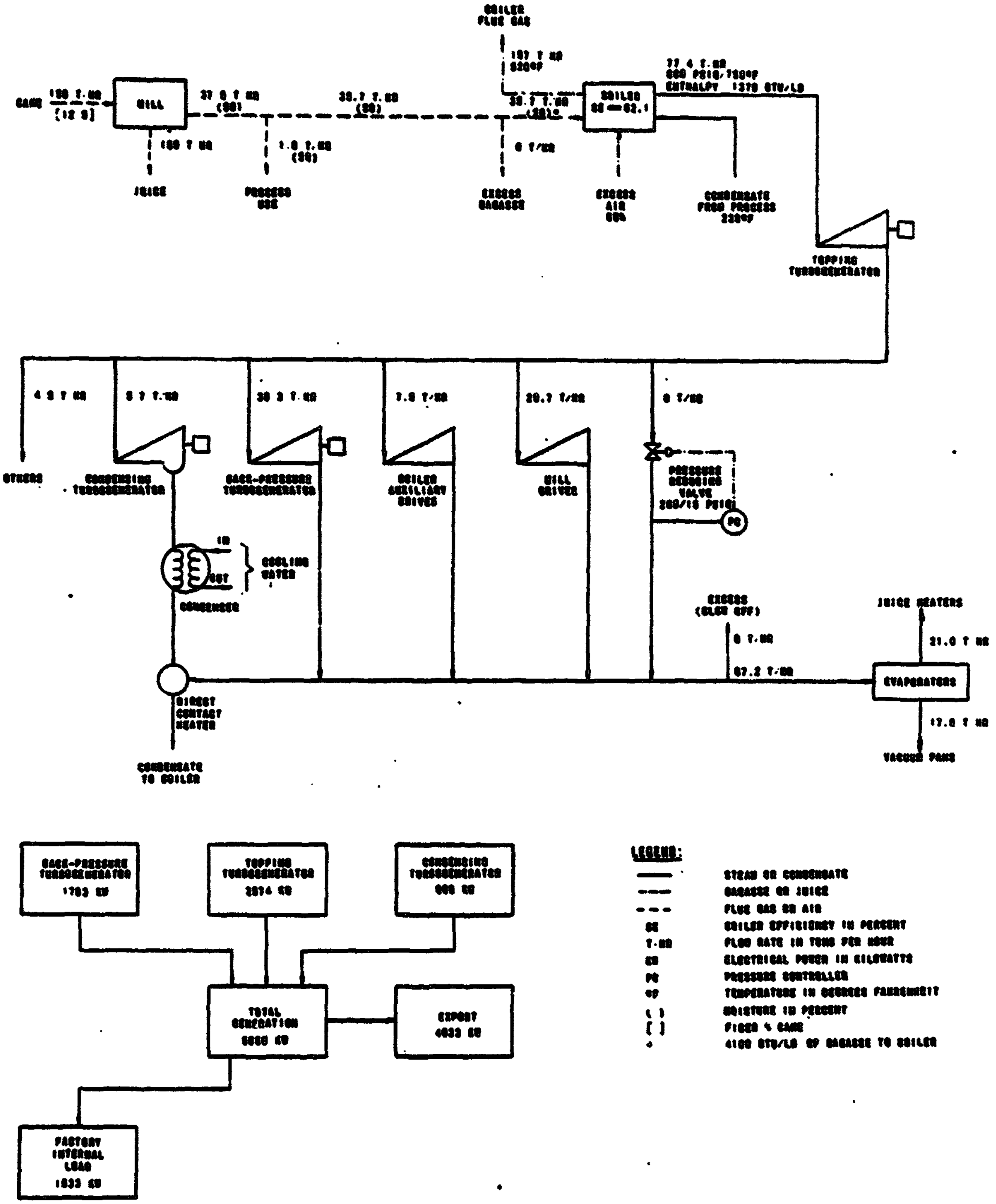
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 5. CONDITION I



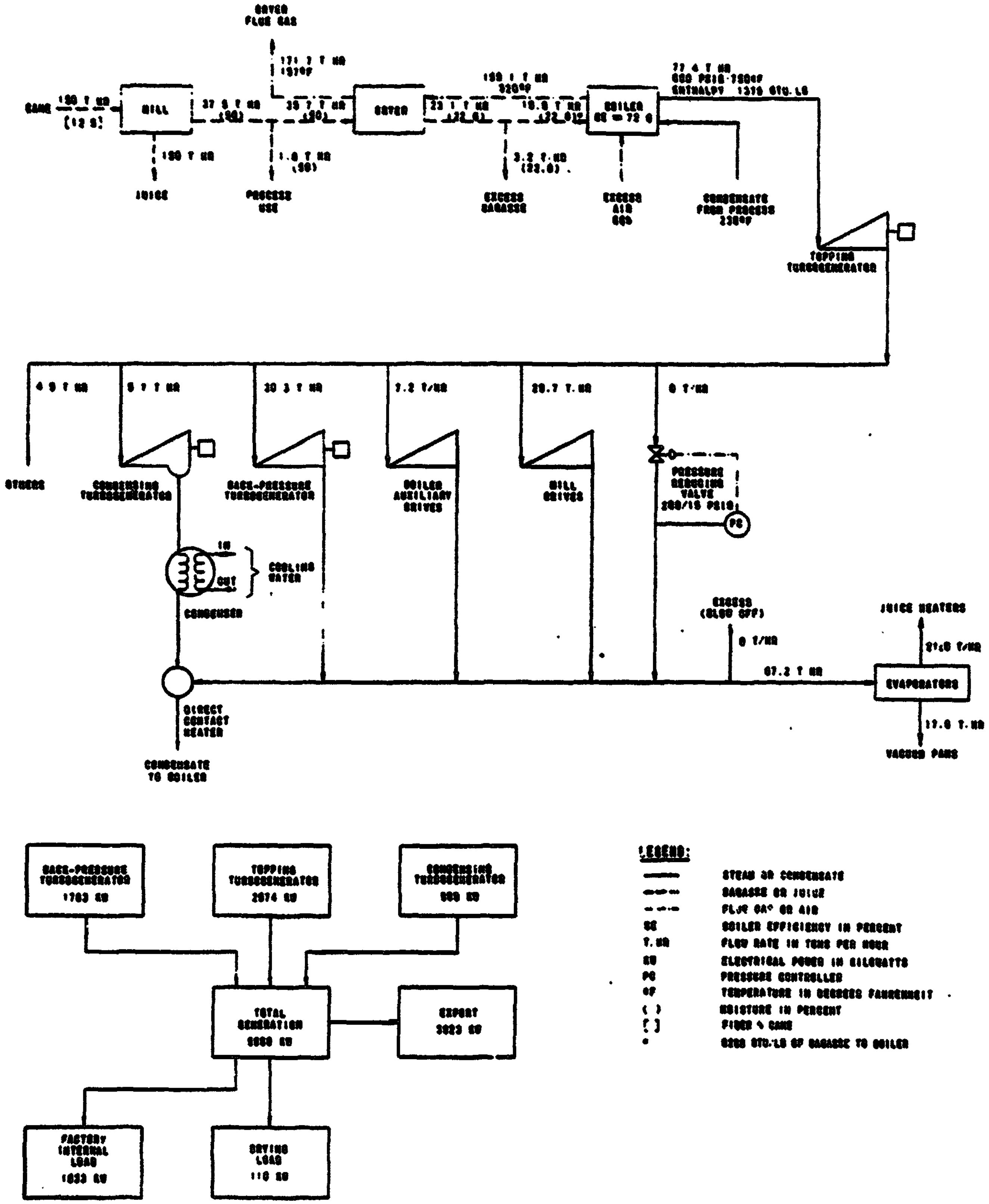
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 5. CONDITION 2



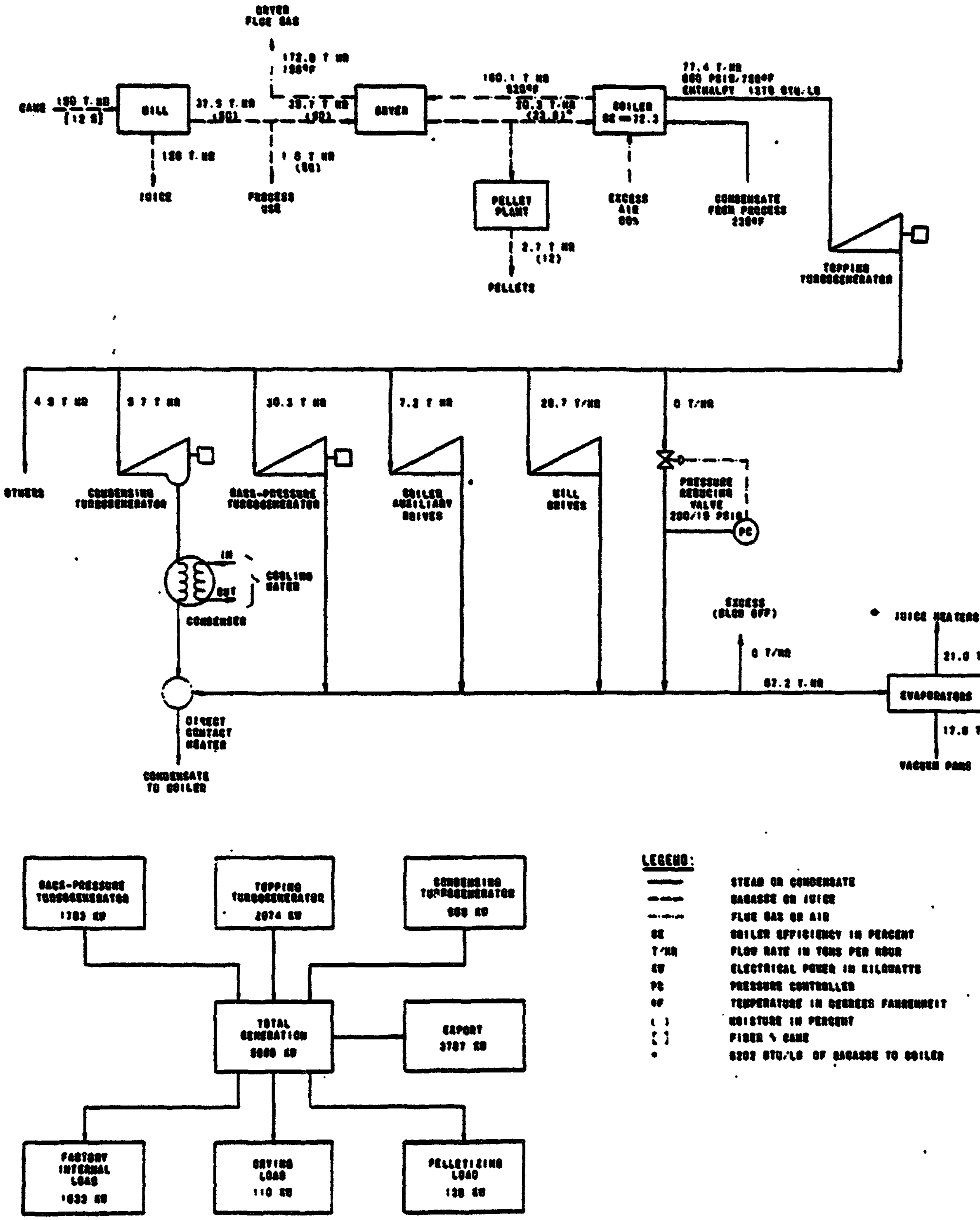
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 5, CONDITION 3



SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 6. CONDITION 1



SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 6, CONDITION 2



<u>**SIMULATION OF FACTORY THERMAL BALANCE**</u>

SCENARIO 6 CONDITION 3

SUMMARY OF COMPUTER SIMULATIONS

	SCENARIO 1			SCENARIO 2			SCENARIO 3			SCENARIO 4			SCENARIO 5			SCENARIO 6		
CONDITION:	1	2	3	1	2	3	1	2	3	1	2	3	1	2	3	1	2	3
SYSTEM:																		
BOILER																		
Steam..............1000 LBS/HR	162.9	162.9	162.9	171.2	171.2	171.2	171.2	171.2	171.2	171.2	171.2	171.2	162.9	162.9	162.9	154.9	154.9	154.9
LBS Steam/LBS Bagasse	2.4	4.2	4.1	2.4	4.3	4.2	2.4	4.3	4.2	2.4	4.3	4.2	2.2	4.0	4.0	2.2	3.9	3.8
Moisture (Feed Bagasse)......%	50.0	24.6	25.2	50.0	22.6	23.8	50.0	22.6	23.8	50.0	22.6	23.8	50.0	20.8	21.4	50.0	22.6	23.8
DRYER																		
Dried Bagasse......1000 LBS/HR	-0-	47.2	47.0	-0-	46.0	45.9	-0-	46.0	45.9	-0-	46.0	45.9	-0-	45.0	45.0	-0-	46.0	45.9
Moisture (Dried Bagasse).....%	---	24.6	24.1	---	22.6	22.4	---	22.6	22.4	---	22.6	22.4	---	20.8	20.6	---	22.6	22.4
PELLETIZATION																		
Pellets............1000 LBS/HR	-0-	-0-	6.8	-0-	-0-	5.4	-0-	-0-	5.4	-0-	-0-	5.4	-0-	-0-	3.9	-0-	-0-	5.4
Moisture......................%	---	---	12.0	---	---	12.0	---	---	12.0	---	---	12.0	---	---	12.0	---	---	12.0
EXCESS BAGASSE																		
Bagasse............1000 LBS/HR	3.5	8.2	-0-	-0-	6.3	-0-	-0-	6.3	-0-	-0-	6.3	-0-	(3.7)	4.4	-0-	-0-	6.3	-0-
LBS Steam/LBS Bagasse	2.4	4.2	---	---	4.3	---	---	4.3	---	---	4.3	---	---	4.0	---	---	3.9	---
Moisture......................%	50.0	24.6	---	---	22.6	---	---	22.6	---	--	22.6	---	---	20.8	---	---	22.6	---
STEAM RATE																		
Process..................LBS/TC	1026	1026	1026	1026	1026	1026	1026	1026	1026	896	896	896	1026	1026	1026	896	896	896
Other....................LBS/TC	60	60	60	60	60	60	60	60	60	60	60	60	60	60	60	60	60	60
STEAM REQUIREMENT																		
Process............1000 LBS/HR	153.9	153.9	153.9	153.9	153.9	153.9	153.9	153.9	153.9	134.4	134.4	134.4	153.9	153.9	153.9	134.4	134.4	134.4
Excess (Blow Off)..1000 LBS/HR	-0-	-0-	-0-	8.3	8.3	8.3	-0-	-0-	-0-	-0-	-0-	-0-	-0-	-0-	-0-	-0-	-0-	-0-
Other..............1000 LBS/HR	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0	9.0
Condensing TG......1000 LBS/HR	---	---	---	---	---	---	8.3	8.3	8.3	27.8	27.8	27.8	---	---	---	11.5	11.5	11.5
TOTAL......1000 LBS/HR	162.9	162.9	162.9	171.2	171.2	171.2	171.2	171.2	171.2	171.2	171.2	171.2	162.9	162.9	162.9	154.9	154.9	154.9
ELECTRICAL POWER																		
Generation.....................kW	2163	2163	2163	2374	2374	2374	2797	2797	2797	3788	3788	3788	5469	5469	5469	5666	5666	5666
Consumption....................kW	1633	1743	1895	1633	1744	1879	1633	1743	1879	1633	1743	1879	1633	1743	1862	1633	1743	1879
Export.........................kW	530	420	268	741	631	495	1164	1054	918	2155	2045	1909	3836	3726	3607	4033	3923	3787

CONSTANT PARAMETERS		VARIABLE PARAMETERS	
MILL CONDITIONS (For All Scenarios)	BOILER CONDITIONS (For All Scenarios)	BOILER CONDITIONS (For Scenarios 1 to 4)	BOILER CONDITIONS (For Scenarios 5 to 6)
95% of Mill Bagasse Available for Steam Generation. i.e. 71,250 LBS/HR @ 50% Moisture	Flue Gas Temperature @ 520°F Boiler Feed Water @ 239°F Excess Air @ 60%	Steam @ 200 psig/500°F Steam Enthalpy @ 1267.48 BTU/LB	Steam @ 600 psig/750°F Steam Enthalpy @ 1379 BTU/LB

Economic Analysis and Evaluation Using Discounted Cash Flow Rate of Return (DCFRR)

By using a generalized model, one can estimate the profitability and analyze the sensitivity of processes. The technique for assessing the relative investment return of potential projects requires a projection of future cash flows which are then reduced to a present value or rate of return.

