USING PINCH TECHNOLOGY TO OPTIMISE EVAPORATOR AND VAPOUR BLEED CONFIGURATION AT THE MALELANE MILL

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Abstract

The Malelane mill has recently embarked on a programme of energy savings. A comprehensive pinch analysis has been undertaken to determine the best investment strategy for TSB Malelane. This paper examines the use of pinch analysis to determine the optimum evaporator and bleed configuration commensurate with the energy saving benefits, minimum inversion losses and minimum colour formation. Evaporator temperature, residence time and brix profiles have been evaluated for each design. Capital/energy trade-offs have been considered in selecting a final design for present and future throughput scenarios.

Keywords: pinch technology, evaporator, vapour bleed, Malelane

Introduction

The TSB Malelane factory operations encompass a raw sugar factory with a back-end refinery, refining all the internally produced raw sugar plus imported raw sugar. Additional activities include the export of bagasse fibre to a board factory, bagasse pith to an animal feeds plant and export of power for irrigation of cane farms. These diverse operations have over a period increased the demand for self-generated energy while at the same time reducing the bagasse available for steam generation. The net consequence has been the need for increased boiler capacity and additional coal burning as an alternate fuel source with concomitant increase in the cost of energy produced. This situation has prompted management to launch an intensive investigation to minimise the energy consumed for processing. It was realised that changes to energy supply or demand must fit the requirements of the whole complex, including possible future expansion. This paper discusses the use of pinch analysis as a systematic means of defining the optimum retrofit strategy.

Background on pinch analysis

Pinch analysis has become an accepted technique for identifying optimum heat recovery opportunities and has also become established as a means of developing low cost options for eliminating bottlenecks in plants and reducing emissions. Pinch analysis has previously been applied to sugar processing and has been useful in identifying some energy saving opportunities and providing new insights into evaporator integration and vapour bleeding. These studies have also confirmed that the sugar industry is well advanced along the learning curve with regard to energy efficiency. The major benefit, however, which has not yet found extensive application in sugar manufacturing, is the use of the technology to aid in systematically optimising energy saving retrofit designs to existing plant, primarily through evaporator integration and vapour bleed configuration.

To understand this paper, a brief review of established pinch concepts and results is necessary: All 'hot' streams (streams requiring cooling) and all 'cold' streams (streams requiring heating) in a plant can be combined in terms of their heat contents in a composite curve and are plotted separately on common temperature, enthalpy axes (Figure 1). The composite curves show the overall profiles of heat availability and heat demand in the process over its entire temperature span. They resemble a counter-current picture of the process heat balance, similar to a counter-current heat exchanger. The 'overlap' of the curves determines the maximum possible heat recovery between streams in the process. On the other hand, the 'overshoots' define the minimum utility requirements. These are the energy 'targets' for both hot and cold utility. The curves are usually separated at one point by the minimum approach temperature (ΔT_{min}). Clearly, the energy targets are maximised by shifting the curves horizontally along the X axis (this is possible since enthalpy is a relative measurement) until they touch at this point. Under this condition heat transfer is not feasible because $\Delta T_{min} = 0$. Infinite area, however, means infinite cost, and a capital energy trade-off can be explored by shifting the curves apart horizontally. An increase in ΔT_{min} will give a higher energy requirement (higher energy cost), but will also provide larger driving forces (lower capital cost). An optimum value of ΔT_{min} can thus be determined. This technique is known as 'supertargeting'.

Composite curves

The point at which ΔT_{min} occurs is known as the heat recovery 'pinch'. The significance of the pinch temperature is that it divides the system into two thermodynamically separate subsystems, each of which is in enthalpy balance with its relevant utility. Above the pinch is a heat sink subsystem and the system below is a heat source. Any transfer across the pinch would be from sink to source and must inevitably lead to increased demand on both utilities. In other words, designs which use minimum utility consumption must have no heat transfer across the pinch.

This discovery led to the pinch design method which can be summarised by the following three basic rules:

- no cold utility above the pinch
- no hot utility below the pinch
- no process heat recovery across the pinch.

Between them these rules ensure maximum energy efficient designs. The 'grand composite curve' of Figure 2 represents one further stage of abstraction of the problem, and is simply constructed by subtracting the composite heat supply curve from the composite heat demand curve at each temperature. The resulting curve provides a convenient summary of the process heat balance which will be found particularly useful when addressing the optimum design of evaporators in the overall energy scheme of a sugar factory.

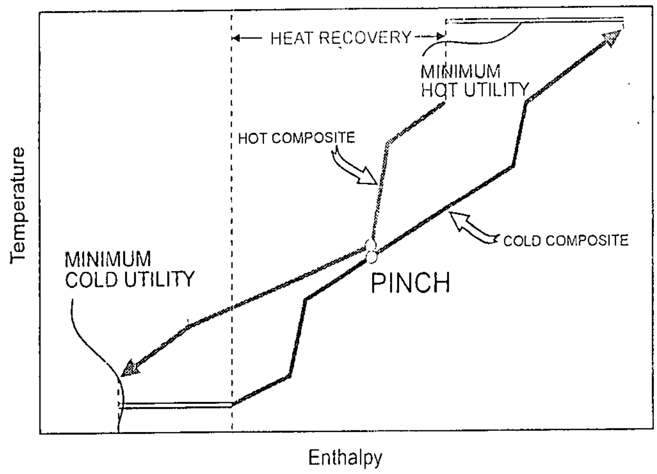

Figure 1. The composite curves.

