ENERGY EFFICIENCY IMPROVEMENTS TO HULETTS SUGAR REFINERY

RG HOEKSTRA[1] AND DJ TAYFIELD[2]

[1]*Tongaat-Hulett Sugar Limited, Private Bag 3, Glenashley, 4022*
[2]*Tongaat-Hulett Sugar Refinery, PO Box 1501, Durban, 4000*

Abstract

The central Tongaat-Hulett Sugar Refinery in Rossburgh generates steam and part of its electrical power requirement from burning coal, with the balance of electricity being purchased from the Durban Corporation grid. Energy is one of the Refinery's major cost components. The efficient use of this commodity is therefore important to its financial viability. Outside consultants examined the energy balance and applied, among others, the principles of pinch technology. On the basis of their report, the following changes were made:

- The single effect evaporator was converted to a double effect, by adding a plate evaporator as first stage and a plate heat exchanger as pre-heater.
- Flash from the condensate, mainly from the pan calandrias, is drawn into the Vapour 2 line of the evaporator instead of venting into the atmosphere.
- Vapour 3 from the syrup flash is used for pre-heating the boiler feed water make-up, using a plate heat exchanger.
- High pressure (HP) steam is no longer let down directly into the 510 kPa(a) process vapour line, but through a thermo-compressor, which draws 220 kPa(a) exhaust vapour from the back-pressure turbo-alternator, enabling this machine to generate additional in-house electrical power.

The use of plate evaporators on a factory scale and of thermo-compressors is fairly new to the South African sugar industry, and these have thus far performed without any problems.

Among the advantages arising from this project are:

- Reduced overall steam consumption, resulting in coal cost savings.
- Lower purchased electricity cost.
- Increased availability of condensate for boiler feed water, with correspondingly reduced need for expensive treated raw water for make-up.

Introduction

All the raw sugar produced from the mills of the Tongaat-Hulett Sugar group in South Africa, which is destined for white sugar, is refined by the central Tongaat-Hulett Sugar Refinery at Rossburgh in Durban. Although a central refinery has the economic advantage of large scale operation and a longer season, compared with white-end refineries attached to individual sugar mills, it has the disadvantage that, energy-wise, it cannot be integrated with any raw sugar mills. All heat energy has to be produced by burning coal, which is an important factor in the economics of running the refinery.

Most of the energy is required for evaporation, be it for pre-concentrating the sugar liquor or for evaporative crystallisation in batch pans. The pan steam requirement could be reduced by utilising the vapour from the pans to heat a second stage of pans operating at lower pressures, as is sometimes practised in the beet sugar industry, but this would require major capital expenditure. Thus most of the attention in recent times has been on reducing the steam required by the existing evaporator.

Other cost components under consideration were:

- the treatment cost for boiler feed water make-up, which could be reduced by increasing the amount of recovered condensate and/or reducing the amount of HP steam required.
- the cost of bought-in electricity from the power utility, which could be reduced by being more self sufficient in own power production.

Plant set-up prior to modifications

Evaporation

The evaporator and its associated units are shown as a process diagram in Figure 1. It is a single-effect evaporator, consisting of two similarly sized Robert vessels in parallel. Tayfield (1988) provided a description of the system, in that case with the emphasis on condensate recovery. As can be seen from the diagram, part of the Vapour 1 (V1) from the evaporators is usefully applied in the vapour melter for melting the raw sugar and in the C-heater for pre-heating the fine water going to process, but the surplus as well as the flash vapour coming from the syrup flash vessel after the evaporator are not utilised, but condensed in the B- and the A-condenser respectively. The evaporator uses low pressure (LP) vapour at 220 kPa(a).

Pans

Four of the five white pans are of the ribbon calandria design, and have to operate at a higher calandria pressure of 510 kPa(a), referred to as the intermediate pressure (IP) vapour.