In this presentation, the discounted cash flow rate of return (DCFRR) will be used to estimate the revenue-to-capital ratio to establish the profitability of various processes. A DCFRR of, say, 15% implies that 15% per year will be earned on the investment, in addition to which the project generates sufficient money to repay the original investment. The cutoff point of the DCFRR is established by management policy and differs from company to company.

The payback period, which is not a real measure of profitability but of time it takes for the cash flow or net annual income before taxes to recoup the original fixed-capital expenditure, will be provided only as an indicator.

To compute the DCFRR, the following components must be identified:

1) Total capital cost, consisting of fixed-capital cost, erection cost, cost of land and other non-depreciable costs;
2) Depreciation method;

3) Revenues;
4) Operating expenses.

From (1) the cash outflow is calculated by applying to the total capital cost the specific year discount factor. From (2), (3) and (4) the net annual income before taxes is computed. Specific year discount factors are then applied to the latter to obtain the net cash outflow. Although a single value of DCFRR is computed from a given set of cash flow data, which are usually subjective estimates of sales revenue, total expenses, fixed capital cost, etc., sensitivity analysis of the profitability is beyond the scope of this project and will not be attempted. The relevant costs (and benefits) are the opportunity costs of the economy of the specific inputs and outputs. Thus taxes are not included and shadow pricing of some inputs (wages, energy) may be appropriate. However, the authors believe that the above mentioned components have been derived from well documented and established processes, thus minimizing the degree of risk. A review of the construction of each of these components is presented below.

Total Capital Cost

- Working capital and cost of land have not been taken into consideration.

- The fixed-capital cost is based on existing designs to which scaling factors have been applied. Collection of costs was from literature, company records and quotations.

- A factor of 1.03 was applied to the purchased cost of equipment to approximate the delivered cost.

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3) Revenues;

4) Operating expenses;

5) Tax rate.

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- The fixed-capital cost is based on existing designs to which scaling factors have been applied. Collection of costs was from literature, company records and quotations.
- A factor of 1.03 was applied to the purchased cost of equipment to approximate the delivered cost.

- A factor of 1.17 to 1.25 was applied to the delivered cost of large single unit processes to approximate the installed costs of equipment, e.g. evaporator, boiler, turbo-generator. The factor was increased to 1.35 to 1.37 for a complex process containing a combination of unit processes, e.g. drying and pelletizing.
- Cost of engineering is 3% to 5% of installed cost for single unit processes and 9% to 11% for complex processes.

Depreciation Method

- Straight-line based on 15 years.

Revenues

- These will be expressed as incremental revenues due to the fact that the various processes are added on to a basic factory. The three products that provide the revenues are:

  Dried bagasse;

  Pellets;

  Exported electrical power (KWH).
- The wet bagasse is considered as a zero revenue item for reasons mentioned earlier in this presentation. In the design of the various scenarios however, if an excess of wet bagasse was found, rather than attributing a zero value to it another scenario was created in which all the excess wet bagasse is burned to produce additional steam converted into exportable electrical energy to which a value is applied.

- For the purpose of establishing a common base in the economic evaluation of the various scenarios and conditions, respective dollar values are attributed to the respective three products.

    1) Exported Electrical Power

       The revenue per KWH is taken as \$0.06, which is assumed to be 90% of the avoided cost of the utility company. The replacement value if purchased from a utility company would probably be \$0.12/KWH.

    2) Dried Bagasse and Pellets

       For these an equivalency is determined and expressed as "recoverable BTU's in steam" from the assumption that, if additional steam had to be produced for power generation, other fuel would have to be purchased. The standard comparison taken is number 6 bunker fuel oil, and the recoverable BTU in steam resulting from the burning of this fuel oil is calculated as follows.

       BTU/lb. of oil = 18,300

       Cost per barrel = \$38

       Boiler efficiency when burning oil = 86%

       Cost of 10^6 BTU recovered in steam from oil burning

       $$= \frac{38 \times 10^6}{42 \times 8.5 \times 18{,}300 \times 0.86}$$

       $$= \$6.7634$$

Therefore, the products are evaluated as follows.

a) Dried Bagasse

Lb./hr. of dried bagasse = 6,310

Moisture % dried bagasse = 22.6%

BTU/lb. of dried bagasse = 6,298

Boiler efficiency = 72.59%

Recoverable BTU/hr. in steam from dried bagasse

= 6,310 x 6,298 x 0.7259

= 28.848 x 10^6 BTU/hr.

∴ Value of dried bagasse/hr. (in terms of oil cost)

= 28.848 x 6.7634

= $195.11

Value of dried bagasse/year (for crop of 2,640 hours)

= $515,100

Note: This value will vary according to the moisture content of the dried bagasse.

b) Pellets

Lb./hr. of pellets = 5,412

Moisture % pellets = 12%

BTU/lb. of pellets = 7,148

Boiler efficiency = 82.01%

Recoverable BTU/hr. in steam from pellets

= 5,412 x 7,148 x 0.8201

= 31.727 x 10^6 BTU/hr.

.˙. Value of pellets/hr. (in terms of oil cost)

= 31,727 x 6.7634

= $214.58

Value of pellets/year (for crop of 2,640 hours)

= $566,490

Operating Expenses

- Although the manufacturing cost of a product is the sum of the processing or conversion cost and the cost of raw material, this presentation excludes the latter because it is considered that the raw material, mill bagasse, is a by-product of the sugar manufacturing process, which absorbs the associated cost.

- Labor cost consists of both direct and indirect cost components. The direct component is based on a rate of $4.50/man-hour but it excludes the cost of supervision, which is assumed absorbed by the sugar manufacturing process as the add-on processes are integrated with the latter. The indirect component, i.e. benefits, is at $2/man-hour.

- Operating material cost is based on documented actual operational conditions of the add-on process. The cost of utilities excludes the six months of operation power requirement as the final products have been discounted by the power requirement equivalency.

- The overhead cost, excluding those costs accounted for above, is also considered absorbed by the sugar manufacturing process due to the integrating nature of the add-on process.

- The cost of maintenance is an investment related cost and, again from available documentation and records, is taken at 1.5% of the major equipment cost, half of which is for the associated labor.

- The non-controllable fixed costs include depreciation but exclude real estate taxes and insurance cost. Should we consider the accounting of the exclusions, the following ranges are recommended:

  Indirect manufacturing cost: plant overhead, which includes the cost of control, safety, medical, etc., at 50% to 150% of labor cost.

  Fixed manufacturing cost:
  Property tax at 2% of direct capital cost of the process;
  Insurance at 1% of the direct capital cost of the process.

Inputs to Discounted Cash Flow

The inputs necessary for developing the discounted cash flows have been taken from the following schedules, which will be found in the appendices indicated below.

Schedules of Capital Cost Estimates, Appendix V

Schedules of Operating Expenses, Appendix VI

Schedule of Revenues, Appendix VII

Cash Inflow Schedules, Appendix VIII

The results of the DCF are tabulated in the following summary. The worksheets will be found in Appendix IX.

Summary of Discounted Cash Flow Rate of Return Analysis

Scenario	2	2	2	3	3	3	4	4	4	6	6	6
Condition	1	2	3	1	2	3	1	2	3	1	2	3
Capital Cost in ($ x 10^6) of:												
3,000 KW Condensing T.G. and Cooling Tower	-	-	-	1.26	1.26	1.26	1.26	1.26	1.26	-	-	-
4,500 KW Extracting/ Condensing T.G. and Cooling Tower	-	-	-	-	-	-	-	-	-	1.41	1.41	1.41
Pre-evaporator	-	-	-	-	-	-	0.23	0.23	0.23	0.23	0.23	0.23
Boiler Retrofit	-	-	-	-	-	-	-	-	-	2.1	2.1	2.1
Drying System	-	1.77	-	-	1.77	-	-	1.77	-	-	1.77	-
Drying/Pelletizing System	-	-	2.06	-	-	2.06	-	-	2.06	-	-	2.06
Total	-	1.77	2.06	1.26	3.03	3.32	1.49	3.26	3.55	3.74	5.51	5.80
Inflow ($ x 10^6)	-	4.64	4.62	1.04	5.01	5.16	2.08	6.55	6.71	5.04	9.68	9.67
Outflow ($ x 10^6)	-	1.77	2.06	1.26	3.03	3.32	1.49	3.26	3.55	3.74	5.51	5.80
% Rate of Return	-	22.85	18.3	.05	10.35	8.95	6.55	15.2	13.6	5.80	11.8	10.55
Payback Period in Years	-	3.81	4.44	12.08	6.05	6.42	7.16	4.98	5.29	7.42	5.69	6.0

Impact of Higher Fibre Content and Throughput

To test the impact of a higher fibre content on the ROI, Scenario 4 was expanded to include a Condition 4. It will be observed that when the fibre % cane increases for 12.5% to 13.5%, the ROI shows an improvement of 56%, all other conditions being equal.

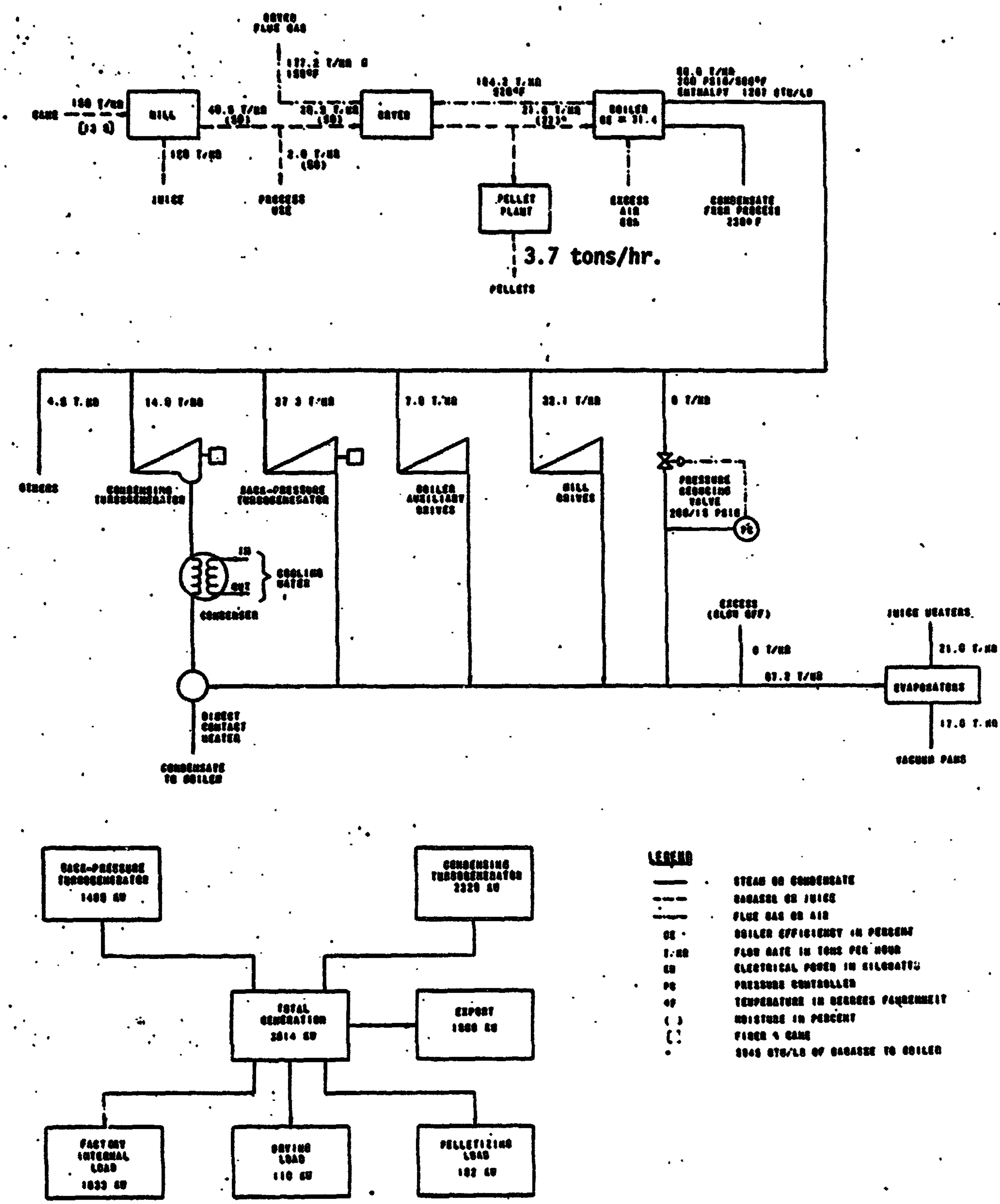
To test the impact of a higher grinding rate, Scenario 4 was further expanded to include Condition 5. In this case the fibre content is maintained at 12.5% but the grinding rate is increased to 200 TCH from 150 TCH. By coincidence the increase in ROI is nearly identical to that of Condition 4, in spite of a slight increase in the cost of the drying and pelletizing plant.

In testing the simultaneous effect of higher fibre content and grinding rate, (Scenario 4 Condition 6) the ROI was found to increase by 119%.