Grand composite curve

Evaporator integration

The first task in applying a pinch analysis to a plant is that of extracting the required thermodynamic data for construction of the grand composite curve. Essentially all that is required is the identification of all the process streams undergoing an enthalpy change, i.e. all the 'hot' and 'cold' streams of the underlying process. The evaporator presents a conceptual problem in this regard since the minimum utility target is greatly affected by the particular arrangement. The required concentration of liquor can be achieved with a variety of effect and bleeding configurations, each leading to a different combination of stream energy/temperature profiles. This is illustrated by the example, where, along the lines of Rillieux's first principle, the simple addition of a single effect further reduces the energy consumption target.

The key to this dilemma may be found in another pinch technology concept, namely, that of multiple level utilities. Essentially, utility is used to satisfy the process heat balance. The grand composite curve may be viewed as a summary of this heat balance as a function of temperature and consequently

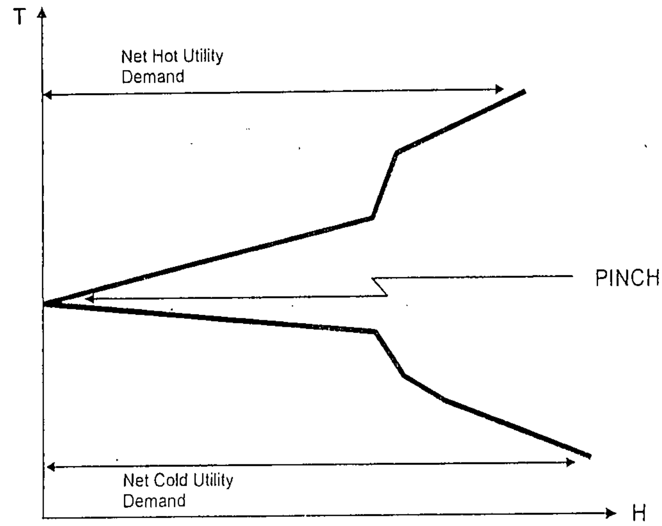

Figure 2. The grand composite curve.

indicates the amount of heat that must be supplied to, or removed from, the process at any given temperature. Thus this balance can often be satisfied with various utilities at different temperatures and the entire utility load may then be matched along the length of the grand composite curve rather than at the extreme temperature levels. To derive any benefit from having a multiple level utility set, some useful work must be achieved in degrading heat from one level to a lower one. A well known example of this is the use of a turbine between the HP and LP steam levels to produce electrical power. Since to the first approximation, an evaporator produces as much steam as it consumes, although at a lower temperature and pressure, while at the same time producing work in the form of a solid/liquid separation, a multiple effect evaporator fits the definition of a multiple level utility. If the heat flows in the evaporators are to be viewed as utilities, then they clearly must be excluded *completely* from the process stream heat requirements.

This insight step means practically that the hot and cold streams comprising the evaporator must be separated from the grand composite curve which now reduces to the heat flows of the remaining process. The evaporator system, being a utility, must now be correctly matched with the process grand composite curve to meet the requirement of the rest of the process. The evaporator can also be plotted on a temperature-enthalpy graph as a series of rectangles which represent the heat loads (supply and demand) for each effect at the relevant temperatures. If the evaporator is located across the pinch, it is taking in heat above the pinch and rejecting it below. This breaks an important rule of pinch technology: do not transfer heat across the pinch. For the most efficient use of available heat, the evaporator must be placed either entirely above or entirely below the pinch. In the sugar industry, for example, the pinch occurs at too low a temperature for placement below to be feasible. Further, by matching the evaporator temperature-enthalpy profile against the process heat sink through manipulation of the evaporator heat loads, the evaporator can be made to fit entirely within the minimum hot utility requirement. In effect, there is now a 'free' evaporator because that much energy would have been required to drive the rest of the process. *We now also have available a technique which provides both ultimate energy targets* as well as a systematic method for evaporator heat integration. Clearly, the process stream heat network comprising the remaining composite curve must be optimised separately using the pinch technique rules.

Considerable freedom exists in manipulating the evaporator system, within the constraint of the total evaporation requirement, to match the process heat sink by varying the number of effects and/or vapour bleeds. There will thus be a number of evaporator configurations that achieve the minimum energy target. Alternatively, some energy from the last effect can be deliberately rejected below the pinch in order to lower the last effect temperature and thus increase the temperature driving forces across the evaporator. In this way the capital cost of the evaporator is reduced, and the factory energy requirement is increased. There is thus a capital energy trade-off that can be explored. There are several well established pinch techniques available, such as the 'supertargeting' method for new plant and the retrofit techniques proposed by Linhoff and Tjoe for existing plant to optimise the capital/energy trade-off in heat exchange networks.

