A STRUCTURED APPROACH TO SUGAR FACTORY DESIGN

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Abstract

Over the years, a number of computer models have been developed and refined by Tongaat-Hulett Sugar for use in sugar factory design. These models are used for greenfield projects, factory expansions, rehabilitations, de-bottlenecking and the integration of new processes such as back-end refining, ethanol production and co-generation. Based on practical experience, a structured approach has been developed that integrates and orders all of the steps necessary for an effective and comprehensive design procedure. The steps include agreement on basic design data, cane supply and throughput strategy, selection of process technology, a boiling house balance, an overall steam and energy balance, detailed evaporation plant and juice heating modelling, equipment sizing and plant layout. Project requirements generally have a number of interdependent objectives that need to be satisfied and this leads to the design procedure involving a number of iterative steps, as the basic design is refined and improved, which are catered for in the structured approach. The paper discusses the design procedures, the models used in the process and the current structured approach that is used by Tongaat-Hulett Sugar.

Keywords: factory design, modelling, mass balance, energy balance, equipment design

Introduction

The effective design of a sugar factory requires a properly structured and comprehensive process design procedure. Because of the complex interactions between the various sections of a sugar factory, the design procedure cannot be a simple 'once-through' linear process. Rather, to accommodate these interactions, the procedure needs to be to be iterative and requires the careful use of engineering judgment to make appropriate assumptions that avoid excessive reiteration of the design.

A well thought-out design, finalised as early as possible in the project, is critical in ensuring that a project meets the standard requirements of cost, performance and timing.

Agreement on basic design data

The design must be based on a set of process data that is agreed to by all relevant parties. This set of data will include:

- *Total quantity of cane to be crushed in a season and length of milling season (LOMS) for crushing the cane.* Obviously the total cane crush requirement, allied with the number of weeks available in the season and the projected overall time efficiency, will determine the weekly, and therefore also daily and hourly crush rates required.

- *Variability in cane quality over the milling season.* Variability in cane quality through the season results in differing constraints occurring over the season. Brix, non-pol and fibre constraints all come into effect at different times of the season. Accurate information on cane quality is vital to the correct sizing of plant equipment.

- *Required process performance of the factory (sugar quality, sugar recovery, energy utilisation).* Whilst the basic sugar process remains the same, differing performance requirements will result in subtle (and sometimes not so subtle) changes to the process. Low steam % cane requirements, for example, will probably result in large amounts of vapour bleeding at the evaporator station, with resultant increased heating surface areas down the tail. Low sugar colour requirements may require differing magma systems or even 'double-curing' installations.

- *Future expansion plans.* It is imperative that future expansion plans are communicated to the designers of the plant. Cognisance of these requirements will often result in the design of plant the role of which may change in the future. This can result in the installation of slightly oversized equipment in the short term, where the duty will change at a later expansion. For example, a slightly oversized second effect evaporator may become a first effect in a later expansion.

Throughput strategy

Using the agreed process data, a throughput strategy will be developed. This strategy adjusts the cane input to the factory (on a monthly basis) to take account of the varying cane quality. Typically, at different times of the season, the throughput of a sugar factory will be constrained, either by a maximum fibre rate (determined by the sizing of the extraction plant), a maximum sucrose rate (determined by the sizing of the high grade portion of the boiling house) or by a maximum non-sucrose rate (determined by the sizing of the low grade portion of the boiling house). In developing a cost-effective process design for a greenfields facility, the factory throughput should ideally be limited for approximately equal periods of the season by each one of these constraints. This objective is achieved by appropriate sizing of the relevant factory equipment. For the expansion of an existing mill, the capacity of the existing installed equipment will determine how closely this goal may be approached.

The overall throughput strategy, and the proposed month-by-month crush budget which is the result of this strategy, forms the basis for formulating the factory mass and energy balances and prevents the over-sizing of certain portions of the plant to cope with peak loadings that are well above the average loading and of short duration.