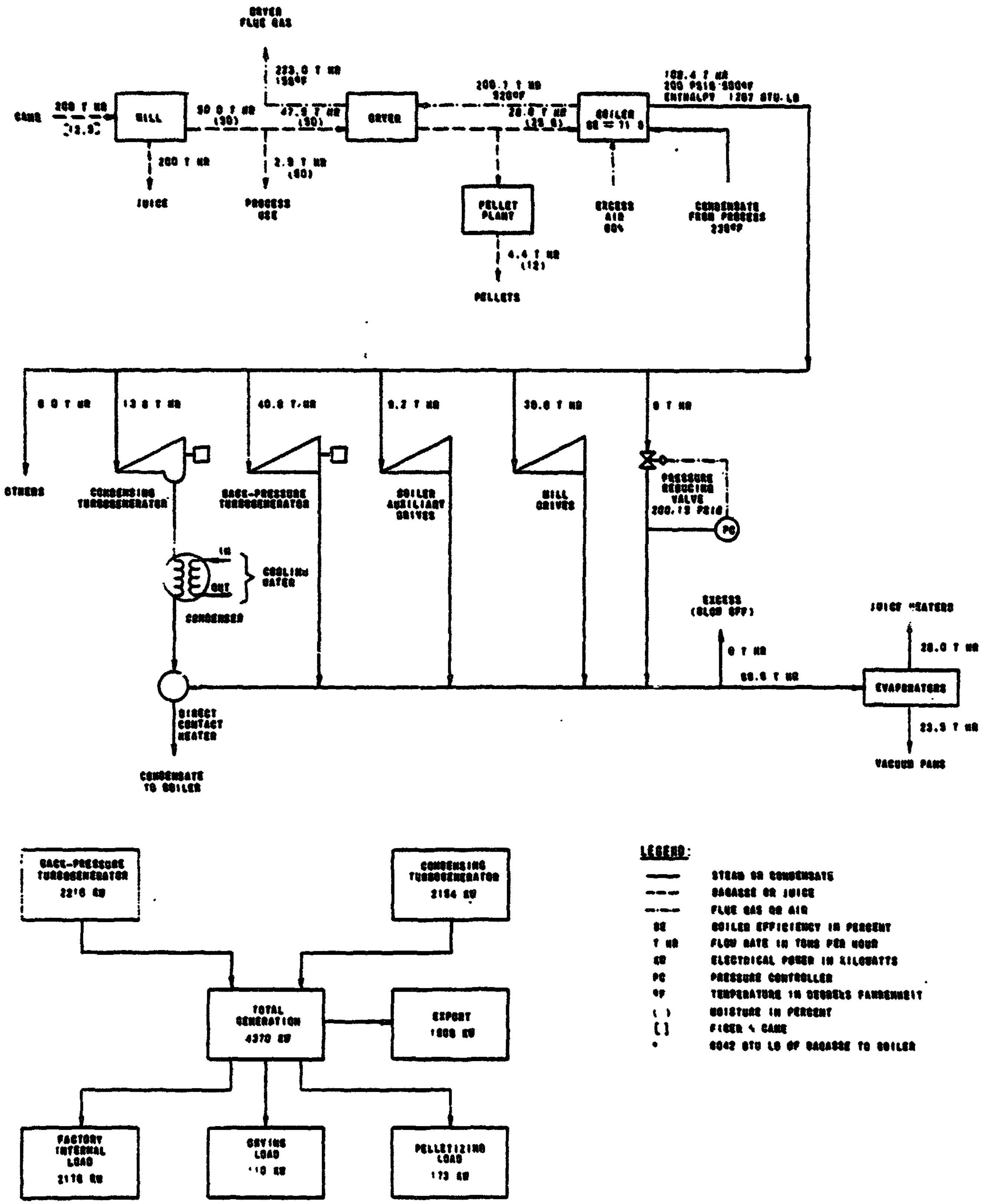
These examples indicate clearly that the higher the fibre content of the cane and the higher the grinding rate the more favorable are the results.

In Scenario 6 we wanted to test the impact of removing the cost of the boiler upgrading on the ROI. This is represented in the Summary of DCF chart as Scenario 6 Condition 3a with an ROI of 22.75% versus 10.55% where the cost of boiler upgrade is included (Scenario 6 Condition 3).

Finally we tested the impact of higher fibre % cane only on Scenario 6, and again the importance of high fibre was borne out in the results. Comparing Scenario 6 Condition 3 with Condition 4, the rate of return increases by 47%.



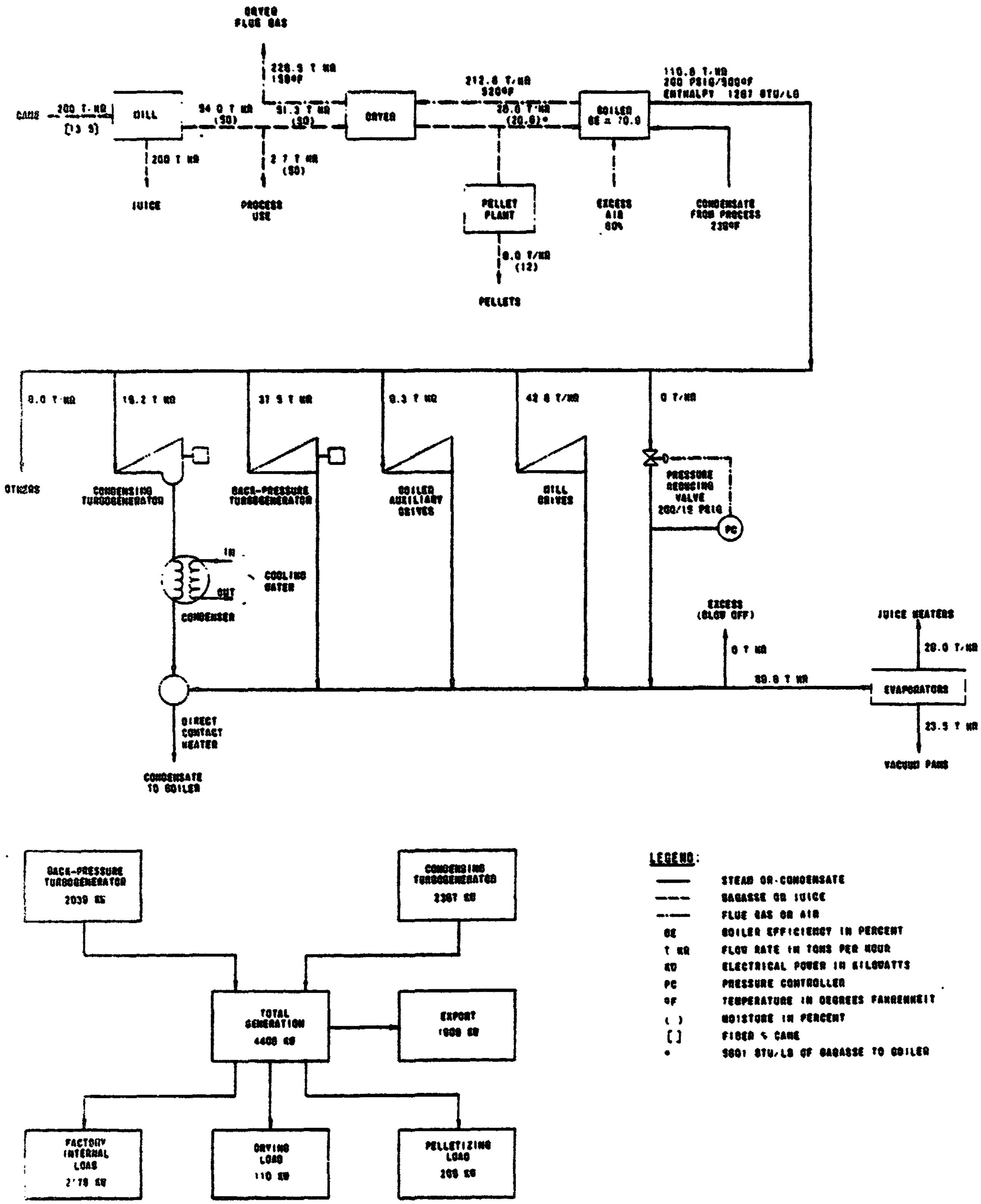
SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 4, CONDITION 4



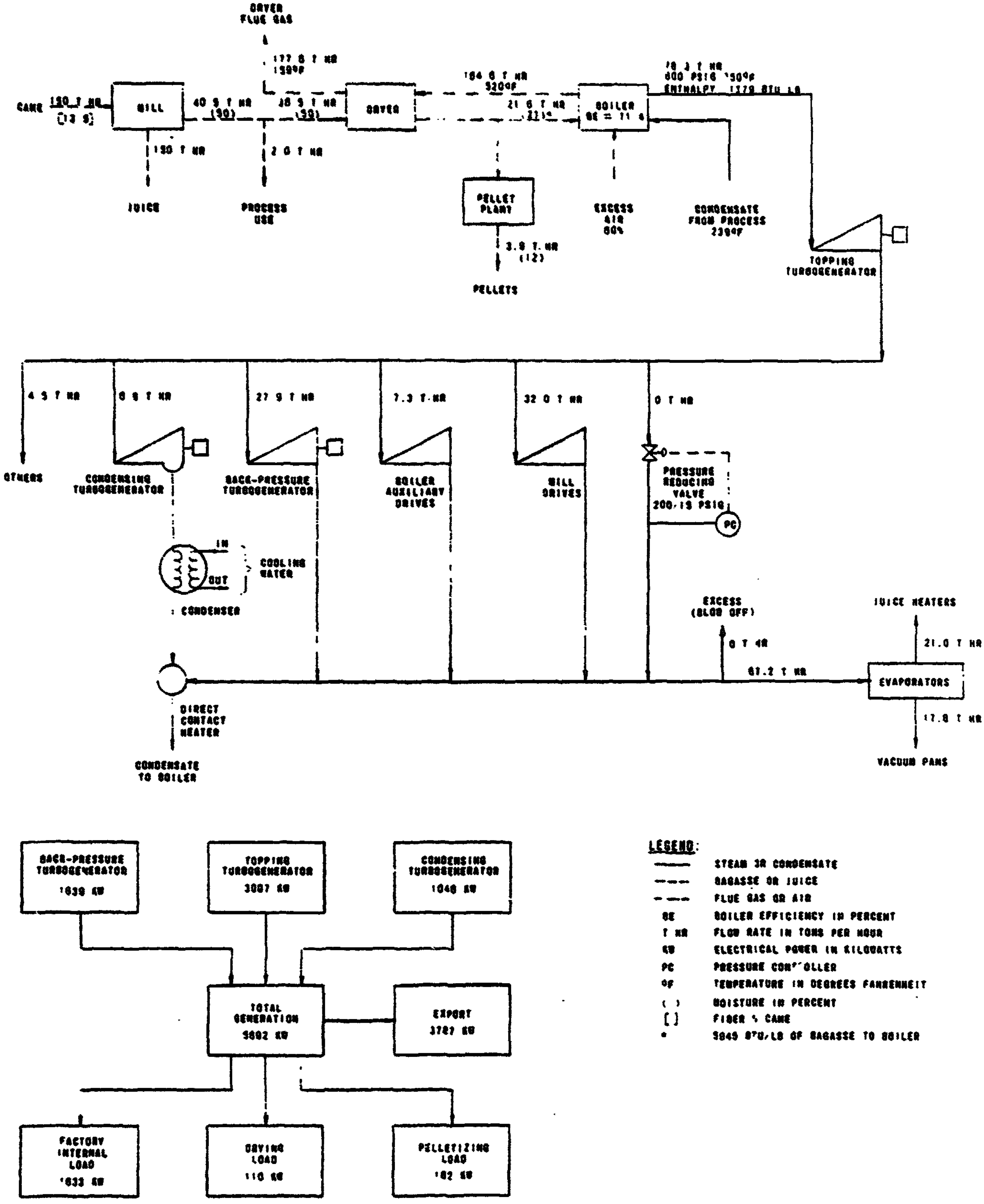
LEGEND:

———	STEAM OR CONDENSATE
– – –	BAGASSE OR JUICE
—·—·—	FLUE GAS OR AIR
BE	BOILER EFFICIENCY IN PERCENT
T HR	FLOW RATE IN TONS PER HOUR
KW	ELECTRICAL POWER IN KILOWATTS
PC	PRESSURE CONTROLLER
°F	TEMPERATURE IN DEGREES FAHRENHEIT
( )	MOISTURE IN PERCENT
[ ]	FIBER % CANE
*	8042 BTU LB OF BAGASSE TO BOILER

SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 4. CONDITION 5



SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 4. CONDITION 6



SIMULATION OF FACTORY THERMAL BALANCE
SCENARIO 6. CONDITION 4

SUMMARY OF COMPUTER SIMULATIONS

	SCENARIO 2				SCENARIO 4						SCENARIO 6
CONDITION:	1	1a	1b	1c	1	2	3	4	5	6	4
BOILER											
Steam..............1000 LBS/HR	162.9	184.9	228.3	246.6	171.2	171.2	171.2	173.3	218.8	221.5	156.6
LBS Steam/LBS Bagasse	2.4	2.4	2.4	2.4	2.4	4.3	4.2	4.0	4.1	3.9	3.6
Moisture (Feed Bagasse)......%	50.0	50.0	50.0	50.0	50.0	22.6	23.8	27.0	25.8	28.9	27.0
DRYER											
Dried Bagasse......1000 LBS/HR	-0-	-0-	-0-	-0-	-0-	46.0	45.9	51.1	62.4	69.2	51.1
Moisture (Dried Bagasse).....%	---	---	---	---	---	22.6	22.4	24.7	23.8	25.9	24.7
PELLETIZATION											
Pellets............1000 LBS/HR	-0-	-0-	-0-	-0-	-0-	-0-	5.4	7.8	8.8	12.1	7.9
Moisture....................%	---	---	---	---	---	---	12.0	12.0	12.0	12.0	12.0
EXCESS BAGASSE											
Bagasse............1000 LBS/HR	3.5	-0-	-0-	-0-	-0-	6.3	-0-	-0-	-0-	-0-	-0-
LBS Steam/LBS Bagasse	2.4	---	---	---	---	4.3	---	---	---	---	---
Moisture....................%	50.0	---	---	---	---	22.6	---	---	---	---	---
STEAM RATE											
Process..................LBS/TC	1026	1026	1026	1026	896	896	896	896	846	846	846
Other....................LBS/TC	60	60	60	60	60	60	60	60	60	60	60
STEAM REQUIREMENT											
Process............1000 LBS/HR	153.9	153.9	205.2	205.2	134.4	134.4	134.4	134.4	179.2	179.2	134.4
Excess (Blow Off)..1000 LBS/HR	-0-	22.0	11.1	29.4	-0-	-0-	-0-	-0-	-0-	-0-	-0-
Other..............1000 LBS/HR	9.0	9.0	12.0	12.0	9.0	9.0	9.0	9.0	12.0	12.0	9.0
Condensing TG......1000 LBS/HR	---	---	---	---	27.8	27.8	27.8	29.9	27.6	30.3	13.2
TOTAL......1000 LBS/HR	162.9	184.9	228.3	246.6	171.2	171.2	171.2	173.3	218.8	221.5	156.6
ELECTRICAL POWER											
Generation..................kW	2163	2594	3207	3500	3788	3788	3788	3814	4370	4406	5693
Consumption.................kW	1633	1633	2178	2178	1633	1743	1879	1905	2351	2497	1906
Export......................kW	530	961	1029	1322	2155	2045	1909	1909	1909	1909	3787
PSIG	200	200	200	200	200	200	200	200	200	200	600
TCH	150	150	200	200	150	150	150	150	200	200	150
Fibre % Cane	12.5	13.5	12.5	13.5	12.5	12.5	12.5	13.5	12.5	13.5	13.5

Summary of Discounted-Cash-Flow Rate of Return Analysis

Scenario	4	4	4	6	6
Condition	4	5	6	3a	4
Capital Cost in ($ x 10^6) of:					
3,000 KW Condensing T.G. and Cooling Tower	1.26	1.26	1.26	-	-
4,500 KW Extracting/ Condensing T.G. and C.T.	-	-	-	1.41	1.41
Pre-evaporator	0.23	0.23	0.23	0.23	0.23
Boiler Retrofit	-	-	-	-	2.1
Drying System	-	-	-	-	-
Drying/Pelletizing System	2.06	2.40	2.40	2.06	2.06
Total	3.55	3.89	3.89	3.70	5.80
Inflow ($ x 10^6)	8.83	9.72	12.56	9.67	11.81
Outflow ($ x 10^6)	3.55	3.89	3.89	3.70	5.80
% Rate of Return	21.25	21.30	29.85	22.75	15.55
Payback Period in Years	4.02	4.01	3.10	3.82	4.91

Note: Scenario 6 Condition 3a is identical to Scenario 6 Condition 3 with the exception that in 6/3a the cost of the boiler upgrading has not been included.