None of these techniques specifically addresses the unique situation of evaporator integration where multiple configurations are possible, each leading to different optimum values of ΔT_{min} for the overall heat exchange network. The authors have devised a method for near optimal design of evaporators based on the existing pinch techniques of 'supertargeting' described above and 'remaining problem analysis'. Essentially the technique allows the designer to simultaneously determine the optimum ΔT_{min} for the background heat exchange network and the optimum approach temperature (ΔT_{min}) between the evaporator and the background network or process heat sink, for a fixed number of effects. The latter ΔT_{min} value determines the appropriate evaporator bleed profile for the best capital/energy trade-off for the evaporator. The procedure is then repeated for different numbers of effects to determine the best overall scheme.

A detailed description of the technique is beyond the scope of this paper. It is important to note, however, that the methodology accounts for both heat exchange area and additional installation costs (e.g. piping, valves, etc) associated with each stream match as well as the cost of the integrated evaporator. The evaporator design must then be checked for sucrose inversion losses and possible increased colour formation; several iterations of the procedure may be required in order to satisfy these requirements. Once the best overall, integrated, energy design has been determined in this way, the individual energy saving projects are ranked in decreasing order of financial return to form a project path or 'road map'.

Structure of the pinch study

The starting point for a study of this nature is an accurate status quo heat and mass balance. To ensure the integrity of this information TSB obtained balances from three independent sources. These were checked for congruency, and only after discrepancies were resolved was a single balance accepted for further work.

The approach adopted in the Malelane study was as follows: the first phase commenced with the maximum energy recovery (MER) design for the factory as a whole during the crushing season. This design, which represents the minimum heat energy required for the Malelane factory, was performed as a reference or benchmark for the optimisation work to follow. Several alternative scenarios were then developed for heat recovery networks where the MER design was relaxed around certain key parameters. The second phase of analysis subjected these scenarios to customised economic investment criteria, plant operability, availability and flexibility criteria. Capital costs versus energy savings trade-offs were comprehensively explored during this phase to yield the optimum investment strategy to save heat energy at the Malelane mill.

Maximum energy recovery design

Stream data

The 'hot' and 'cold' streams identified in Appendix 1 constitute the 'background' heat exchange network for the Malelane mill.

The grand composite curve

The minimum energy requirements for hot and cold utilities to satisfy the operation of the raw mill and sugar refinery are indicated in Figure 3, the grand composite curve (GCC). This curve shows the minimum amount of heating or cooling required at each temperature level of the process.

TSB Malelane grand composite curve

A global approach temperature (ΔT_{min}) of 3°C, based on previous experience, was used in the construction of the GCC. Because larger approach temperatures are required for gas and vacuum pan heat exchange applications, additional ΔT_{min} contributions have been considered for the following applications:

A and Refined Pans	ΔT_{min} 27°C
B and C Pans	ΔT_{min} 30°C
air heating	ΔT_{min} 20°C.

The GCC reveals a pinch temperature at 65°C (referenced to hot streams), or 62°C (referenced to cold streams). The pinch is thus shown on the GCC at 63,5°C (63,5 − $\Delta T_{min}/2$ for cold streams and 63,5 + $\Delta T_{min}/2$ for hot streams).

Energy saving opportunities

The GCC of the raw mill and refinery at Malelane allows the identification of any violations of the fundamental pinch principles. These violations, when corrected, represent opportunities of saving energy, as highlighted in Table 1.

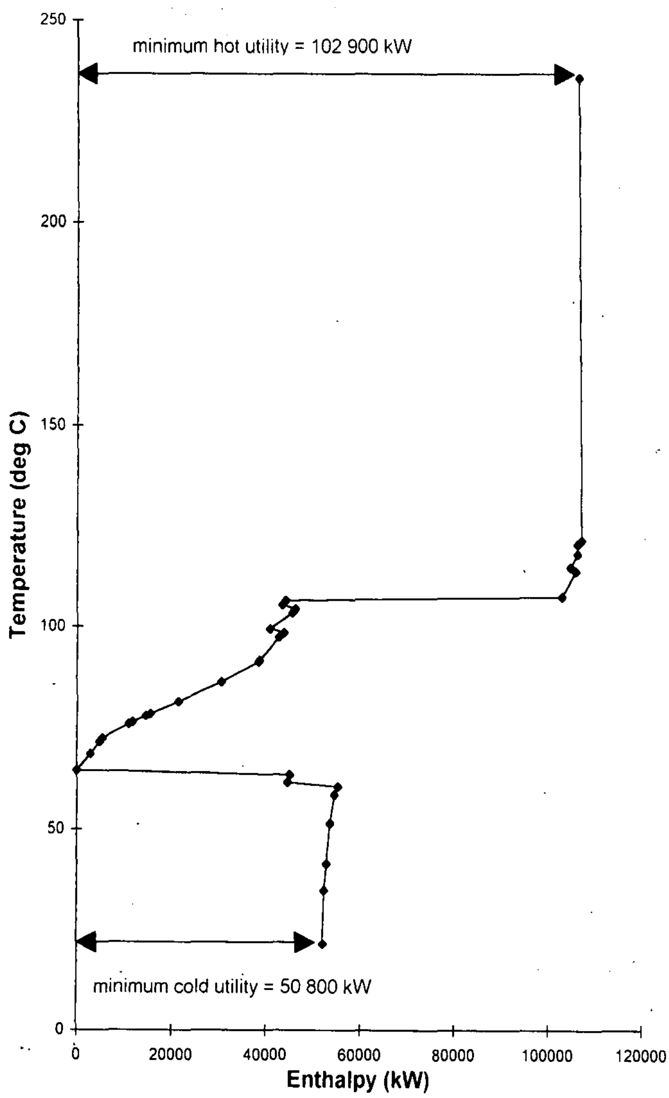

Figure 3. TSB Malelane grand composite curve.