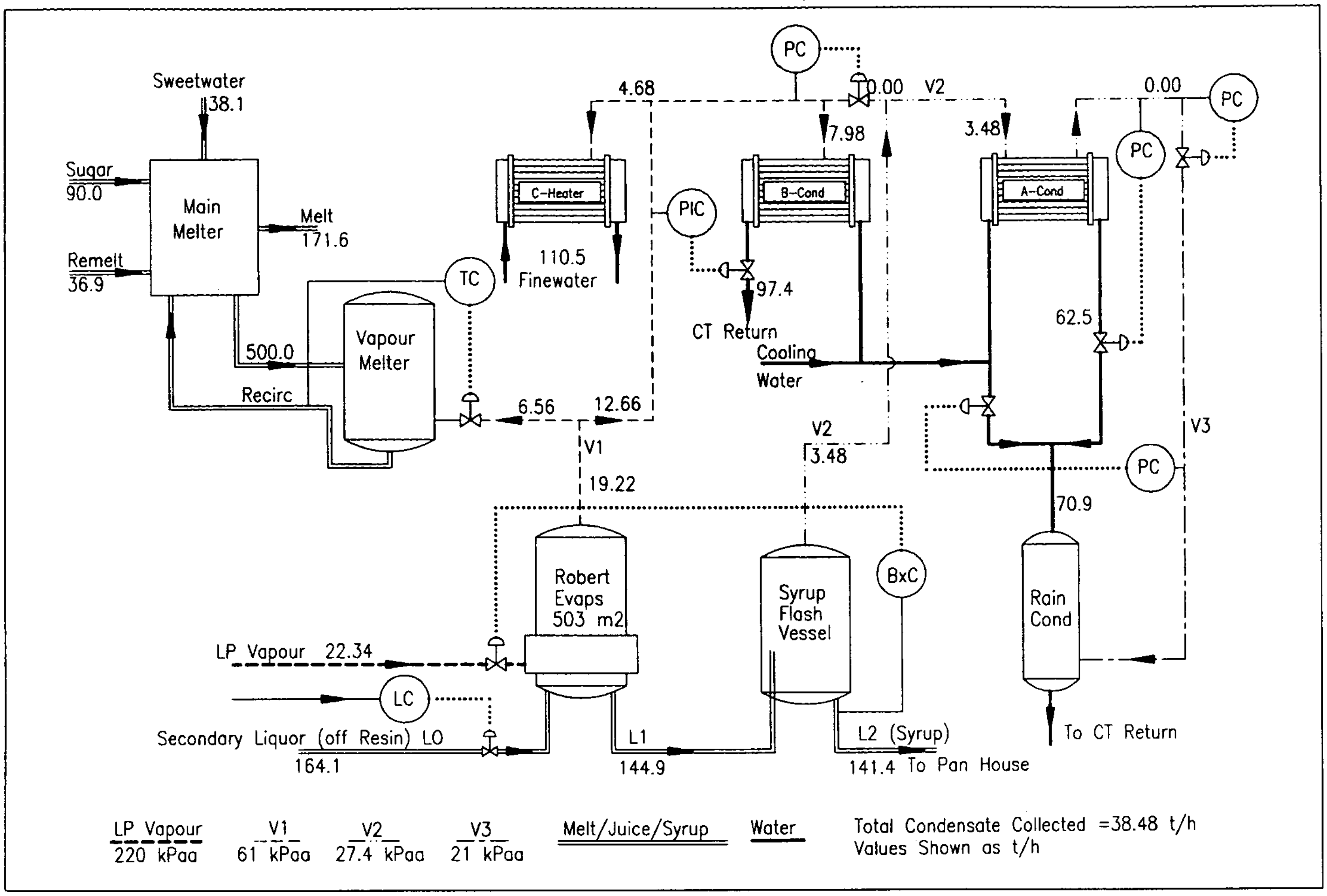

Figure 1. Process diagram of single-effect evaporator system prior to changes.

The remaining white pan and the recovery pans with tubular calandrias as well as the evaporator and process in general use exhaust vapour which, after desuperheating, becomes part of the LP vapour.

Boilers

There are two classes of boilers which deliver high pressure steam: the 3 200 kPa(a) (HP1) boilers are the main supply, and are used for the turbo-alternator (TA), and the remainder of their steam is let down to be added to that from the 1 360 kPa(a) (HP2) boilers which, on letting down and desuperheating, is used mainly for heating purposes.

The IP vapour for the ribbon calandria white pans is obtained by letting down from the HP2 main. The LP vapour for the evaporator is obtained from the exhaust of the TA and, if necessary, can be made up by letting down from the HP2 line.

The basics of the steam supply are illustrated in Figure 2, excluding desuperheating and the accumulator.