Overall Summary of Results

Scenario/ Condition	Improvement	ROI	Results KW Export	Tons Dry Bagasse	Tons Pellets
2/2	Drying	22.85	631	6.3	-
2/3	Drying, pelletizing	18.3	495	-	5.4
3/1	3,000 KW condensing T.G.	.05	1,164	-	-
3/2	3,000 KW T.G., drying	10.35	1,054	6.3	-
3/3	3,000 KW T.G., drying, pelletizing	6.42	918	-	5.4
4/1	3,000 KW T.G., pre-evaporator	7.16	2,155	-	-
4/2	3,000 KW T.G., pre-evaporator, drying	4.98	2,045	6.3	-
4/3	3,000 KW T.G., pre-evap., drying, pell'g.	5.29	1,909	-	5.4
6/1	4,500 KW condensing/extraction T.G., high pressure boiler, pre-evaporator	7.42	4,633	-	-
6/2	4,500 KW T.G., boiler, pre-evap., drying	5.69	3,923	6.3	-
6/3	4,500 KW T.G., boiler, pre-evap., drying, pelletizing	6.0	3,787	-	5.4
6/3a	Same as 6/3 but omit boiler upgrading	22.75	3,787	-	5.4
	Impact of Higher Fibre and Grinding Rate				
6/4	4,500 KW T.G., boiler, drying, pell'g., but fibre % cane at 13.5%	15.55	3,787	-	7.9
4/4	3,000 KW T.G., pre-evap., drying, pelletizing, 13.5% fibre in cane	21.25	1,909	-	7.8
4/5	3,000 KW T.G., pre-evap., drying, pelletizing, 200 TCH	21.30	1,909	-	8.8
4/6	3,000 KW T.G., pre-evap., drying, pelletizing, 200 TCH, 13.5% fibre in cane	29.85	1,909	-	12.1

SECTION IV

Conclusion

The burning of wet bagasse is inherently inefficient, leading to large losses of energy in the flue gases. By capturing a large proportion of this energy and using it for drying bagasse prior to burning, the efficiency of the combustion process is greatly improved, leading to a surplus of a high quality fibrous fuel which may be densified into a marketable commodity.

Our economic analysis of the drying and pelletizing process shows that a reasonable rate of return calculated by the discounted cash flow rate of return method can be achieved even in the case of marginal conditions of our example (low fibre % cane, low grinding rate, short grinding season).

If the goal is simply to produce pellets for sale to others, the investment may be confined to the acquisition of a drying and pelletizing plant and a storage facility, assuming the sugar factory has enough electrical generation capacity to carry the extra load required for the drying plant and pelletizing plant.

Alternatively, if the goal is to produce the maximum amount of electricity for sale to others, then it becomes necessary not only to dry the bagasse and pelletize the surplus but also to minimize the process steam need and maximize the generation of electrical power. These goals can be achieved by carrying out some or all of the improvements suggested in this study. The size of the investment in this case will depend on the number of favorable conditions already pre-existing in the factory.

In the case of our example, the basic factory is a bare minimum unit with a high process steam consumption and low electrical generation capacity. In other words it is really a "worst case" condition and consequently large investments are necessary to achieve optimum utilization of bagasse and steam. If we consider that the acceptable level of ROI is around 10%, then most of the scenarios show a marginally acceptable return until we increase either the grinding rate or the fibre content. All the scenarios, except Scenario 5, show a positive rate of return. The combination of higher grinding rate and higher fibre content produces a large increase in ROI, enough to satisfy the highest realistic expectations.

As was pointed out in the text of this study, if improvements such as replacement of boilers, acquisition of condensing/extracting turbo-generators, pre-evaporators, etc., are charged entirely against the production of additional power, ignoring the beneficial results they may have for the sugar operation, it is likely that the return on investment will be less attractive.

A more realistic approach would be to keep in mind what improvements are necessary for the production of surplus power in a factory and to work toward that goal by replacing equipment as it becomes due for replacement with the type and size of equipment that could eventually or gradually fit into the power program. The best example would be the case of boilers. If a low pressure boiler is due for replacement, it should be replaced by a 600 psig boiler even if in the beginning it has to be operated at a lower pressure.

The next step would be to install a topping turbine or better still replace the existing back-pressure turbo-generator with a double extraction/condensing turbo-generator. Once a condensing turbo-generator has been obtained, the next step could be to install a large pre-evaporator that could absorb all the exhaust from the prime movers and provide vapor for juice heaters, evaporators and vacuum pans. This would realize large steam savings with the result that more steam could be made available for power generation. Our study also shows that, of all improvements reviewed, the pre-evaporator shows the best return on investment.

The best time to plan, of course, is whilst new factories are being conceptualized. Under most normal conditions a large new factory could be made to produce several times more power than it needs for its own operation and it is hoped that in this era of energy awareness maximum advantage will be taken of such situations.

If a country in which a sugar industry is located depends on imported fossil fuel for power generation, substituting pelletized bagasse for fossil fuel may become an important factor in improving the balance of payment. In such a case, even if the ROI is small, as long as it is positive, there will be a gain for the country as a whole if import of fuel is eliminated.

Finally, in this study we have considered only pelletization as the means for densifying bagasse. Another form of densification is cubing and it is likely that this process will find some application in the sugar industry.

Cubing of bagasse is not yet well proven, though trials have been conducted at the Haina factory by Papakube Corporation of San Diego. These preliminary trials showed that bagasse can be cubed but to a lesser density than that of pellets. The handling of the cubes is not as easy as that of the pellets and of course the cubes require more storage capacity than pellets for a given weight. There have been a few problems that developed with the cubes in storage and at this time we are not prepared to recommend the use of cubing until further research has taken place and cubes are proven to be as stable as pellets under prolonged storage conditions.

Appendix I

Developing Countries Where Sugar Cane is Grown

Central America/Caribbean

Barbados
Belize
Costa Rica
Cuba
Dominican Republic
El Salvador
Gaudeloupe
Guatemala
Haiti
Honduras
Jamaica
Martinique
Mexico
Nicaragua
Panama
St. Kitts
Trinidad-Tobago

South America

Argentina
Bolivia
Brazil
Chile
Colombia
Ecuador
Guyana
Paraguay
Peru
Surinam
Uruguay
Venezuela

Asia

Bangladesh
Burma
China
India
Indonesia
Iran
Iraq
Malaysia
Pakistan
Philippines
Sri Lanka
South Vietnam
Thailand

Oceania

Fiji
Western Samoa

Africa

Angola
Cameroon
Chad
Congo
Egypt
Ethiopia
Gabon
Ghana
Guinea
Ivory Coast
Kenya
Malagasy Republic
Malawi
Mali
Mauritius
Morocco
Mozambique
Nigeria
Rwanda
Senegal
Somalia
South Africa
Sudan
Swaziland
Tanzania
Uganda
Upper Volta
Zaire
Zambia
Zimbabwe

Appendix II

Comparison Between Use of High Pressure and Low Pressure Steam and Between Back Pressure and Condensing Turbo-Generator

	Case I	Case II
Boiler pressure	200 psig	600 psig
Steam temperature	500^{0}F	750^{0}F
Bagasse burned per hour	25 tons (5×10^4 lb.)	25 tons (5×10^4 lb.)
Boiler efficiency	58%	58%
Steam turbine inlet pressure	200 psig	600 psig
Exhaust pressure	10 psig	10 psig
Use of steam	to T.G.	to T.G.
Higher Calorific Value of bagasse at 50% moisture	4,200 BTU	4,200 BTU
Heat value of bagasse	$5 \times 10^4 \times 4.2 \times 10^3$ $= 2.1 \times 10^8$ BTU	$5 \times 10^4 \times 4.2 \times 10^3$ $= 2.1 \times 10^8$ BTU
Heat transferred to steam in boiler	$0.58 \times 2.1 \times 10^8$ $= 1.218 \times 10^8$ BTU	$0.58 \times 2.1 \times 10^8$ $= 1.218 \times 10^8$ BTU
Temperature of feed water	190^{0}F	190^{0}F
Total heat value in 1 lb. of superheated steam	1,267.4 BTU	1,378.9 BTU
Heat in feed water	158 BTU	158 BTU
.˙. Net heat required to raise 1 lb. water to steam	1,109.4 BTU	1,220.9 BTU

Appendix II, page 2

	Case I	Case II
∴ Lb. of steam produced	$\frac{1.218 \times 10^8}{1.1094 \times 10^3}$	$\frac{1.218 \times 10^8}{1.2209 \times 10^3}$
	$= 1.0979 \times 10^5$	$= .99762 \times 10^5$
	= 109,790 lb.	= 99,762 lb.
Theoretical steam rate of turbine exhausting at 10 psig	19.27 lb./KWH	11.64 lb./KWH
Assumed thermodynamic efficiency of T.G.	.70	.72
Expected steam rate	27.52 lb./KWH	16.16 lb./KWH
∴ KW output	3,989 KW	6,173 KW

If instead of a back pressure unit a condensing turbo-generator is used, the temperature of the condensate would drop to 101°F but much more power would be generated, as can be observed from the following:

	Case I	Case II
Total heat value in 1 lb. of steam	1,267.4 BTU	1,378.9 BTU
Heat in feed water	69.1 BTU	69.1 BTU
∴ Net heat required to raise 1 lb. water to steam	1,198.3 BTU	1,309.8 BTU

Appendix II, page 3

	Case I	Case II
.·. Lb. of steam produced	$\frac{1.218 \times 10^8}{1.1983 \times 10^3}$	$\frac{1.218 \times 10^8}{1.3098 \times 10^3}$
	= 101,644 lb.	= 92,991 lb.
Theoretical steam rate of turbine exhausting at 2" Hg abs.	9.32 lb./KWH	7.09 lb./KWH
Assumed thermodynamic efficiency of T.G.	.72	.74
Expected steam rate	12.944 lb./KWH	9.58 lb./KWH
.·. KW output	7,853 KW	9,707 KW

Appendix III

Combination of Topping Turbo-Generator With Existing 200 psig Condensing Turbo-Generator

From the Mollier diagram the theoretical work by the steam expanding from 600 psig at 750°F to 200 psig is 119 BTU. Applying to this value a thermodynamic efficiency of .55, which is typical for a machine of this type and size, the actual work done = 65.45 BTU

$$= 65.45 \times 2.93 \times 10^{-4} \text{ KWH}$$

$$= 1.92 \times 10^{-2} \text{ KWH/lb. of steam}$$

$$\therefore \text{KW output} = 99{,}762 \times 1.92 \times 10^{-2}$$

$$= 1{,}915 \text{ KW}$$

Again referring to the Mollier diagram it will be found that the steam temperature of the exhaust at 200 psig is 550°F. This steam will have to be desuperheated to 500°F before being admitted to the 200 psig condensing unit.

Total heat in steam at 200 psig and 550°F = 1,294.6 BTU

Total heat in steam at 200 psig and 500°F = 1,267.4

Difference = 27.2 BTU/lb.

$$\therefore \text{Heat to be removed from exhaust steam} = 99{,}762 \times 27.2$$

$$= 2.713 \times 10^{6} \text{ BTU}$$

Water used for desuperheating is at 101°F (condensate).

Let X = lb. of water at 101°F required.

Appendix III, page 2

Saturated temperature of steam at 200 psig = 388°F.

.'. Heat required to raise temperature of desuperheating water

to 388°F = X(388 - 101)

= 287X

L.H. of vaporization at 200 psig = 837.4 BTU/lb.

.'. Heat required to vaporize X lb. = 837.4X BTU

Hence, equating heat required to raise temperature of X lb. of water + heat required to vaporize X lb. of water with heat to be removed from superheated exhaust steam, we have:

$$287X + 837.4X = 2.713 \times 10^6$$

$$1,124.4X = 2.713 \times 10^6$$

$$X = \frac{2.713 \times 10^6}{1.1244 \times 10^3}$$

$$= 2.413 \times 10^3$$

$$= 2,413 \text{ lb./hr.}$$

.'. Weight of desuperheated exhaust steam available to 200 psig

unit = 99,762 + 2,413

= 102,175 lb./hr.

Steam rate of 200 psig turbo-generator exhausting at 2" Hg abs.

= 12.94 lb./KWH.

.'. KW output = $\frac{102,175}{12.94}$ = 7,896 KW, which added to the 1,915 KW obtained from the topping unit would total 9,811 KW.

Appendix IV

Boiling House Steam Requirement

To evaluate the economic impact of various types of improvements to enhance the effective use of mill bagasse, a basic factory is defined in the following to establish a base against which these improvements will be assessed. The parameters of the basic factory will be grouped under three headings:

1) Operating conditions;
2) Material balance;
3) Thermal balance.

Operating Conditions

Grinding season: 6 months/year

5 days/week

22 hours of operation/day

2 hours of idle time/day where idle time = breakdown and/or shortage of cane

2,640 operating hours/season

Rate: 150 tons cane/hour

Cane quality: Bagasse % cane = 25%

Fibre % cane = 12.5%

Normal juice % cane = 75%

Bagasse quality: Moisture % bagasse = 50%

Appendix IV, page 2

Imbibition water: Imbibition % fibre = 200% expressed as a % of fibre instead of cane, as it is the amount of fibre that determines the quantity of water that can be absorbed.

Material Balance

Juice extraction: This process may be represented as follows:

Cane + imbibition water = raw juice + bagasse.

From the operating conditions, the following are calculated:

$$\text{Tons of imbibition water/hour} = \frac{12.5}{100} \times 150 \times \frac{200}{100} = 37.5$$

$$\text{Tons of bagasse/hour} = 150 \times \frac{25}{100} = 37.5$$

∴ Therefore, from the above equation, the amount of raw juice is determined to be 150 tons.

Clarified juice and syrup:

The raw juice from the extraction process is treated with lime and brought to boiling prior to clarification. From the clarification process, the clarified juice is usually found to be somewhat higher in quantity than the raw juice, mainly due to the water introduced in the washing of impurities from the raw juice at the filter station. This increase in quantity is found in practice to be on the order of 10% of the raw juice. Thus, from 150 tons of raw juice, the amount of clarified juice is calaculated as 165 tons.

Appendix IV, page 3

If the juice is assumed to contain 12.5% total solids, and the syrup 65% total solids (65^{0} brix), then the quantity of water evaporated from the clarified juice to syrup is found to be:

$$165 \times \frac{65 - 12.5}{65} = 133.3 \text{ tons/hr., leaving } 31.7 \text{ tons/hr. of syrup.}$$

Raw sugar and final molasses:

The total dissolved solids, namely sucrose, reducing sugars, ash impurities, etc., contained in the clarified juice will travel throughout the whole process to provide finally two commercial products: raw sugar and final molasses.

Tons dissolved solids in clarified juice and syrup

$$= 165 \times \frac{12.5}{100}$$

$$= 20.625$$

Assuming a syrup purity ranging between 80^{0} and 84^{0} and raw sugar at 96^{0} pol, the following distribution is obtained:

	Syrup 80^{0} Purity	Raw Sugar	Final Molasses
Sucrose	16.5	15.9	0.6
Other Solids	4.125	0.45	3.675
Process Water Added		0.15	0.75
Total		16.50	5.025

Appendix IV, page 4

	Syrup 84° Purity	Raw Sugar	Final Molasses
Sucrose	17.325	16.725	0.6
Other Solids	3.3	0.45	2.85
Process Water Added		0.15	0.6
Total		17.325	4.05

Though the molasses composition represented above is only an approximation, the amount of molasses approximates closely to what is found in practice.