Energy recovery opportunities

In addition to the energy saving opportunities presented by the pinch violations, the following modifications, based on good sugar factory operation, have been included in the maximum energy recovery design:

- raw evaporator syrup increased from 65 to 68° brix
- refinery melt brix increased from 65 to 68°brix
- MJ flash temperature reduced from 104 to 102°C.

The utility requirements can be read from Figure 3, as:

minimum hot utility	=	102 900 kW
minimum hot utility (including losses at 7,5%)	=	110 600 kW
minimum cold utility	=	50 800 kW.

The minimum hot utility equates to 49% steam on cane or 172 t/h HP steam requirement. A comparison can be made with the existing hot utility (including losses) of 150 300 kW indicating a potential reduction in process steam demand of 26,4%. The GCC indicates that the minimum cold utility for the raw mill and refinery at Malelane is 50 800 kW. A reduction in cooling utility of 39 700 kW (44%) would also be realised. Considering that Malelane uses river water make-up as cooling water, this potential reduction is of strategic benefit.

Evaporator integration for MER design

Within the constraint that the minimum energy target is dictated by the GCC, the evaporator must be integrated so as to minimise the overall area requirement for evaporation and process heating. For Malelane mill the best match emerges as a triple effect evaporator. The superimposed evaporator configuration is shown in Figure 4, the GCC for the MER design.

Table 1
Energy recovery opportunities.

No.	Current cases of cross-pinch heat transfer	Proposed MER solution below the pinch (<62°C)	Proposed MER solution above the pinch (>65°C)	Associated duty saved/kW
1	B and C remelt heating on exhaust steam	Heat with pan vapour	N/A	311
2	Refinery raw melt heating from 34,9 to 70°C with exhaust steam	Heat with pan vapour from 34,9 to 62°C	N/A	1 554
3	Heating of sweet water from 50 to 60°C with exhaust steam	Heat with pan vapour	N/A	402
4	Heat from boiler blow-down rejected to cooling water	N/A	Heat boiler feed water to de-aerators	1 015
5	Refinery second effect vapour rejected to cooling water	N/A	Appropriate evaporator integration	3 605
6	Raw house fourth effect vapour rejected to cooling water	N/A	Appropriate evaporator integration	22 175
7	Excess condensate to drain	N/A	Heat MJ	2 544

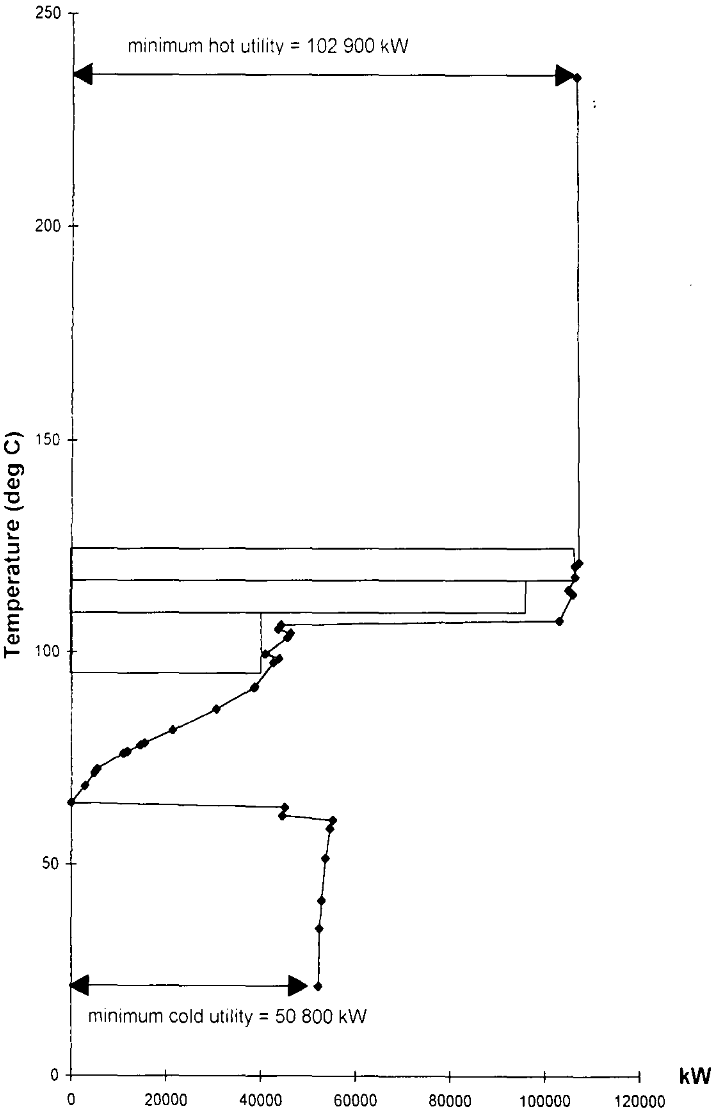

Figure 4. Integrated evaporator.