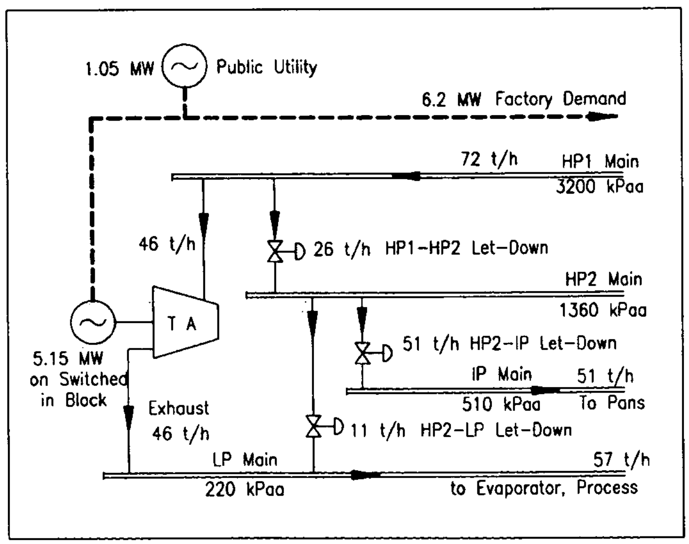

Figure 2. Steam supply arrangement prior to alterations (average conditions).

Turbo-alternator

Power is usually generated by a large turbo-alternator, using high pressure (HP1) steam at 3 200 kPa(a) and exhausting at 220 kPa(a) into the LP vapour line.

Under South African conditions, it is usually cheaper to buy power from an electricity utility than to produce it in-house through a condensing TA. If, however, the power is produced by a back-pressure TA, with the exhaust in any case being needed for process heating purposes, then such power is far cheaper than from an electric power utility. Compared with directly letting the high pressure steam down and desuperheating, a small additional amount of HP steam is required when producing such power from a back pressure TA, because the directly let-down steam with its higher enthalpy will absorb more desuperheating water than the exhaust vapour, thus requiring a smaller amount of HP steam for the same amount of process heating steam. This increase is, however, much less than the amount of HP steam required for producing the same power on a condensing TA.

At the refinery there is a problem in that the TA cannot be used to supply the full factory power requirements, not through lack of capacity, but because the exhaust vapour from the TA is limited by the LP process vapour consumption of the evaporator and tubular calandria pans. This obviously limits the amount of HP1 steam to the TA by the same amount, and hence the electric power produced.

One solution would be to shift the existing IP vapour consumption to additional LP vapour consumption by converting the existing white pans from ribbon to tubular calandria, but that would be a major capital project. Furthermore, the refinery does not have a facility for parallel generation, whereby the TA could produce as much power as possible within the constraints of LP vapour demand *at any moment,* with the public utility automatically providing whatever additional power is required. The refinery currently uses block allocation, by which the TA is coupled to blocks of power consuming equipment which best match the power the TA will produce at the *minimum likely* LP process vapour demand of about 46 t/h. The remaining factory power consuming equipment is connected to and supplied by the power utility. When operating at an average LP vapour demand of 57 t/h, the additional 11 t/h have to be made up by letting down from the 1 360 kPa(a) HP2 line, because it is not feasible to block-switch automatically to vary the TA load to suit its exhaust rate to the LP vapour consumption. The result is that the refinery has to purchase a significant part of its power requirement, which it could otherwise more economically have generated in-house.

Furthermore, the above 'minimum likely' LP process vapour demand is not an absolute lower limit; on occasions it can go below that value, in which case the TA exhaust will be in over production and result in blow-off to atmosphere, which wastes an average of about 2 t/h of LP vapour – and of treated boiler feed-water.

Earlier investigations

The most obvious scope for reducing energy consumption in the refinery was to convert the existing single effect evaporator to a double-effect arrangement, by using the new vessel as the first effect and the existing Robert vessels as the second effect. A spreadsheet based steady state simulation model of the evaporator section was developed, and was extremely useful in simulating various scenarios. At the outset it was decided that the new vessel should be either a falling film (FF) or a plate evaporator (PE), because of the lower retention times. The FF evaporator was favoured initially, because the necessary technology was already established in the beet sugar industry. A drawback was fitting the great height of the evaporator into the building.