We see from the above, therefore, that from 150 tons of cane approximately 17 tons of sugar and 4 to 5 tons of molasses are produced. With the higher purity, more sugar and less molasses are obtained.

Massecuite:

Further evaporation of the syrup leads to crystallization into massecuites, which in turn are treated in centrifugals where the sugar crystals are separated from the mother liquor.

A variety of massecuite boiling systems are practiced but in this context only two simple methods will be dealt with to determine the quantities of massecuites and molasses which have to be handled (are produced.) Without considering the distribution of sucrose and non-sucrose (i.e. the purity ratios of the intermediate products), this analysis will be based on the distribution of total solids previously established under the heading "Raw Sugar and Final Molasses."

From the flow diagram (Figure 1), the total quantities of massecuite on a solids basis are: 2-boiling system 33.45;
3-boiling system 36.975.

It is apparent from the foregoing that with the 3-boiling system about 10% more massecuite has to be processed, as well as more molasses, with a correspondingly higher steam consumption.

The amount of massecuite to be boiled may be assumed to be typical, although considerable variation is noted in juice quality. With variation in juice purity, the boiling scheme may therefore be modified and such modifications will not alter the above analysis.

It should be noted that the total solids in massecuites are approximately 60% to 80% higher than the amount present in the incoming syrup.

Thermal Balance

Juice Heating:

One pound of steam, having a latent heat of approximately 960 BTU, will raise the temperature of 1,000 lb. of juice by 1°F, when the specific heat of the juice is approximately 0.96; i.e. 1 ton of

Appendix IV, page 6

steam (or vapor) will raise the temperature of 100 tons of juice by 10^{0}F. Therefore, for heating 150 tons of raw juice from 80^{0}F to the temperature required for clarification, 220^{0}F, the steam flow required is 21 tons.

Further, due to the long retention time in the clarifier, the clarified juice will reach the evaporator at about 200^{0}F and will have to be reheated to its boiling point of 230^{0}F under 6 psig pressure before evaporation can commence. The preheating by 30^{0}F, whether done in the evaporator itself or in a juice preheater (the preferred method), will require about 5.0 tons of steam for 165 tons of clarified juice.

Evaporation:

From the steam table, 1 lb. of steam upon condensation will give up 950 BTU, which is not quite sufficient to produce 1 lb. of first vapor, for which 958 BTU is required. So in each evaporation vessel 1 lb. of vapor appears to evaporate less than 1 lb. of water. However, if the flash evaporation from the juice, as it passes from one vessel to the subsequent one (from a higher to a lower temperature), is taken into account, one may assume that 1 lb. of steam entering the calandria will produce 1 lb. of vapor. Therefore, 1 ton of steam (vapor) will evaporate 1 ton of water.

Appendix IV, page 7

As the basic factory is fitted with a straight quadruple effect evaporator, i.e. without bleeding of vapor, each vessel will, according to ton per ton evaporation, evaporate as much as the others, or one fourth of the total. Under the heading "Material Balance" it was established that the total evaporation is 133.3 tons; therefore the first vessel will receive $\frac{133.3}{4}$ = 33.325 tons of exhaust steam.

Boiling of Massecuites:

In this presentation, the steam requirement of a 3-boiling system will be considered.

As determined under the heading "Material Balance," the total quantity of massecuites, on a solid basis, is 37 tons but, as all massecuites are boiled to a water content of 10%, the actual quantity of massecuites is 41.1 tons.

Further it is assumed that, for re-boiling, the molasses will be diluted and the low grade sugar will be dissolved to the same concentration as the incoming syrup (65^0 brix), so that the total solid will be $\frac{37}{0.65}$ = 57 tons of liquor. Therefore the evaporation will be 57 - 41.1 = 15.9 tons of water.

As evaporation is performed under vacuum with steam at a pressure higher than that of the vacuum vapor, it will be inaccurate to

Appendix IV, page 8

assume the previously determined evaporation of "ton per ton." From the steam table it is noted that the steam requirement will be 10% higher than the amount of water evaporated, i.e. 17.5 tons.

Total Process Steam Requirement:

From the foregoing, the following summary is established and is shown in Figure 2:

(1)	Hea... of limed juice (80^oF to 220^oF)	21.125 tons
(2)	Heating of clarified juice (200^oF to 230^oF)	5.0 tons
(3)	Juice evaporation	33.325 tons
(4)	Boiling of massecuites	17.5 tons
	Total	76.95 tons

In summary, the process steam requirement can be expressed as 1,026 lb./ton of cane.

Improvement of the Thermal Balance of the Basic Factory

The process steam requirement of 1,026 lb./ton cane, when utilizing a straight quadruple effect evaporation station without vapor bleeding, can be considerably improved by using the first vapor for juice heating and massecuite boiling.

The first vapor bleed will result in the evaporation of 1(17.5 + 21) = 38.5 tons of water in single effect. Therefore 133.3 - 38.5 = 94.8 tons

Appendix IV, page 9

of water are left to be evaporated in quadruple effect, i.e. $\frac{94.8}{4}$ = 23.7 tons for each vessel. Thus the first body has to evaporate 38.5 + 23.7 = 62.2 tons and the last three vessels 23.7 tons each, providing a total evaporation of 133.3 tons. The exhaust steam to the first vessel is 62.2 + 5 tons = 67.2 tons (for preheating) as shown in Figure 3. This arrangement results in a saving of 12.7% on exhaust steam or a requirement of 896 lb. of exhaust/ton cane.

Further improvement is possible if the limed juice heating is done in two stages, with first vapor and second vapor. (It is not possible to reach the final temperature of 220^{o}F with second vapor alone. The steam saving is as shown in the following.

The first vapor bleed will evaporate in single effect 1(17.5 + 6) = 23.5 tons and the second vapor bleed will evaporate in double effect 2 x 15 = 30 tons, thus leaving 133.3 - (23.5 - 30) = 79.8 tons to be evaporated in quadruple effect, i.e. $\frac{79.8}{4}$ = 19.95 tons per vessel. Under this condition the first vessel has a total evaporation of 17.5 + 6 + 15 + 19.95 = 58.45 tons. The exhaust steam needed by the first vessel is 58.45 tons + 5 tons for preheating of the clarified juice. This would result in a net saving of 17.5% or a requirement of 846 lb. of exhaust steam/ton cane.

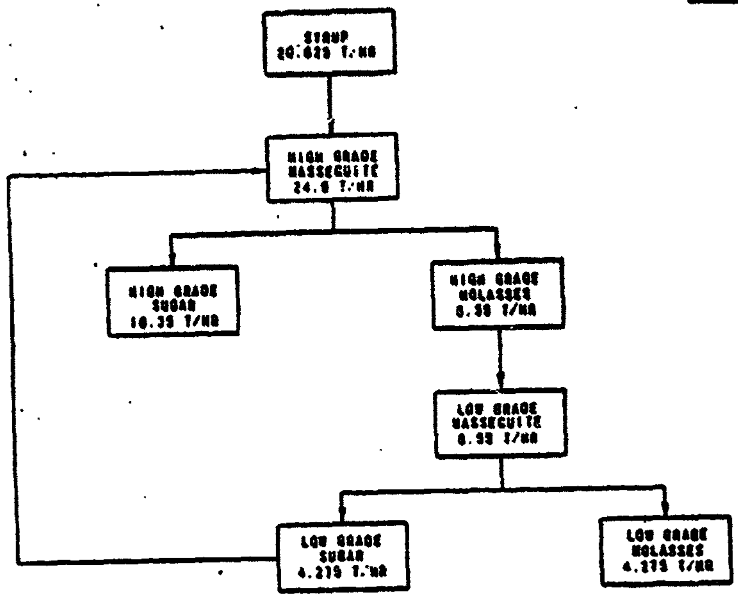
In summary, the steam consumption under the following conditions is:

Figure 2	Without bleeding	76.95 tons or 1,026 lb./ton cane
Figure 3	With bleeding of first vapor	67.2 tons or 896 lb./ton cane
Figure 4	With bleeding of first and second vapor	63.45 tons or 846 lb./ton cane

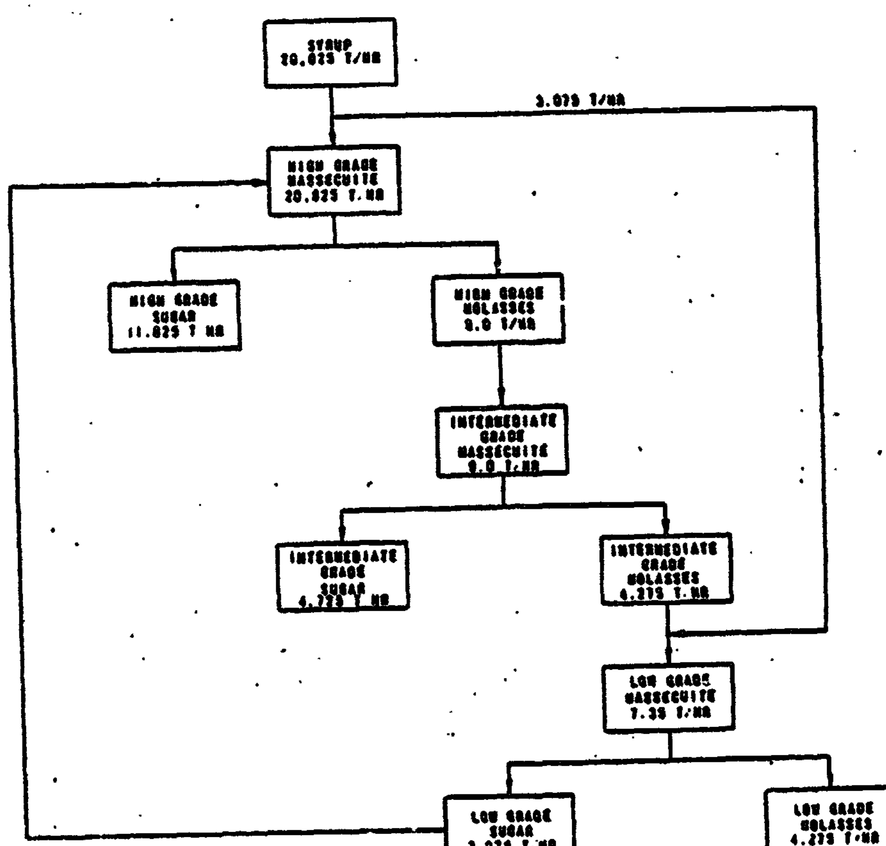
The straight quadruple effect of the basic factory is taken as having a performance of 4.5 lb. of water evaporated per sq.ft. of heating surface, i.e. 4 vessels of 8,000 sq.ft. each. With first vapor bleeding, the first vessel will have to evaporate 62.2 tons of water and therefore needs to have a heating surface of approximately 15,000 sq.ft. or an additional vessel of 8,000 sq.ft. in parallel with the existing one.