Integrated evaporator

For the MER design, the application of a quadruple or quintuple evaporator would involve additional capital cost (due to lower temperature driving forces per effect and additional complexity) with no further energy savings. The only possible benefit of operating quad or quin effects would be reduced sucrose losses or lower colour. This is discussed briefly under inversion losses and colour formation.

It must be borne in mind that the MER design does not consider the practicality of stream matches nor does it take into account capital costs of new heat exchange networks.

The evaporator integration was re-examined after the elimination of non-practical heat exchange matches, e.g. use of pan vapours for heating duty. The optimum match still emerged as a triple effect evaporator design with very attractive savings. For a cost effective evaluation to be made, this benefit must be traded-off against the effect of higher brix syrup at higher temperatures and longer retention times in the second and third effect vessels of a triple. Due to the lack of empirical data on high temperature evaporation in a cane sugar evaporator, model simulations were done to determine sucrose inversion and colour formation for typical Malelane conditions for pH and residence times. The model for sucrose inversion is based on the SMRI equation which considers the effect of time, temperature, pH and sucrose concentration, but is adapted to account for the residence time distribution occurring in either Robert, Kestner or falling film evaporators.

Sucrose inversion

The results of the model of the *status quo* evaporator at Malelane and of two relaxed energy recovery cases are presented in Table 2 by way of example. A total sucrose inversion loss of 0,231% is predicted for the *status quo* at Malelane and losses of 0,325% for a triple with additional first effect area. The loss is 0,094% higher than the *status quo*. This result may, however, be reduced to below the level of the *status quo* by judicious replacement of some existing vessels with either falling film or plate evaporators. This is highlighted when three existing Robert vessels are replaced with falling film evaporators. The inversion losses reduce to 0,209%. This value is less than the *status quo* quad.

These results have revealed that there is a complex trade-off between energy savings, sucrose inversion and capital costs. The results presented here are indicative and are not optimised.

Table 2
Sucrose inversion losses modelled on 350 TCH.

	Sucrose loss (t/h)	Sucrose loss (%)
Malelane status quo (existing quad)	0,101	0,231
Triple using existing vessels with additional first effect area	0,140	0,325
Replace three second effect Robert vessels with plate falling film evaporators in the triple above	0,090	0,209

Sucrose inversion losses modelled on 350 TCH

Colour formation

Due to the complexity of colour formation at high temperature and the interactive effect of pH, brix and time it is extremely difficult to model colour formation for different evaporator designs. To obtain some insight, a series of laboratory trials were conducted by the SMRI to determine the effect of brix, pH, temperature and time on the colour of evaporator first effect juice and syrup at 120 and 90°C, and pH values of 5 and 6. These conditions best represent the range of conditions encountered in the triple effect design. The summarised data (Table 3) have been used to predict possible conditions in a triple evaporator and the following equation was determined from multi-linear regression.

Colour = 37646 – 6,22E6/T + 18,7*brix - 447*pH + 0,732 t
T = temperature (K)
Bx = degrees brix
t = residence time (min).

An R^2 value of 0,685 was obtained for 48 sets of results. Because of the low correlation coefficient, it should be noted that the conclusions serve only as a guideline for future selection.

Table 3
Effect of brix temperature and time on colour formation.

Condition	Typical retention (min)	Colour increase (colour units)	Colour increase (%)
Low brix/high temp; typically first effect (18 brix/120°C)	5,0	1 083	5,6
*Intermediate brix/intermediate temp; typically second effect (43 brix/105°C)	7,5	794	4,1
High brix/high temp; typically third effect (68 brix/90°C)	10,0	505	2,6
Total	22,5	2 382	12,3

Effect of brix, temperature and time on colour formation

The magnitude of the colour increase calculated is lower than expected for higher temperature exposure; however, the downstream effect of the increased colour in raw and refined sugar boiling is unknown. Sound sugar technology practice predicates minimisation/avoidance of colour increase in processing to keep final sugar colours down, therefore this factor has been adopted as a constraint in the next phase of the analysis.

Design parameters

Sensitivity parameters and proposed factory modifications

The next step has been the determination of the optimum evaporator design and vapour bleeding scheme for Malelane in accordance with sensitivity parameters and proposed factory modifications as well as constraints on the energy network. These take cognisance of various production scenarios of the raw mill and refinery and in addition consider the inherent energy savings associated with process improvements (Table 4).

Sensitivity parameters and proposed factory modifications

Constraints on heat exchanger design

In addition to the above considerations, the following constraints and design parameters shown in Table 5 were imposed on heat exchanger designs.

Heat exchanger design criteria

All the above parameters ensure that adequate robustness is built into the evaporator in keeping with future capacity changes and process energy saving modifications. Incorporation of the operability and flexibility constraints as well as due regard to sensitivity parameters has shifted the evaporator match to a quadruple design. Main reasons for the shift can be attributed to:

- reduced vapour demand due to process modifications
- capital costs for a triple effect evaporator dominate the capital/energy trade-off
- relaxations for the various sensitivity and constraint parameters identified earlier.

Optimum approach temperature

The pinch analysis technique known as 'supertargeting' has been employed to determine the optimum ΔT_{min} for the maximum return on investment. The starting point for the supertargeting exercise is the calculation of heat exchanger cost curves for new equipment. The following cost curve was developed from suppliers' prices for conventional gasketed plate exchangers:

Cost (R)	=	13562 * Area ^{0,3725}
Exchange rate used R/$	=	4,42
Data range	=	10-150m^2
r^2	=	0,99.