As the PE became more established, subsequent investigations favoured the PE, because:

- it was compact and fitted easily into the existing layout
- it could easily be expanded in the future by adding plates
- from a scaling point of view, the refinery operating conditions are probably less severe than in a beet sugar factory, where the PE had already proven itself.

As expected, the model showed that a fair amount of LP vapour could be saved by a double effect evaporator. Unfortunately, a counter-effect reduced the economics of the proposal: reduced LP vapour consumption of the evaporator also meant reduced exhaust vapour production from the TA and hence its ability to provide in-house electric power, with a correspondingly larger requirement for the more expensive power from the public utility.

The outcome from these investigations was that the net financial savings could not adequately justify the capital expenditure for the double effect arrangement.

Investigation by consultants

Refinery Management decided that a more comprehensive investigation into the energy economics of the refinery was required, and appointed specialist consultants in this field. This mainly involved applying the principles of pinch technology, which aspect of the project is briefly described by Rein (1997). By using this method of analysis, various other ways to save energy, which had not been so obvious, were noted.

The main findings were:

- Convert the single effect to a double effect evaporator, as expected.
- Only one liquor pre-heater plate heat exchanger (PHE), using V1, would be required, without the need for a second stage using LP vapour, because a plate evaporator does not require feed to be at its boiling point (another advantage over the FF evaporator).
- The treated water or feed to the boilers could be preheated by a PHE, using a lower grade Vapour 3, thus saving on LP vapour at the de-aerator.
- The flash to atmospheric pressure of condensate from the various calandrias could be used in a PHE to heat the liquor in the C-saturator prior to filtering, instead of using open LP vapour as at present. However, this proposal was not accepted because refinery Management had misgivings that the carbonatated liquor, with its suspended solids, would cause blockages in the PHE.
- A thermo-compressor, using HP1 steam as motive steam, could be used to draw in exhaust (LP) vapour and

discharge into the 510 kPa(a) (IP) vapour line. In so doing, more 'space' for the exhaust vapour from the TA is created, thus enabling the TA to produce more in-house power. The amount of 1 350 kPa(a) HP2 steam let down into the IP line will of course be reduced by the amount of discharge from the thermo-compressor.

New evaporator arrangement

Figure 3 shows the process flow diagram for the new double effect evaporator arrangement. Note that the former V1 becomes V2, etc, due to the additional effect.

Seeing that the abovementioned flash from pan calandria condensates could not feasibly be applied to heating the C-saturator liquor, thought was given to using it elsewhere. In the original concept, the amount of Vapour 2 (V2) produced by the second stage evaporators would have been too low to supply the vapour melter and the C-heater for pre-heating the fine water used in process, so that supplementing by 1,5 t/h LP vapour would have been necessary. By increasing the area of the juice pre-heater PHE, reducing the area of the first effect PE and by drawing flash vapour from the calandria condensates into the V2 line (which is feasible, because the V2 line operates below atmospheric pressure), the model showed that the above LP vapour would not normally be necessary, and 4 t/h of treated boiler feed water would be saved.

Thermo-compressor arrangement

The integration of the thermo-compressor is diagrammatically shown in Figure 4. An important characteristic of a thermo-compressor is its limitation to a single throughput rate, i.e. it is not feasible to throttle it for reduced throughputs. When a varying throughput is desired, an array of units of different size should be used, with on/off switching to provide the desired rate.

By the nature of the refinery, there are various constraints which will affect the operation of the thermo-compressors:

- The maximum boiler supply rate of 3 200 kPa(a) (HP1) steam, which drives the turbo-alternator and provides motive steam for the thermo-compressor, is 72 t/h.
- The total thermo-compressor discharge rate may not exceed the demand for 510 kPa(a) (IP) vapour from the white pans, which could be as low as 35 t/h.
- The exhaust vapour production rate less the thermo-compressor entrainment rate may not exceed the rate of 220 kPa(a) (LP) process vapour consumption, which could go as low as 33,6 t/h when the steam saving measures are implemented.