LEGEND:
T/HR FLOW RATE IN TONS SOLIDS PER HOUR

Appendix IV, page 10a

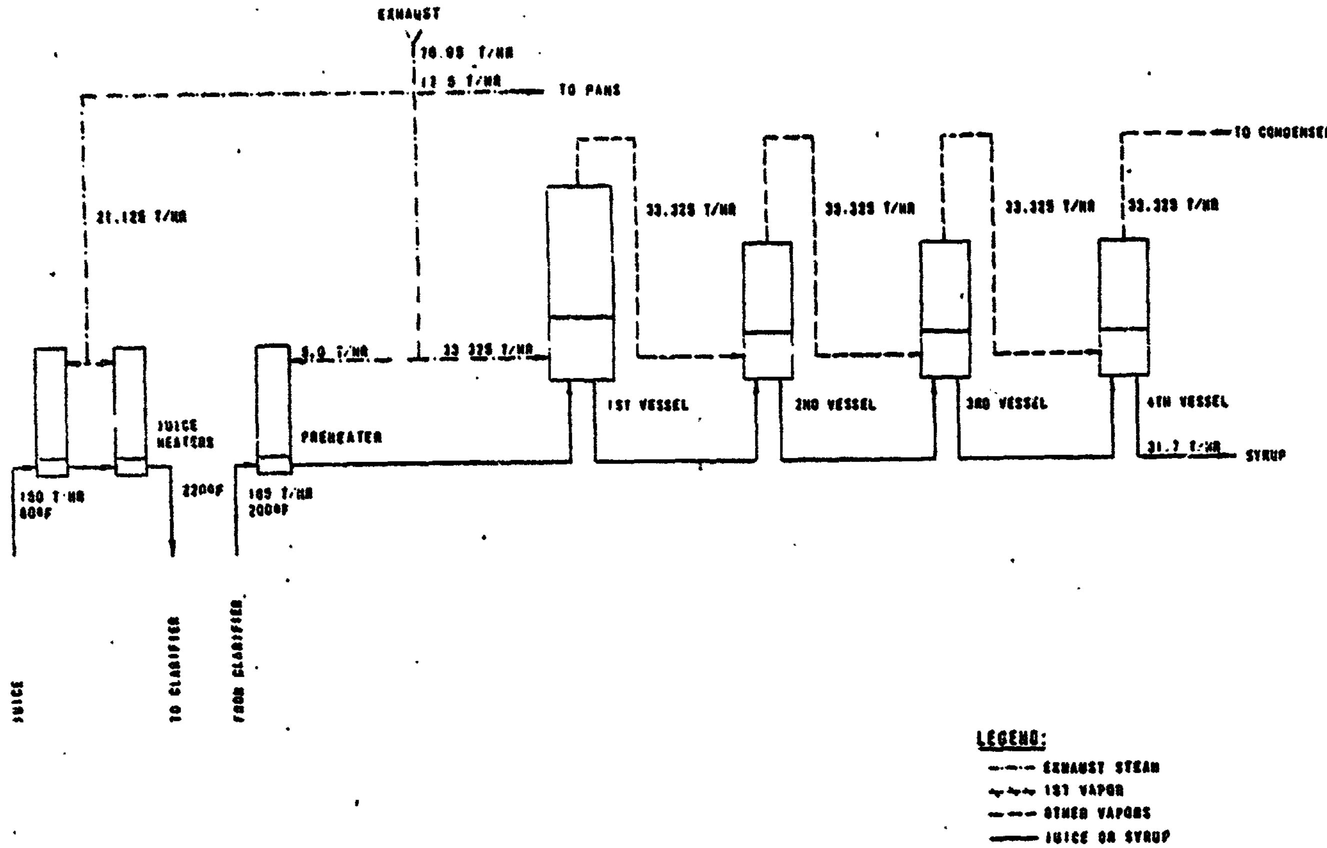


TWO-BOILING SYSTEM



THREE-BOILING SYSTEM

FIGURE 1
MASSECUITE BOILING SCHEME



TOTAL EVAPORATION

33.325 x 4 = 133.3 T/HR

RATE

150 TONS CANE PER HOUR
150 TONS RAW JUICE PER HOUR
165 TONS CLARIFIED JUICE PER HOUR

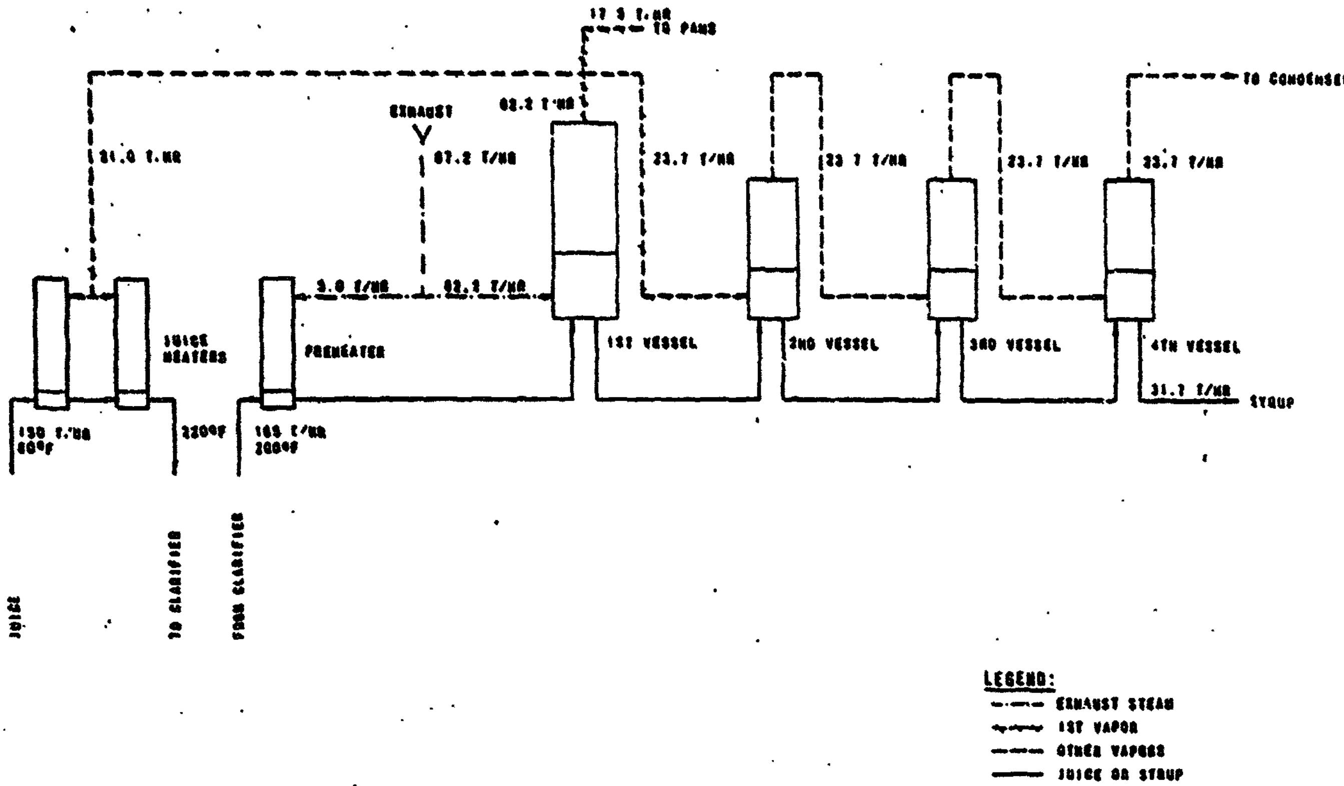
DISTRIBUTION

WITHOUT BLEEDING OF 1ST VAPOR
EXHAUST STEAM TO JUICE HEATERS, EVAPORATORS AND PANS

EQUIPMENT

STRAIGHT QUADRUPLE EFFECT OF EQUAL HEATING SURFACES.

FIGURE 2
PROCESS STEAM DISTRIBUTION
FOR HEATING, EVAPORATION AND CRYSTALLIZATION



TOTAL EVAPORATION

62.2 + 23.7 + 23.7 + 23.7 = 133.3 T/HR

RATE

150 TONS CANE PER HOUR

150 TONS RAW JUICE PER HOUR

165 TONS CLARIFIED JUICE PER HOUR

DISTRIBUTION

EXHAUST STEAM TO LARGE 1ST VESSEL PRE-EVAPORATOR

BLEEDING OF 1ST VAPOR FROM 1ST VESSEL TO JUICE HEATERS

BLEEDING OF 1ST VAPOR FROM 1ST VESSEL TO PANS

EQUIPMENT

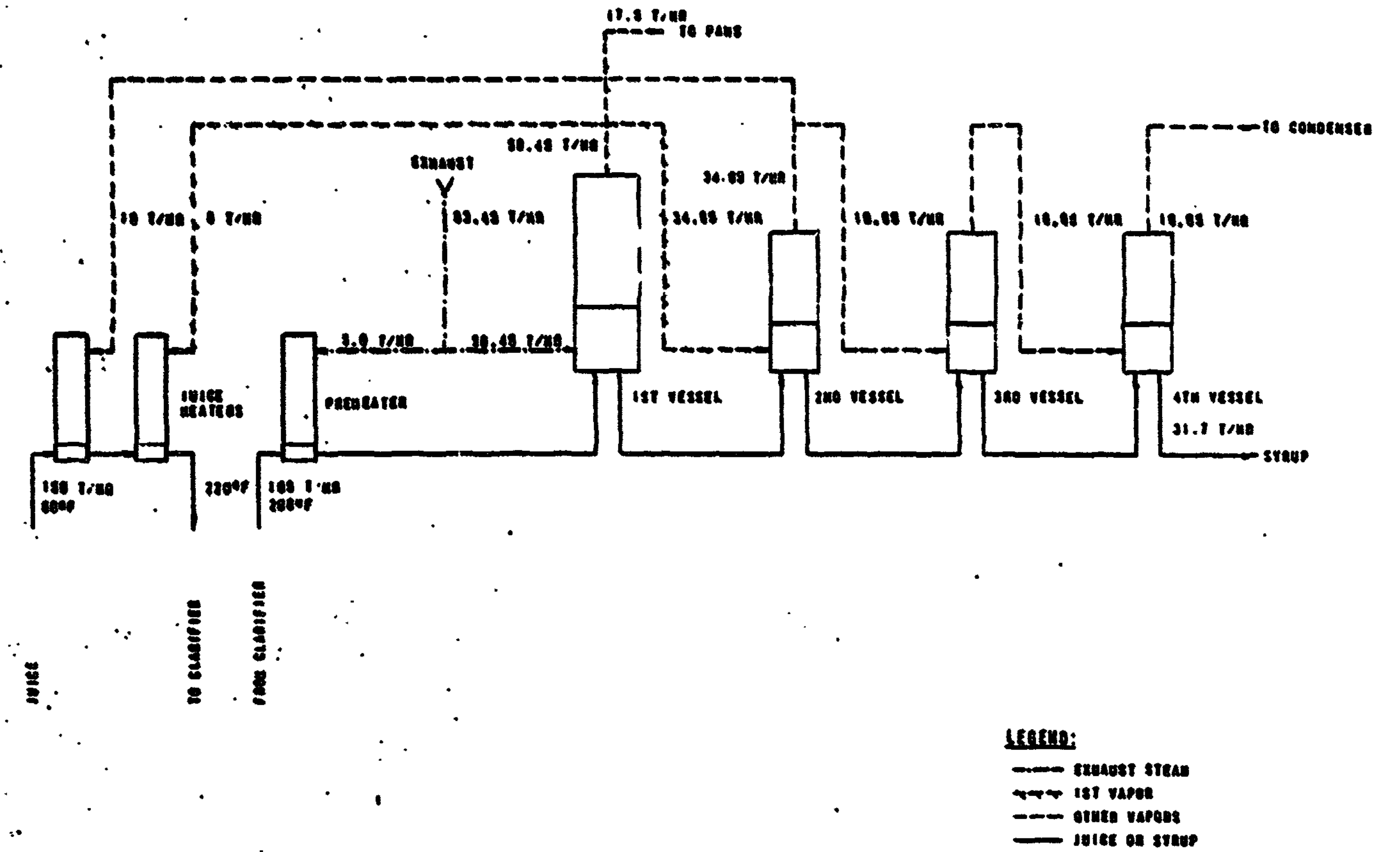
LARGE 1ST VESSEL

2ND, 3RD AND 4TH VESSELS OF EQUAL HEATING SURFACES.

FIGURE 3

PROCESS STEAM DISTRIBUTION

FOR HEATING, EVAPORATION AND CRYSTALLIZATION



TOTAL EVAPORATION:

58.45 + 34.95 + 19.95 + 19.95 = 133.3 T/HR

RATE

150 TONS CANE PER HOUR

150 TONS RAW JUICE PER HOUR

165 TONS CLARIFIED JUICE PER HOUR

DISTRIBUTION

EXHAUST STEAM TO LARGE 1ST VESSEL PRE-EVAPORATOR

BLEEDING OF 1ST VAPOR FROM 1ST VESSEL TO FINAL JUICE HEATING

BLEEDING OF 1ST VAPOR FROM 1ST VESSEL TO PANS

BLEEDING OF 2ND VAPOR FROM 2ND VESSEL TO PRIMARY JUICE HEATING

EQUIPMENT

LARGE 1ST VESSEL

2ND, 3RD AND 4TH VESSELS OF EQUAL HEATING SURFACES.

FIGURE 4
PROCESS STEAM DISTRIBUTION
FOR HEATING, EVAPORATION AND CRYSTALLIZATION

Appendix V

Schedule of Capital Cost Estimates

System: Dryer and Pelletizer

For drying capacity of up to 45 tons/hr. of 50% moisture bagasse with or without pelletizing capacity of up to 5 tons/hr. of 12% moisture pellets.

	Drying Only (000's)		Drying and Pelletizing (000's)	
Cost of major items:				
Mechanical	$ 633		$ 732	
Electrical	60		76	
Instrumentation	75		84	
Structural	92		106	
Others (lagging material, paint, drives, etc.)	36		63	
Total		$ 896		$ 1,061
Installation cost of major items:				
Mechanical	190		217	
Electrical	34		42	
Instrumentation	16		18	
Structural	42		49	
Others	29		51	
Total		311		377
Cost of manufacturing building:				
Material for civil work	27		27	
Bldg. structure & material	36		36	
Installation	31		31	
Total		94		94
Shipping and Insurance		55		62
Total erected cost		1,356		1,594
Engineering:				
Basic	75		80	
Mechanical	26		27	
Electrical	8		9	
Instrumentation	13		13	
Structure	18		19	
Civil	10		10	
Total		150		158
Main contractor (contract mgt.)		205		239
Royalties		41		48
Start-up and commissioning		12		16
Total of project		$ 1,764		$ 2,055

Appendix V, page 2

Schedule of Capital Cost Estimates

System: Dryer and Pelletizer

For drying capacity of 45+ to 65 tons/hr. of 50% moisture bagasse with or without pelletizing capacity of 5+ to 10 tons/hr. of 12% moisture pellets.

	Drying Only (000's)		Drying and Pelletizing (000's)	
Cost of major items:				
Mechanical	$ 728		$ 926	
Electrical	68		82	
Instrumentation	75		84	
Structural	106		122	
Others (lagging material, paint, drives, etc.)	40		70	
Total		$ 1,017		$ 1,284
Installation cost of major items:				
Mechanical	219		250	
Electrical	38		47	
Instrumentation	16		18	
Structural	48		56	
Others	32		56	
Total		353		427
Cost of manufacturing building:				
Material for civil work	30		30	
Bldg. structure & material	39		39	
Installation	33		33	
Total		102		102
Shipping and insurance		62		70
Total erected cost		1,534		1,883
Engineering:				
Basic	75		80	
Mechanical	30		32	
Electrical	8		10	
Instrumentation	13		13	
Structural	20		22	
Civil	10		10	
Total		156		167
Main contractor (contract mgt.)		230		282
Royalties		46		56
Start-up and commissioning		12		16
Total of project		$ 1,978		$ 2,404

Appendix V, page 3

Schedule of Capital Cost Estimates

System: Turbo-Generator and Auxiliaries

1) Turbo-Generator

Type	Condensing	Topping	Extraction Condensing
Rated capacity	3,000 KW	3,500 KW	4,500 KW
Inlet steam condition	200 psig/ 500°F	600 psig/ 750°F	600 psig/ 750°F
Extraction steam cond'n.	None	None	200 psig
Exhaust steam condition	2" Hg abs.	200 psig	2" Hg abs.
Cost:			
Complete unit delivered to site	$ 600,000	$ 510,000	$ 690,000
Other material:			
Mechanical (incl. piping, valving, etc.)	90,000	77,000	104,000
Electrical	42,000	36,000	48,000
Instrumentation	21,000	18,000	24,000
Misc. (lagging, painting, etc.)	15,000	13,000	17,000
Civil (mat. & labor)	60,000	60,000	69,000
Building	20,000	20,000	20,000
Installation (mech., elec., inst., etc.)	150,000	128,000	173,000
Engineering	30,000	26,000	35,000
Total cost	$ 1,028,000	$ 888,000	$ 1,180,000

2) Cooling Tower

Similar to Marley Cross Flow Class 500			
Number of cells	2	None	2
Cost:			
Complete unit delivered to site	$ 120,000		$ 120,000
Other material			
Mechanical	18,000		18,000
Electrical	8,000		8,000
Instrumentation	4,000		4,000
Misc. (as above)	3,000		3,000
Civil (mat. & labor)	40,000		40,000
Installation	30,000		30,000
Engineering	6,000		6,000
Total Cost	$ 229,000		$ 229,000
Total (1) & (2)	$ 1,257,000	$ 888,000	$ 1,409,000
Approximately	$ 1,260,000	$ 890,000	$ 1,410,000

Appendix V, page 4

Schedule of Capital Cost Estimates

System: Condensing Turbo-Generator and Cooling Tower

For steam condition of 200 psig / 500°F.

1) Turbo-Generator of 3,000 KW Condensing Type

Multi-valve, multi-stage, horizontal, reaction, axial flow and condensing type turbo-generator rated at 3,000 KW at steam condition of 250 psig / 500°F and exhausting at 2" Hg abs. to a main surface condenser. Turbine is fitted with single helical and single reduction gear for a rated speed of 9400/1800 RPM. The turbine is to produce 3,000 KW while passing a throttle flow of 38,000 lb./hr. at rated condition of 0.80 lagging power factor. This corresponds to a steam flow of 12.64 lb./KWH, or an overall efficiency of 75%. The turbine-generator set is capable of producing 3,750 KW at 1.0 power factor when the steam flow is increased to 46,500 lb./hr. for a steam rate of 12.39 lb./KWH.

The generator is a revolving field, cylindrical poles, brushless type synchronous generator, totally enclosed with closed air circulation self-ventilated with water cooled air cooler.

High tension voltage:	AC 4,160 V, 60 Hz, 3-phase, 3-wire, neutral earthed through resistor
Low tension voltage:	AC 480 V, 60 Hz, 3-phase, 3-wire
Control voltage:	DC 125 V

The turbine generating set shall be capable of continuous operation in parallel with other generators and with the public utility system. Also included is a molded case, three-pole, single throw, manual operation type circuit breaker.