This type of design was used for matches where scaling or clogging with fibres was not expected, i.e. combinations of vapour/water/condensate/clear liquor matches. A similar cost curve was developed for welded, fully cleanable free-flow plate exchangers:

Cost (R)	=	22697 * Area ^{0,4278}
Exchange rate used R/$	=	4,42
Data range	=	6-220m^2
r^2	=	0,98.

Applying the pinch technique for a two year pay-back, an optimum ΔT_{min} of around 3°C was confirmed.

Overall optimisation

The additional installation costs (e.g. piping, valves, etc) associated with each stream match and the capital cost of an integrated evaporator have been addressed using the pinch techniques mentioned earlier. The investment criteria applied throughout the study are given in Appendix 2. In addition to the general bases and sensitivity parameters already listed, the specific criteria given in Appendix 3 have been applied to the raw house optimisation.

Investment scenarios

Retrofit projects

The short-list of energy network design projects that lie on the optimum retrofit path is given in Table 6. These have been

Table 4
Sensitivity parameters and proposed factory modifications.

	Base case	Sensitivity parameters
General		
Cane throughput	365 TCH	Max - 385 TCH
LP steam pressure	120 kPa maximum	Optimise according to let-down margin and additional evaporator area requirement
Assumed losses	As per mass balance 7,5% exh and 5% on vapour bleeds	5% exh and 3% on vapour bleeds, due to insulation and steam trap improvements
Diffuser		
Imbibition	375%	Min 350%
Diffuser temperature	80°C	N/A
Scalding juice temperature	85°C	N/A
Juice heating		
MJ temperature	Heat to 99°C	Heat to 95°C and vacuum flash
CJ final temperature	Heat to 115°C	Heat to first effect bubble point temperature
Raw house pans		
Movement juice	7,5% on massecuite vol	N/A
B sugar affination	None	6% reduction in A and B pan steam
A/B continuous pan	None	Required after major ref. expansion
A crystalliser cooling	None at present	0,75 t/h raw pan steam reduction
Raw syrup brix	68°	70°
Refinery		
Production	As per existing	Base case +20%, base case +50%, base case +100%
Pan brixes	Optimise with automation	None
Melt brix	66°	68°
Thick liquor brix	72°	74° (with temperature control)
Pan movement water	80% reduction in current use (pan automation)	Maximum 90% reduction
Fifth boiling	None	1,3 t/h reduction at 43 t/h refined sugar output

Table 5
Heat exchanger design criteria.

Heat exchanger network	Constraint
Press water heating	Only direct contact heaters allowed due to fouling.
SJ, MJ, CJ heating	Only tubular or platular heat exchangers to be used. Mixed juice screening to be provided if platular heaters are used.
Minimum pan calandria vapour temperature	104°C for batch pan heating.
Fine liquor pre-heating	Gasketed PHE acceptable.
Carbonatation heating	Platular heat exchanger for screened melt.
Sweet water heating	Only platular heaters to be used or direct injection steam.

screened to meet the criteria for sound technology and economic viability. In addition, since the economics of individual projects are affected by the inclusion or exclusion of others, the integration of the best set of projects has had to be systematically analysed. It is impossible to describe in this report the details of all the complex trade-offs and the optimisation methodology used; rather the results of the optimum investment retrofit are presented.

List of retrofit projects

Improvements brought about in energy reduction by process modifications and the effects of the refinery expansion have being grouped into a logical progression route to form several cases shown in Table 7.

Case descriptions for retrofit scenarios

Case 1 considers the implementation of energy savings projects identified from the pinch recovery opportunities. Cases 2 to 5 include process modifications as well as progressively increased refining rates. The project and process modifications associated with each case are shown in Table 7.

Evaporator configuration for retrofit scenarios

The evaporator configurations for the above scenarios have been summarised in Table 8.

Evaporator configuration for retrofit scenarios

Conservative heat transfer coefficients have been used to determine evaporator areas for the different effects. Cases 1 to 4 can be accommodated with the existing evaporator area although a

Table 6
List of retrofit projects.

Project No.	1 & 2	7a	8	5	32	12 & 17
Description	SJ heating	MJ heating (3 stage heating)	CJ heating (2 stage heating)	Press water heating	BFW heating	Raw pan heating
Hot stream	V3	V3 / V2 / V1	V1 / Exh	V3	Boiler blow-down	V2
Supply temp	86°C	86/100,7/109°C	100,9/123,7°C	86°C	237°C	99°C
Target temp	86°C	86/100,7/109°C	100,9/123,7°C	86°C	85°C	99°C
Cold stream	SJ	MJ	CJ	Press water	BFW make-up	A, B, C cont. pans
Supply temp	66°C	62,5°C	96°C	58,4°C	75°C	N/A
Target temp	75,9°C	99°C	109°C	82°C	79,8°C	N/A
kW exchanged	7 744	20 492	6 578	5 218	1 120	30 955
Heat exchanger	Existing S&T	Existing S&T	Existing S&T	New	New	New

1. Project numbers refer to the original study, which considered more than 70 potential energy saving projects.
2. The temperatures and duties supplied above are typical values and these might differ slightly from case to case. However, the validity of where each project is placed on the optimum retrofit is not affected.