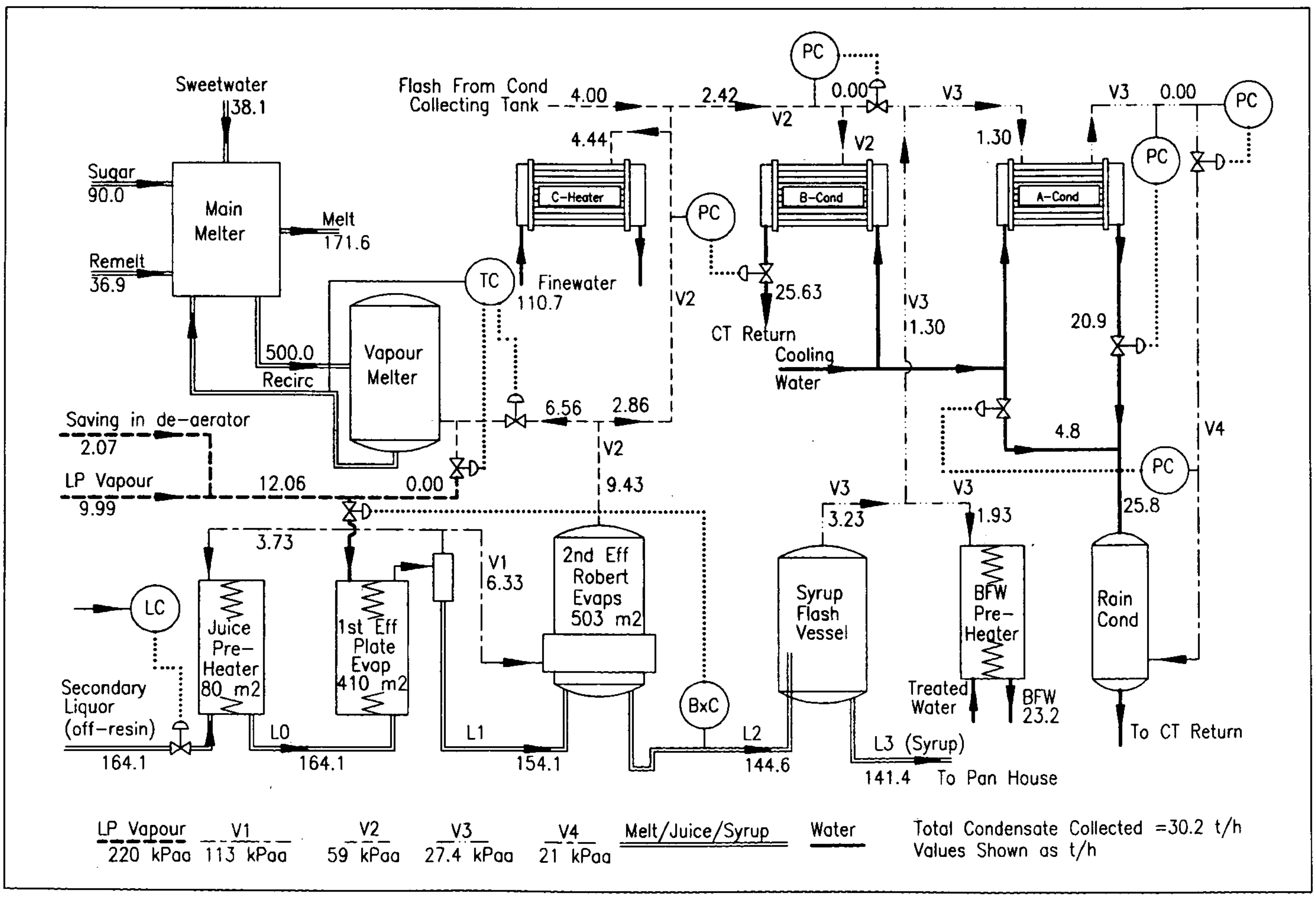

Figure 3. Process diagram of double-effect evaporator system.

Because of these constraints, an array of four units was proposed in the consultant's submission to provide throughputs ranging from 25 to 50 t/h discharge rate. For each unit, the ratio of motive steam:entrainment vapour:discharge vapour was 1:0,634:1,634. To help understand this complexity of the interacting constraints, the situation was simulated on a spreadsheet. Its results, taken for different values of the variables, are shown in Figure 5. Note that all cases, including 'no thermo-compression', apply to having implemented the double effect evaporator and other savings. In considering Figure 5, the objective is that the in-house power that can be generated by the TA for any given LP vapour demand rate should be as high as possible. The estimated probability distribution of LP vapour demand rates is also shown, including the region of blow-off.