The switchboard consists of: Generator panel
Exciter panel
Surge absorber panel
Neutral panel
Synchronizing panel

All above delivered to site	$ 600,000
Other material:	
Mechanical (incl. piping, valving, etc.)	90,000
Electrical	42,000
Instrumentation	21,000
Misc. (lagging, painting, etc.)	15,000
Civil work	60,000
Building	20,000
Installation (mech., elec., inst., etc.)	150,000
Engineering	30,000
Total	$ 1,028,000

2) Cooling Tower

Two-cell cooling tower of the Marley Cross Flow Class 500 type, delivered to site	$ 120,000
Other material:	
Mechanical (incl. piping, valving, etc.)	18,000
Electrical	8,000
Instrumentation	4,000
Misc.	3,000
Civil work	40,000
Installation (mech., elec., inst., etc.)	30,000
Engineering	6,000
Total	$ 229,000

Total of turbo-generator and cooling tower system	$ 1,257,000
Approximately	$ 1,260,000

Appendix V, page 6

Schedule of Capital Cost Estimates

System: Pre-evaporator

Major components:	
8,000 sq.ft. heating surface vessel w/ tubes	$ 80,000
Entrainment separator	8,000
Valving	12,000
Piping	24,000
Insulation	14,000
Structural	18,000
Instrumentation	6,000
Other misc. (electrical, etc.)	4,000
Total	166,000
Civil work (material and labor)	18,000
Installation	40,000
Engineering	8,000
Total of project	$ 232,000

Appendix VI

Schedule of Operating Expenses

System: Dryer of 45 tons/hr.

Depreciation

Capital cost of plant erected	$ 1,770,000
Depreciation based on 15 year (straight line)	$ 118,000/year

Maintenance

The cost of maintenance materials is expected to be no more than 1.5% of the major equipment cost, i.e. .015 x 896,000 = $13,440 approximate per year. This should cover both material and labor, which has proven to be the case at Haina.

Labor

One operator with a guaranteed 2,080 hr./year at an average of $6.50/hr. including benefits, i.e. $13,520/year.

Operating Materials and Utilities

Electrical load:

Base: Grinding season - 6 months/year
5 days/week
22 hours of operation/day
2 hours of idle time/day
Full load condition during an operating hour = 110 KW
Partial load condition during an idle hour = 30 KW
$0.06/KWH

Appendix VI, page 2

Off-season - 6 months/year
5 days/week
8 hours at partial load/day

Electrical energy cost:

During grinding season -

2,640(110 x 0.06) + 240(30 x 0.06) =	\$ 17,856
During off-season -	
120 x 8,730 x 0.06 =	1,728
Total	\$ 19,584
Other utilities: Water = 6 x 50 =	300
Other material: Lubrication & others \$750/mo.	4,500
Total operating materials and utilities	\$ 24,384

Operating Cost Summary

Depreciation	\$ 118,000
Maintenance	13,440
Labor	13,520
Operating materials and utilities	24,384
Total	\$ 169,344

Appendix VI, page 3

For the return on investment computation, the cost of electrical energy used during the grinding season will not be taken into consideration, as the system has been adjusted to produce its energy requirement, and further the input for the DCF computation is expressed as incremental cost associated with the implementation of the proposed system.

Input for DCF computation:

Depreciation	$ 118,000
Maintenance	13,440
Labor	13,520
Operating materials and utilities	6,470
Total	$ 151,430

Appendix VI, page 4

Schedule of Operating Expenses

System: Dryer of 45 tons/hr. and pelletizer of 5 tons/hour

Depreciation

Capital cost of plant erected	$ 2,055,000
Depreciation based on 15 year (straight line)	$ 137,000/year

Maintenance

The cost of maintenance materials is expected to be no more than 1.5% of the major equipment cost, i.e. .015 x 1,061,000 = $15,915 approximate per year. The latter should cover both material and labor, which has proven to be the case at Haina.

Labor

Two operators with a guaranteed 2,080 hr./year at an average of $6.50/hr. including benefits, i.e. $27,040/year.

Operating Materials and Utilities

Electrical load:

Base: Same as for drying

Plus energy used for pelletization system -

$\left[76 + (22 \times \text{tons pellets/hr.})\right]$ KWH, i.e.

$19{,}584 + \left[76 + (22 \times \text{tons pellets/hr.})\right] (0.06)(2{,}640)$

Appendix VI, page 5

Other utilities:	Water = 6 x 50	$ 300
Other material:	Lubrication & others $1,000/mo.	6,000
	Die, rollers, etc. $2/ton pellets	
	(2 x tons pellets/hr.)(2,640)	

Operating Cost Summary

Depreciation	$ 137,000
Maintenance	15,915
Labor	27,040

Operating material and utilities

$$\$25,884 + \left[76 + (22 \times \text{tons pellets/hr.})(0.06)(2,640) + (2 \times \text{tons pellets/hr.})(2,640)\right]$$

Total

$$\$205,839 + \left[76 + (22 \times \text{tons pellets/hr.})(0.06)(2,640) + (2 \times \text{tons pellets/hr.})(2,640)\right]$$

$$= \$205,839 + (2,640)\left[76 + (22 \times \text{tons pellets/hr.})(0.06) + (2 \times \text{tons pellets/hr.})\right]$$

For the return on investment computation, the cost of electrical energy used during the grinding season will not be taken into consideration, as the system has been adjusted to produce its energy requirement, and further the input for the DCF computation is expressed as incremental cost associated with the implementation of the proposed system.

<u>Appendix VI, page 6</u>

Input for DCF computation:

Depreciation	$ 137,000
Maintenance	15,915
Labor	27,040
Operating material and utilities	

$8,028 + 2,640(2 x tons·pellets/hr.)

Appendix VI, page 7

Schedule of Operating Expenses

System: Turbo-Generator and Auxiliaries

Type	Condensing	Topping	Extraction Condensing
Depreciation			
15 years straight line	$ 84,000	$ 59,333	$ 94,000
Maintenance			
Labor and material at 1.5% of delivered equipment cost for generator and cooling tower	10,800	7,650	12,150
Other operating material			
Lubrication, etc.	2,300	1,600	2,500
Total	$ 97,100	$ 68,583	$ 108,650

Appendix VI, page 8

Schedule of Operating Expenses

System: Pre-evaporator

Depreciation	
15 years straight line	$ 15,466
Maintenance	
At 1.5% of major components	2,490
Labor	-
Operating material and utilities	-
Total	$ 17,956

Appendix VII

Schedule of Revenues

	Excess Bagasse or Pellets BTU/lb.	Boiler Efficiency %	10^6 BTU Recoverable in Steam From Excess Bagasse or Pellets /Crop	Export 10^6 KWH /Crop	$ Revenues: From Recoverable BTU /Crop	$ Revenues: From Export KWH /Crop	$ Revenues: Total Revenues /Crop
Scenario 1							
Cond 1*	4,100	62.15	-	1.397	-	83,832	83,832
Cond 2	6,138	72.08	95,750	1.107	647,597	66,408	714,005
Cond 3	7,148	82.01	105,782	0.708	715,447	42,459	757,906
Scenario 2							
Cond 1	4,100	62.15	-	1.956	-	117,366	117,366
Cond 2	6,298	72.59	76,160	1.666	515,101	99,942	615,043
Cond 3	7,148	82.01	83,759	1.308	566,497	78,474	644,971
Scenario 3							
Cond 1	4,100	62.15	-	3.072	-	184,300	184,300
Cond 2	6,298	72.59	76,160	2.781	515,101	166,876	681,977
Cond 3	7,148	82.01	83,759	2.423	566,497	145,408	711,905
Scenario 4							
Cond 1	4,100	62.15	-	5.687	-	341,249	341,249
Cond 2	6,298	72.59	76,160	5.397	515,101	323,825	838,926
Cond 3	7,148	82.01	83,759	5.039	566,497	302,357	868,854
Scenario 5							
Cond 1	4,100	62.15	(24,792)	10.127	(167,678)	607,638	439,959
Cond 2	6,443	73.03	54,554	9.837	368,971	590,214	959,185
Cond 3	7,148	82.01	60,111	9.523	406,555	571,407	977,962
Scenario 6							
Cond 1	4,100	62.15	-	10.646	-	638,773	638,773
Cond 2	6,298	72.59	76,160	10.365	515,101	621,349	1,136,450
Cond 3	7,148	82.01	83,759	9.998	566,497	599,881	1,166,378

Constant: $6.763412/$10^6$ BTU recoverable in steam

$0.06/KWH

* No value is given to surplus of wet bagasse.

Appendix VII, page 2

Schedule of Revenues

	Excess Bagasse or Pellets BTU/lb.	Boiler Efficiency %	10^6 BTU Recoverable in Steam From Excess Bagasse or Pellets /Crop	Export 10^6 KWH /Crop	From Recoverable BTU /Crop	From Export KWH /Crop	Total Revenues /Crop
Scenario 2							
Cond 1a	4,100	62.15	-	2.536	-	152,138	152,138
Cond 1b	4,100	62.15	-	2.718	-	163,090	163,090
Cond 1c	4,100	62.15	-	3.491	-	209,452	209,452
Scenario 4							
Cond 4	7,148	82.01	121,176	5.039	819,563	302,357	1,121,920
Cond 5	7,148	82.01	136,868	5.039	925,697	302,357	1,228,054
Cond 6	7,148	82.01	187,026	5.039	1,264,932	302,357	1,567,289
Scenario 6							
Cond 4	7,148	82.01	121,594	9.998	822,389	599,881	1,422,270

Constant: $6.763412/$10^6$ BTU recoverable in steam

$0.06/KWH

Appendix VIII

Cash Inflow Schedule

Cash Inflow From Pre-Evaporator

Investment	Depreciation	Operating Cost
$ 232,000	$ 15,466	$ 17,956

Incremental Revenue

Scenario 3 Condition 1 vs Scenario 4 Condition 1

341,249 - 184,300 = $ 156,949

Scenario 3 Condition 2 vs Scenario 4 Condition 2

838,926 - 681,977 = $ 156,949

Scenario 3 Condition 3 vs Scenario 4 Condition 3

868,854 - 711,905 = $ 156,949

Total Incremental Revenue	$ 156,949
Operating Expenses	17,956
Net Incremental Revenue	138,993
Add Depreciation	15,466
Net Cash Inflow	154,459

Appendix VIII, page 2

Cash Inflow Schedule

Cash Inflow From Drying System

(45 tons/hr. bagasse)

Investment	Depreciation	Operating Cost
$ 1,770,000	$ 118,000	$ 151,488

Incremental Revenue

Scenario 2 Condition 1 vs Condition 2

615,043 - 117,366 = $ 497,677

Scenario 3 Condition 1 vs Condition 2

681,977 - 184,300 = $ 497,677

Scenario 4 Condition 1 vs Condition 2

838,926 - 341,249 = $ 497,677

Scenario 6 Condition 1 vs Condition 2

1,136,450 - 638,773 = $ 497,677

Incremental Revenue	$ 497,677
Operating Expenses	151,488
Net Incremental Revenue	346,189
Add Depreciation	118,000
Net Cash Inflow	464,189

Appendix VIII, page 3

Cash Inflow Schedule

Cash Inflow From Drying and Pelletizing System

(45 tons/hr. bagasse, 5 tons/hr. pellets)

Investment	Depreciation	Operating Cost
$ 2,055,000	$ 137,000	$ 202,271

Incremental Revenue

Scenario 2 Condition 1 vs Condition 3

644,971 - 117,366 = $ 527,605

Scenario 3 Condition 1 vs Condition 3

711,905 - 184,300 = $ 527,605

Scenario 4 Condition 1 vs Condition 3

868,854 - 341,249 = $ 527,605

Scenario 6 Condition 1 vs Condition 3

1,166,378 - 638,773 = $ 527,605

Total Incremental Revenue	$ 527,605
Operating Expenses	202,271
Net Incremental Revenue	325,334
Add Depreciation	137,000
Net Cash Inflow	462,334

Appendix VIII, page 4

Cash Inflow Schedule

Cash Inflow From Scenario 3 Condition 2

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Drying System	1,770,000	118,000	168,695
Total	3,030,000	202,000	265,795

Incremental Revenue

Scenario 2 Condition 1 vs Scenario 3 Condition 2

681,977 - 117,366 = $ 564,611

Total Incremental Revenue	$ 564,611
Operating Expenses	265,795
Net Incremental Revenue	298,816
Add Depreciation	202,000
Net Cash Inflow	500,816

Appendix VIII, page 5

Cash Inflow Schedule

Cash Inflow From Scenario 3 Condition 3

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Drying/Pelletizing System	2,055,000	137,000	202,271
Total	3,315,000	221,000	299,371

Incremental Revenue

Scenario 2 Condition 1 vs Scenario 3 Condition 3

711,905 - 117,366 = $ 594,539

Total Incremental Revenue	$ 594,539
Operating Expenses	299,371
Net Incremental Revenue	295,168
Add Depreciation	221,000
Net Cash Inflow	516,168

Appendix VIII, page 6

Cash Inflow Schedule

Cash Inflow From 3,000 KW Condensing Turbo-Generator, Cooling Tower and Pre-evaporator

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Pre-evaporator	232,000	15,466	17,956
Total	1,492,000	99,466	115,056

Incremental Revenue

Scenario 2 Condition 1 vs Scenario 4 Condition 1

341,249 - 117,366 = $ 223,883

Scenario 2 Condition 2 vs Scenario 4 Condition 2

838,926 - 615,043 = $ 223,883

Scenario 2 Condition 3 vs Scenario 4 Condition 3

868,854 - 644,971 = $ 223,883

Total Incremental Revenue	$ 223,883
Operating Expenses	115,056
Net Incremental Revenue	108,827
Add Depreciation	99,466
Total Cash Inflow	208,293

Appendix VIII, page 7

Cash Inflow Schedule

Cash Inflow From Scenario 4 Condition 2

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Pre-evaporator	232,000	15,466	17,956
Drying System	1,770,000	118,000	168,695
Total	3,262,000	217,466	283,751

Incremental Revenue

Scenario 2 Condition 1 vs Scenario 4 Condition 2

838,926 - 117,366 = $ 721,560

Total Incremental Revenue	$ 721,560
Operating Expenses	- 283,751
Net Incremental Revenue	437,809
Add Depreciation	217,466
Net Cash Inflow	655,275

Appendix VIII, page 8

Cash Inflow Schedule

Cash Inflow From Scenario 4 Condition 3

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Pre-evaporator	232,000	15,466	17,956
Drying/Pelletizing System	2,055,000	137,000	202,271
Total	3,547,000	236,466	317,327

Incremental Revenue

Scenario 2 Condition 1 vs Scenario 4 Condition 3

868,854 - 117,366 = $ 751,488

Total Incremental Revenue	$ 751,488
Operating Expenses	317,327
Net Incremental Revenue	434,161
Add Depreciation	236,466
Net Cash Inflow	670,627

Appendix VIII, page 9

Cash Inflow Schedule

Cash Inflow From Scenario 4 Condition 4

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Pre-evaporator	232,000	15,466	17,956
Drying System	2,055,000	137,000	208,654
Total	3,547,000	236,466	323,710

Incremental Revenue

Scenario 2 Condition 1a vs Scenario 4 Condition 4

1,121,920 - 152,138 = $ 969,782

Total Incremental Revenue	$ 969,782
Operating Expense	323,710
Net Incremental Revenue	646,072
Add Depreciation	236,466
Net Cash Inflow	882,538

Appendix VIII, page 10

Cash Inflow Schedule

Cash Inflow From Scenario 4 Condition 5

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Pre-evaporator	232,000	15,466	17,956
Drying/Pelletizing System	2,404,000	160,266	237,943
Total	3,896,000	259,732	352,999

Incremental Revenue

Scenario 2 Condition 1b vs Scenario 4 Condition 5

1,228,054 - 163,090 = $ 1,064,964

Total Incremental Revenue	$ 1,064,964
Operating Expenses	352,999
Net Incremental Revenue	711,965
Add Depreciation	259,732
Net Cash Inflow	971,697