Table 7
Case descriptions for retrofit scenarios (* = refined sugar output, # = process modifications). Higher RSO rates have been considered as part of the sensitivity analysis.

Case	Basis			
1	Proj 1&2, 7a, 8, 5, 32	+b/case RSO		
2	Proj 1&2, 7a, 8, 5, 32	+b/case RSO		+#1,2,3,4
3	Proj 1&2, 7a, 8, 5, 32	+b/case RSO	+ 20%	+#1,2,3,4,5
4	Proj 1&2, 7a, 8, 5, 32	+b/case RSO	+ 50%	+#1,2,3,4,5,6
5	Proj 1&2, 7a, 8, 5, 32, 12, 17	+b/case RSO*	+100%	+#1,2,3,4,5,6

#1 = A-crystalliser cooling
#2 = 90% reduction in refinery pan movement water usage
#3 = Reduced steam losses (exhaust from 5 to 3%, vapour from 7 to 5%)
#4 = Reduced MJ heating from 102 to 95°C and flash under vacuum
#5 = Refinery energy saving steps (melt bx 68, thick liquor bx 72°, 5th boiling)
#6 = B sugar affination

shift of 740 m^2 from the second to the third effect is required to accommodate the change from vapour 2 to vapour 3 for projects (1, 2, 5, 7a).

It is only in case 5, because of the increased steam demand from the higher refining rate, that larger evaporator area is required. The proposed strategy to address this entails using projects 12 and 17, i.e. a change from batch to continuous raw pans and a corresponding vapour bleed change from vapour 1 to vapour 2 for these pans. It is only when 'A' and 'B' continuous pans are installed that lower pressure vapour can be used. The evaporator remains a quad; however, new second effect surface, larger third effect area by incorporating the existing fourth effect area and new, albeit small, fourth effect area would be required. Due to colour formation problems in effect three for this case, the alternative option is for the entire third effect to be a new short retention time type evaporator. In this case the existing 1 500 m^2 third effect may be used in second effect service to make up the shortfall in area and the fourth effect may remain as is, even though it is vastly oversized and is acting primarily as a flash vessel.

Table 8
Evaporator configurations for retrofit scenarios (FF = falling film).

Effect	Vessel	Case 1, 2, 3, 4	Case 5
1	SK1, SK2, 1A	SK1, SK2, 1A	SK1, SK2, 1A
2	2A, 2B, 2C, 2D, 2E	2A, 2B, 2C, 2E	2A, 2B, 2C, 2D, 2E + new FF tubular evap . (400 m^2)
3	3	2D, 3	3, 4
4	4	4	New FF tubular evap . (100 m^2)

SK1	Semi-Kestner No. 1	1 580 m^2
SK2	Semi-Kestner No. 2	2 325 m^2
1A, 2A, 2B, 2C	Robert vessels	1 020 m^2 each
2D, 2E	Robert vessels	740 m^2 each
3	Robert vessel	1 500 m^2
4	Robert vessel	1 500 m^2

The investment return for energy savings obtained from V2 bleeding to batch pans for cases 1 to 4 was found to have poor pay back (>4 years). The reason is that V2 is required to be at least 104°C to provide sufficient temperature driving force for good batch pan evaporation rates. The higher temperature reduces the temperature driving forces for evaporation across the first two effects of the raw factory evaporator, resulting in an increase in the evaporator area requirement. This is further exacerbated by the significantly increased sucrose inversion that will occur due to higher first and second effect temperatures. Existing conventional evaporators in certain areas would therefore have to be replaced by low residence time evaporator technology to counter the sucrose destruction. Under this scenario the increased capital greatly outweighs the energy saving benefit.

Economic summary for various scenarios

The summary of optimum energy saving scenarios for the five strategic processing cases is shown in Table 9.

Table 9
Economic summary for various scenarios.

(Existing HP steam on cane: 67%)	Case				
	1	2	3	4	5
HP steam % cane	56,8	54,2	54,4	54,4	56,2
Pay-back years	1,11	1,20	1,26	1,20	1,19

Pay-back excludes cost and benefit of process modifications.
Pay-back costs based on budget prices for supply and installation.

Economic summary for various scenarios

The best energy saving scenario is realised by reconfiguring the evaporator bleed to include the projects and process modifications identified in cases 1 to 5. The order of the cases provide a project road-map for progressive implementation to cater for future possible expansion of the refinery.

Implementation of case 1 essentially retains a quad evaporator with reconfigured vapour bleed and achieves reduction in HP steam % cane from 67 to 56,8%. This is reduced to 54,2% with the addition of the first four process modifications in case 2. In cases 3 and 4 the increased steam demand for 20 and 50% increase in RSO is negated with the addition of process modifications five and six respectively. Hence the steam % cane for cases 3 and 4 remains at 54,4% with the existing evaporator configuration. Notwithstanding the improvements in evaporator and vapour bleed configuration for 100% increase in RSO (case 5) the steam % cane increases slightly to 56,8%.

The pay-back periods in all five cases is approximately 1,2 years and is within the two year requirement of TSB.