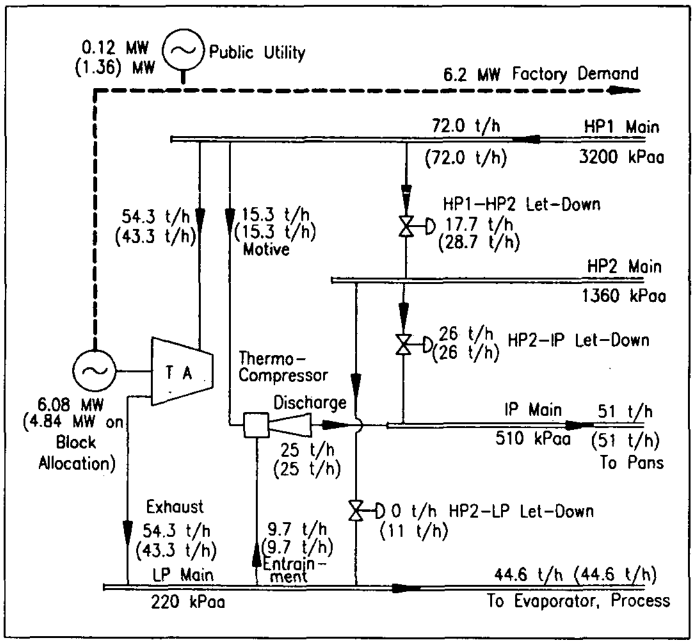

Figure 4. Integration of 25 t/h thermo-compressor, assuming parallel generation. Values in parenthesis apply to case of block allocation.

When using the system of block allocation, TA operation is constrained to operate at the likely minimum LP vapour consumption, i.e. on line FGH, and will produce in-house power as follows:

Case	Point	MW generated
No thermo-compression	H	3,75
One 25 t/h unit	G	4,84
Full array	F	5,40

For comparison, point L is shown, which represents the single effect situation, without thermo-compression. Its fairly good power self sufficiency of 5,15 MW is due to the (uneconomically) high LP vapour consumption. It is immediately obvious that a considerable amount of in-house power can be gained by using thermo-compression, although most of the gain is achieved by using only the single 25 t/h unit.

When parallel generation is used, the TA is no longer constrained to operate on the line FGH, but can follow the range of LP vapour demand along curves FJK or GJK, depending on whether the full array or a single unit is used, and produce far more in-house power. The full array, with all its additional expense and control complexity, now shows a lesser power self sufficiency advantage over the single unit, particularly because the probability of operating near the unfavourable points F or G is low. Indeed, at the average LP vapour consumption rate of 44,6 t/h, the array gains little. In addition, at 25 t/h discharge rate, constraint (b) above will not be encountered. It was therefore decided on the single 25 t/h unit. Figure 4 compares the operating values for parallel generation and for block switching applicable to average LP vapour demand.

Another bonus of parallel generation is that the TA output can follow the LP vapour demand at all times, including below the 'likely minimum', i.e. to the left of line FGH, without blow-off taking place, thus saving on LP vapour lost to atmosphere and on treated BFW.

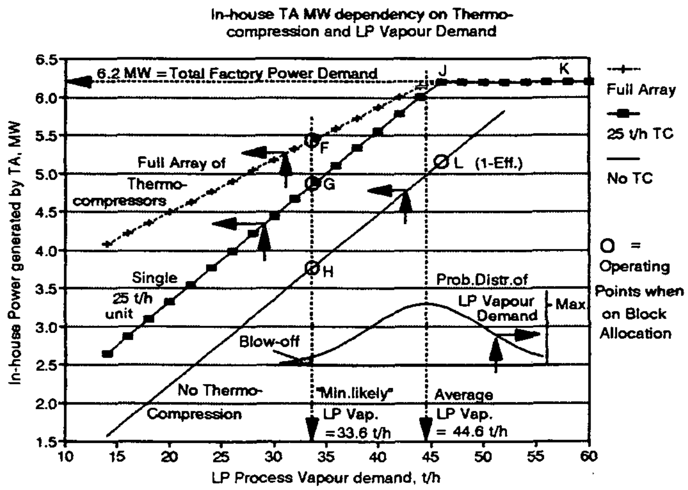

Figure 5. Effect of LP vapour demand and thermo-compression on in-house power generated.