Appendix VIII, page 11

Cash Inflow Schedule

Cash Inflow From Scenario 4 Condition 6

	Investment	Depreciation	Operating Cost
3,000 KW Condensing Turbo-Generator and Cooling Tower	$ 1,260,000	$ 84,000	$ 97,100
Pre-evaporator	232,000	15,466	17,956
Drying/Pelletizing System	2,404,000	160,266	246,494
Total	3,896,000	259,732	361,550

Incremental Revenue

Scenario 4 Condition 6 vs Scenario 2 Condition 1c

1,567,289 - 209,452 = $ 1,357,837

Total Incremental Revenue	$ 1,357,837
Operating Expenses	361,550
Net Incremental Revenue	996,287
Add Depreciation	259,732
Net Cash Inflow	1,256,019

Appendix VIII, page 12

Cash Inflow From Scenario 6 Condition 1

	Investment	Depreciation	Operating Cost
Boiler Differential	$ 2,100,000	$ 140,000	$ 140,000
4,500 KW Extracting/ Condensing Turbo-Generator and Cooling Tower	1,410,000	94,000	108,650
Pre-evaporator	232,000	15,466	17,956
Total	3,742,000	249,466	266,606

Incremental Revenue

Scenario 6 Condition 1 vs Scenario 2 Condition 1

638,773 - 117,366 = $ 521,407

Total Incremental Revenue	$ 521,407
Operating Expenses	266,606
Net Incremental Revenue	254,801
Add Depreciation	249,466
Net Cash Inflow	504,267

Appendix VIII, page 13

Cash Inflow Schedule

Cash Inflow From Scenario 6 Condition 2

	Investment	Depreciation	Operating Cost
Boiler Differential	$ 2,100,000	$ 140,000	$ 140,000
4,500 KW Extracting/ Condensing Turbo-Generator and Cooling Tower	1,410,000	94,000	108,650
Pre-evaporator	232,000	15,466	17,956
Drying System	1,770,000	118,000	151,488
Total	5,512,000	367,466	418,094

Incremental Revenue

Scenario 6 Condition 2 vs Scenario 2 Condition 1

1,136,450 - 117,366 = $ 1,019,084

Total Incremental Revenue	$ 1,019,084
Operating Expenses	418,094
Net Incremental Revenue	600,990
Add Depreciation	367,466
Net Cash Inflow	968,456

Appendix VIII, page 14

Cash Inflow Schedule

Cash Inflow From Scenario 6 Condition 3

	Investment	Depreciation	Operating Cost
Boiler Differential	$ 2,100,000	$ 140,000	$ 140,000
4,500 KW Extracting/ Condensing Turbo-Generator and Cooling Tower	1,410,000	94,000	108,650
Pre-evaporator	232,000	15,466	17,956
Drying/Pelletizing System	2,055,000	137,000	202,271
Total	5,797,000	386,466	468,877

Incremental Revenue

Scenario 6 Condition 3 vs Scenario 2 Condition 1

1,166,378 - 117,366 = $ 1,049,012

Total Incremental Revenue	$ 1,049,012
Operating Expenses	468,877
Net Incremental Revenue	580,135
Add Depreciation	386,466
Net Cash Inflow	966,601

Appendix VIII, page 15

Cash Inflow Schedule

Cash Inflow From Scenario 6 Condition 4

	Investment	Depreciation	Operating Cost
Boiler Differential	$ 2,100,000	$ 140,000	$ 140,000
4,500 KW Extracting/ Condensing Turbo-Generator and Cooling Tower	1,410,000	94,000	108,650
Pre-evaporator	232,000	15,466	17,956
Drying/Pelletizing System	2,055,000	137,000	208,726
Total	5,797,000	386,466	475,332

Incremental Revenue

Scenario 6 Condition 4 vs Scenario 2 Condition 1a

1,422,270 - 152,138 = $ 1,270,132

Total Incremental Revenue	$ 1,270,132
Operating Expenses	475,332
Net Incremental Revenue	794,800
Add Depreciation	386,466
Net Cash Inflow	1,181,266

Appendix VIII, page 16

Cash Inflow Schedule

Cash Inflow From Scenario 6 Condition 2a

(Excluding boiler)

	Investment	Depreciation	Operating Cost
4,500 KW Extracting/ Condensing Turbo-Generator and Cooling Tower	$ 1,410,000	$ 94,000	$ 108,650
Pre-evaporator	232,000	15,466	17,956
Drying System	1,770,000	118,000	151,488
Total	3,412,000	227,466	278,094

Incremental Revenue

Scenario 6 Condition 2a vs Scenario 2 Condition 1

1,136,450 - 117,366 = $ 1,019,084

Total Incremental Revenue	$ 1,019,084
Operating Expenses	278,094
Net Incremental Revenue	740,990
Add Depreciation	227,466
Net Cash Inflow	968,456

Appendix VIII, page 17

Cash Inflow Schedule

Cash Inflow From Scenario 6 Condition 3a

(Excluding boiler)

	Investment	Depreciation	Operating Cost
4,500 KW Extracting/ Condensing Turbo-Generator and Cooling Tower	$ 1,410,000	$ 94,000	$ 108,650
Pre-evaporator	232,000	15,466	17,956
Drying/Pelletizing System	2,055,000	137,000	202,271
Total	3,697,000	246,466	328,877

Incremental Revenue

Scenario 6 Condition 3a vs Scenario 2 Condition 1

1,166,378 - 117,366 = $ 1,049,012

Total Incremental Revenue	$ 1,049,012
Operating Expenses	328,877
Net Incremental Revenue	720,135
Add Depreciation	246,466
Net Cash Inflow	966,601

SUMMARY OF DISCOUNTED CASH FLOW RETURN ON INVESTMENT AND PAYBACK PERIOD CALCULATIONS

1	Pre-Evaporator	Scenario 4 vs Scenario 3	66.15%
2	Drying System	Applicable to All Scenarios	22.85%
3	Drying and Pelletizing System	Applicable to All Scenarios	18.30%
4	3 MW Turbo-Generator and Drying System	Scenario 3 Condition 2 vs Scenario 2 Condition 1	10.35%
5	3 MW Turbo-Generator, Drying and Pelletizing System	Scenario 3 Condition 3 vs Scenario 2 Condition 1	8.95%
6	3 MW Turbo-Generator and Pre-Evaporator	Scenario 4 Condition 1 vs Scenario 2 Condition 1	6.55%
7	3 MW Turbo-Generator, Pre-Evaporator and Drying System	Scenario 4 Condition 2 vs Scenario 2 Condition 1	15.20%
8	3 MW Turbo-Generator, Pre-Evaporator, Drying and Pelletizing	Scenario 4 Condition 3 vs Scenario 2 Condition 1	13.60%
9	3 MW Turbo-Generator, Pre-Evaporator and Drying System	Scenario 4 Condition 4 vs Scenario 2 Condition 1a	21.25%
10	3 MW Turbo-Generator, Pre-Evaporator, Drying and Pelletizing	Scenario 4 Condition 5 vs Scenario 2 Condition 1b	21.30%
11	3 MW Turbo-Generator, Pre-Evaporator, Drying and Pelletizing	Scenario 4 Condition 6 vs Scenario 2 Condition 1c	29.85%
12	Boiler Differential, 4.5 MW Turbo-Generator and Pre-Evaporator	Scenario 6 Condition 1 vs Scenario 2 Condition 1	5.80%
13	Boiler Differential, 4.5 MW Turbo-Generator, Pre-Evap. & Drying	Scenario 6 Condition 2 vs Scenario 2 Condition 1	11.80%
14	Boiler Differential, 4.5 MW Turbo-Gen., Pre-Evap., Dry, Pelletize	Scenario 6 Condition 3 vs Scenario 2 Condition 1	10.55%
15	Boiler Differential, 4.5 MW Turbo-Gen., Pre-Evap., Dry, Pelletize	Scenario 6 Condition 4 vs Scenario 2 Condition 1a	15.55%
16	4.5 MW Turbo-Generator, Pre-Evaporator and Drying System	Scenario 6 Condition 2a vs Scenario 2 Condition 1	25.40%
17	4.5 MW Turbo-Generator, Pre-Evaporator, Drying and Pelletizing	Scenario 6 Condition 3a vs Scenario 2 Condition 1	22.75%

Appendix IX

Page 1

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

PRE-EVAPORATOR - SCENARIO 4 vs SCENARIO 3

Cash Outflow	$
Year #0	232,000
Year #1	
Total Outflow	232,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$1,544,590
Average Inflow per Year	$154,459

Trial Results

Discounted Cash Flow Rate of Return	66.15%
Payback Period	1.50 Years

Appendix IX

Page 2

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

DRYING SYSTEM - APPLICABLE TO ALL SCENARIOS

Cash Outflow	$
Year #0	1,770,000
Year #1	
Total Outflow	1,770,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$4,641,890
Average Inflow per Year	$464,189

Trial Results

Discounted Cash Flow Rate of Return	22.85%
Payback Period	3.81 Years

Appendix IX

Page 3

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

DRYING & PELLETIZING SYSTEM - APPLICABLE TO ALL SCENARIOS

Cash Outflow	$
Year #0	2,055,000
Year #1	
Total Outflow	2,055,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$4,623,340
Average Inflow per Year	$462,334

Trial Results

Discounted Cash Flow Rate of Return	18.30%
Payback Period	4.44 Years

Appendix IX

Page 4

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR & DRYING SYSTEM -
SCENARIO 3 CONDITION 2 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	3,030,000
Year #1	
Total Outflow	3,030,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$5,008,160
Average Inflow per Year	$500,816

Trial Results

Discounted Cash Flow Rate of Return	10.35%
Payback Period	6.05 Years

Appendix IX

Page 5

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR, DRYING & PELLETIZING -
SCENARIO 3 CONDITION 3 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	3,315,000
Year #1	
Total Outflow	3,315,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$5,161,680
Average Inflow per Year	$516,168

Trial Results

Discounted Cash Flow Rate of Return	8.95%
Payback Period	6.42 Years

Appendix IX

Page 6

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR & PRE-EVAPORATOR - SCENARIO 4 CONDITION 1 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	1,492,000
Year #1	
Total Outflow	1,492,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$2,082,930
Average Inflow per Year	$208,923

Trial Results

Discounted Cash Flow Rate of Return	6.55%
Payback Period	7.16 Years

Appendix IX

Page 7

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR, PRE-EVAPORATOR & DRYING SYSTEM - SCENARIO 4 CONDITION 2 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	3,262,000
Year #1	
Total Outflow	3,262,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$6,552,750
Average Inflow per Year	$655,275

Trial Results

Discounted Cash Flow Rate of Return	15.20%
Payback Period	4.98 Years

Appendix IX

Page 8

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR, PRE-EVAPORATOR, DRYING & PELLETIZING - SCENARIO 4 CONDITION 3 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	3,547,000
Year #1	
Total Outflow	3,547,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$6,706,270
Average Inflow per Year	$670,627

Trial Results

Discounted Cash Flow Rate of Return	13.60%
Payback Period	5.29 Years

Appendix IX

Page 9

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR, PRE-EVAPORATOR & DRYING SYSTEM - SCENARIO 4 CONDITION 4 vs SCENARIO 2 CONDITION 1a

Cash Outflow	$
Year #0	3,547,000
Year #1	
Total Outflow	3,547,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$8,825,380
Average Inflow per Year	$882,538

Trial Results

Discounted Cash Flow Rate of Return	21.25%
Payback Period	4.02 Years

Appendix IX

Page 10

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR, PRE-EVAPORATOR, DRYING & PELLETIZING - SCENARIO 4 CONDITION 5 vs SCENARIO 2 CONDITION 1b

Cash Outflow	$
Year #0	3,896,000
Year #1	
Total Outflow	3,896,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$9,716,970
Average Inflow per Year	$971,697

Trial Results

Discounted Cash Flow Rate of Return	21.30%
Payback Period	4.01 Years

Appendix IX

Page 11

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR, PRE-EVAPORATOR, DRYING & PELLETIZING - SCENARIO 4 CONDITION 6 vs SCENARIO 2 CONDITION 1c

Cash Outflow	$
Year #0	3,896,000
Year #1	
Total Outflow	3,896,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$12,560,190
Average Inflow per Year	$1,256,019

Trial Results

Discounted Cash Flow Rate of Return	29.85%
Payback Period	3.10 Years

Appendix IX

Page 12

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

BOILER DIFFERENTIAL, 4.5 MW TURBO-GENERATOR & PRE-EVAPORATOR - SCENARIO 6 CONDITION 1 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	3,742,000
Year #1	
Total Outflow	3,742,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$5,042,670
Average Inflow per Year	$504,267

Trial Results

Discounted Cash Flow Rate of Return	5.80%
Payback Period	7.42 Years

Appendix IX

Page 13

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

BOILER DIFFERENTIAL, 4.5 MW TURBO-GENERATOR, PRE-EVAPORATOR & DRYING - SCENARIO 6 CONDITION 2 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	5,512,000
Year #1	
Total Outflow	5,512,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$9,684,560
Average Inflow per Year	$968,456

Trial Results

Discounted Cash Flow Rate of Return	11.80%
Payback Period	5.69 Years

Appendix IX

Page 14

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

BOILER DIFFERENTIAL, 4.5 MW TURBO-GENERATOR, PRE-EVAPORATOR, DRY & PELLETIZE - SCENARIO 6 CONDITION 3 vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	5,797,000
Year #1	
Total Outflow	5,797,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$9,666,010
Average Inflow per Year	$966,601

Trial Results

Discounted Cash Flow Rate of Return	10.55%
Payback Period	6.00 Years

Appendix IX

Page 15

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

BOILER DIFFERENTIAL, 4.5 MW TURBO-GENERATOR, PRE-EVAPORATOR, DRY & PELLETIZE - SCENARIO 6 CONDITION 4 vs SCENARIO 2 CONDITION 1a

Cash Outflow	$
Year #0	5,797,000
Year #1	
Total Outflow	5,797,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$11,812,660
Average Inflow per Year	$1,181,266

Trial Results

Discounted Cash Flow Rate of Return	15.55%
Payback Period	4.91 Years

Appendix IX

Page 16

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

4.5 MW TURBO-GENERATOR, PRE-EVAPORATOR & DRYING SYSTEM - SCENARIO 6 CONDITION 2a vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	3,412,000
Year #1	
Total Outflow	3,412,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$9,684,560
Average Inflow per Year	$968,456

Trial Results

Discounted Cash Flow Rate of Return	25.40%
Payback Period	3.52 Years

Appendix IX

Page 17

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

4.5 MW TURBO-GENERATOR, PRE-EVAPORATOR, DRYING & PELLETIZING - SCENARIO 6 CONDITION 3a vs SCENARIO 2 CONDITION 1

Cash Outflow	$
Year #0	3,697,000
Year #1	
Total Outflow	3,697,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$9,666,010
Average Inflow per Year	$966,601

Trial Results

Discounted Cash Flow Rate of Return	22.75%
Payback Period	3.82 Years

Appendix IX

Page 18

Discounted Cash Flow Return on Investment and Payback Period Calculations

Project Description Title

3 MW TURBO-GENERATOR - SCENARIO 2 CONDITION 1 vs SCENARIO 3 CONDITION 1

Cash Outflow	$
Year #0	1,260,000
Year #1	
Total Outflow	1,260,000

Number of Years of Cash Inflow	10
Total Inflow (Year 1 to 10)	$1,042,740
Average Inflow per Year	$104,274 .

Trial Results

Discounted Cash Flow Rate of Return	.05%
Payback Period	12.08 Years

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