Conclusion

The application of pinch technology at Malelane has not necessarily resulted in the identification of new energy saving opportunities. Rather, the techniques employed have guided the retrofit in a systematic manner and provided confidence that the best project path has been selected. The analysis has quickly determined the optimum investment strategy from over 70 potential energy saving projects without having to address each one separately.

Capital, energy and technology trade-offs have confirmed that the existing quad design evaporator is optimum for the Malelane mill for current as well as future processing requirements. Vapour bleed reconfiguration with existing and some new heat exchange equipment has identified a potential steam consumption reduction of approximately 18% with a capital repayment period of 1,2 years.

Acknowledgements

The authors are grateful to TSB for permission to present this paper. Particular thanks are due to members of staff of the SMRI and TSB who collaborated on this project.

BIBLIOGRAPHY

Jenkins, DM and Kumana, JD (1987). Pinch technology applications in sugar processing. SIT Congress: Paper 561.

Linhoff, B (1994).Use of pinch analysis to knock down capital costs and emissions. *Chem Eng Progress* (August, 1994) pp 33-57.

Markham, S (1996). The effect of brix, pH, temperature and time on colour formation. Sugar Milling Research Institute Technical Note No. 56/96: 1-10.

Radford, D (1996). Heat and mass balance for TSB Malelane Mill. Report on energy study conducted at Malelane Mill.

Riley, R and Seillier, D (1996). Phase 1 Pinch Analysis TSB Malelane Mill. Report on energy study conducted at Malelane Mill.

Riley, R and Seillier, D (1996). Phase 2 Pinch Analysis TSB Malelane Mill. Report on energy study conducted at Malelane Mill.

APPENDIX 1
Data set (MER design).

Stream description	T1 (°C)	T2 (°C)	Cp (kJ/kg °C)	Flow rate (t/h)	Latent heat (kJ/kg)	Enthalpy (kW)
Hot streams						
Clarification flash	100,0	100,0		1,63		1 022
A pan vapour	65,0	65,0		38,50	2 256,6	23 232
B pan vapour	62,0	62,0		10,95	2 386,1	7 276
C pan vapour	62,0	62,0		5,60	2 392,0	3 736
First boiling vapour	65,0	65,0		21,61	2 401,8	13 689
Second boiling vapour	65,0	65,0		10,01	2 361,6	6 567
Third boiling vapour	65,0	65,0		5,27	2 361,6	3 457
Fourth boiling vapour	65,0	65,0		2,92	2 361,6	1 916
Excess condensate	100,0	65,0	4,187	62,50	2 361,6	2 544
Combined condensate flash	100,0	100,0		4,96		3 109
Boiler blow-down	237,0	80,0	4,187	5,30	2 256,6	968
Exh flash to V1	115,3	115,3		2,51		1 549
V1 flash to V2	106,0	106,0		5,39	2 221,6	3 355
Raw evaps syrup	93,3	70,0	2,910	75,92	2 240,5	1 430
Cold streams						
Scalding juice	71,0	90,0	4,000	831,10		-17 545
D1 steam to diffuser	106,0	106,0		0,55	2 235,1	-341
Press water to diffuser	57,0	85,0	4,187	172,80		-5 627
Mixed juice	62,0	102,0	3,980	461,60		-20 413
Clear juice	96,0	116,5	4,000	413,30		-9 414
A pans	106,0	106,0		39,44	2 235,1	-21 638
B pans	106,0	106,0		10,99	2 235,1	-6 823
C pans	106,0	106,0		5,42	2 235,1	-3 365
Remelt (B and C)	40,0	60,0	2,700	20,73		-311
Cents and mol blow-up	120,0	120,0		1,52	2 202,0	-930
Raw melt	33,5	70,0	2,740	70,17		-1 949
Carb liquor	70,0	70,0	2,880	73,57		-883
Fine liquor	77,0	85,0	2,920	74,11		-1 142
First boiling	106,0	96,0		24,55	2 235,1	-13 018
Second boiling	106,0	106,0		11,43	2 235,1	-7 096
Third boiling	106,0	106,0		6,01	2 235,1	-3 831
Fourth boiling	106,0	106,0		3,31	2 235,1	-2 055
Refined sugar drier air	20,0	90,0	1,100	55,00		-1 176
BFW	75,0	106,0	4,187	181,60		-6 548
Carb liquor re-heat	76,5	80,0	2,890	74,11		-208
Carb liquor ex filt re-heat	74,5	80,0	2,890	74,11		-327
Sweet water	50,0	60,0	4,140	34,96		-402

APPENDIX 2
Investment calculation criteria.

The bases for all heat energy saving calculations are:

- Overall boiler efficiency (kW_{in}/kW_{out}) = 75%
- Overall time efficiency = 84%
- Coal cost (delivered) = R115/ton
- Coal CV = 28 000 kJ/kg
- Pay back period required = 2 years
- Coal price inflation = 6% per annum
- No additional operating costs
- Capital cost includes 20% contingency
- Season length = 40 weeks

APPENDIX 3
Heat exchanger design criteria.

- HTCs for shell and tube HEs derived from Hugot.
- Cleaning cycle for shell and tube HEs optimised using pinch techniques; fouling rates determined from Hugot.
- All energy network designs based on cane throughput of 385 TCH, but savings were calculated on 365 TCH in order to add a conservative bias.