Calculated benefits of the project

Unfortunately, at the time of writing, the operating results thus far have not been sufficient in quantity nor of adequate consistency to justify quoting any factory figures. The new equipment has been operating for only 2½ months; the system for the recovery of the calandria condensate flash vapour has just been completed, and the replacement of block switching by parallel generation to improve power self- sufficiency is still underway. Furthermore, due to temporary filtration problems caused by raw sugar quality, the liquor feed to the evaporators is more dilute than normal, which affects steam consumption adversely.

At this stage, it is only possible to give the calculated benefits of the project. These are shown in Table 1. Note that two possible outcomes are shown for the project: for block allocation, as is presently still applied, and for parallel generation, which is the final intention.

Table 1. Changes in tonnages and costs with energy saving measures.

	Tonnages and costs			Savings w.r.t. 1-Effect	
	1-Effect	2-Effect, Block	2-Effect Parallel	2-Effect, Block	2-Effect, Parallel
Consumed by factory					
IP vapour for white pans, t/h	51,0	51,0	51,0	0	0
LP vapour for process and deaerator, t/h	57,0	44,6	44,6	12,4	12,4
LP (exhaust) vapour blow-off, t/h	2,0	2,0	0	0	2,0
Produced from					
HP1 steam (3 200 kPaa), t/h	72,0	72,0	72,0	0	0
HP2 steam (1 360 kPaa), t/h	29,1	17,7	17,2	11,4	11,9
Desuperheating water, t/h	8,9	7,9	6,4	1,0	2,5
Coal for above HP steam, t/h	13,1	11,6	11,5	1,5	1,6
(a) Cost of coal, Rand/h	R2 241,42	R1 980,67	R1 968,88	R260,75	R272,54
Treated water, t/h	36,2	31,6	29,6	4,6	6,6
(b) Cost of treated water, Rand/h	R79,67	R69,60	R65,16	R10,07	R14,51
Purchased power, MW	1,05	1,36	0,40	−0,31	0,65
(c) Cost of purchased power, Rand/h	R115,50	R149,60	R44,00	−R34,10	R71,50
Total cost = (a)+(b)+(c), Rand/h	R2436,58	R2199,87	R2078,03	R236,71	R358,55
Total annual cost, Rand/year	R18,274 million	R16,499 million	R15,585 million	R1,775 million	R2,689 million

The tonnages and cost figures might exclude some values which would be common and constant to all cases, but in calculating the savings of the project, such common values would cancel out anyway.

The bottom line of Table 1 shows the calculated annual savings for the project. Even when allowing that some of the benefits might not turn out as ideally as calculated, the savings remain considerable.

Remarks on operations thus far

At the time of writing, most of the new arrangment, including the PE and the PHEs, have been operating for about 2½ months. The connection of condensate flash to the V2 line has yet to be done, and parallel power generation continues to be investigated. The following points might be of interest:

- The PHEs and the PE have given no problems at all. There has been no indication of scaling in the PE, and capacity-wise it is adequately coping with the load.
- The thermo-compressor is working successfully, although its performance is sensitive to pressures in the LP and the IP vapour lines.
- The thermo-compressor has provided a bonus in that pressure regulation in the IP vapour line has become more stable.
- The existing two Robert vessels, which are now in the second effect, are not boiling as satisfactorily as previously, the probable cause being that, with V1 at about 113 kPa(a) instead of LP vapour at 220 kPa(a) on the calandria side, the overall temperature difference Δt across the calandria tubes is now considerably less. Because heat transfer coefficient is, in itself, dependent on Δt, the relative performance could drop. Overall, the evaporator-section is coping perfectly well.
- There had been misgivings about the noise level of the thermo-compressor, but the thermal insulation around it and the connecting pipework reduces the noise level to no higher than that of the general factory background.

REFERENCES

Rein, PW (1997). The use of pinch technology in identifying energy savings. *Sugar Ind Techn Inc* Vol. 56: 313-317.

Tayfield, DJ (1988). Optimisation of condensate recovery at Huletts Refineries. *Proc S Afr Sug Technol Ass* 62: 90-93.