Selection of process technology

Before proceeding with the development of mass and energy balances, there must be agreement on the major process technologies to be used in the factory (e.g. the use of either milling or diffusion for extraction, the use of continuous or batch pans in the boiling house, or the choice of decolourisation processes to be implemented in the back-end refinery). Once this has been agreed, an overall process flow sheet for the entire factory can be drawn up.

Mass and energy balances

Using the overall process flow sheet, it is then possible to develop detailed process flow diagrams for individual items of plant. These process flow diagrams form the basis for mass and energy balances that are used to determine the flow rates and characteristics (compositions) of all the major streams.

Because of the complexity of the sugar factory and the need to iterate designs in terms of process performance and equipment requirements, it is necessary to use a number of separate mass and energy balances of varying levels of detail and sophistication to achieve the final design. In general, balances that cover larger areas of the plant will include less detail, and balances specific to smaller areas of the plant will be used to confirm the details of the plant in that area.

Major types of balance used are discussed below. Note that, for consistency, all balances presented refer to a recent expansion exercise at a Tongaat-Hulett mill in Moçambique.

Dual Mass and Colour Balance

This model generates factory balances for sucrose, non-sucrose and colour, allowing a detailed investigation into both the selection and operation of the boiling scheme.

The model used calculates two simultaneous mass-balances, which allows various scenarios to be directly compared; for example, a mass-balance at the peak brix period of the season can be compared to a mass-balance for the peak non-sucrose loading period. Alternatively, the effects of improved performance – such as increased A-exhaustion – can easily be compared in a single model.

The model is based on the traditional three boiling system (A, B and C massecuites); however, any combination of magma systems can be employed (B-magma to A-seed, C-magma to B-seed, or both, or no magma usage at all, with graining for all three boil types).

An annexed (or back-end) refinery can also be modelled, with between one and four white boilings. All or part of the raw sugar produced in the raw-house section of the model can be 'dispatched' to the refinery. Raw sugar imports from other factories can also be accommodated.

The model calculates expected equipment sizing of major plant items. Examples of these major items are pans (vessel capacities in m^3), crystallisers (vessel capacities in m^3), centrifugals (massecuite handling capabilities in m^3 massecuite per hour), tanks (m^3), remelters (m^3), and sugar drying capacity. The capacity calculations are obtained using standard norms for equipment sizing. For example, in calculating pan capacities, empirically determined pan time indices (Archibald and Smith, 1975) are used. Crystalliser volumes are obtained using predefined retention times.

In greenfield designs, preliminary sizings for major items are thus immediately generated. For expansion exercises, the sizes of currently installed plant are input into the model. The model then compares installed plant capacities with the expected capacities after the expansion, allowing problem areas to be easily identified.

Examples of the outputs of the model are presented in Figures 1 to 3.

Figure 1 shows the monthly brix, non-sucrose and fibre loadings at the given cane quality. The limiting of the throughput for the brix, fibre and non-sucrose constraints can be seen.

MASS BALANCE CALCULATIONS		Ver	1,00	6,10
Case	XINAVANE	Monday, 21 March 2005		
	Cane quality estimated, XN 2002 and SM 2002			

Assumptions

Tons cane / season (factors)	403 753	
Tons cane / season (target/wght)	605 399	1 006 918
Season length weeks (factors)	30,1	
Season length weeks (target)	30,0	30,0
Start date	early April	
Stop date	early December	
Overall time efficiency (factors)	74,07	
Overall time efficiency (target)	80,00	80,00
Lost time % available	8,00	8,00
Cake loss % in MJ	0,48	0,48
UDL % in MJ to Syrup	1,25	1,25
Total UDL % in MJ	2,50	2,50
MJ-cane (not DAC) pty	0,57	0,57
Suspended solids % MJ	1,16	1,16
Extraction % (sucrose)	94,0	94,0
Imbibition % fibre	[illegible]	[illegible]
Fibre % bagasse	[illegible]	[illegible]
Filter wash index	[illegible]	[illegible]
Syrup brix	[illegible]	[illegible]
Non-Sucrose Recovery	0,87	0,87
Mol Monosaccharide/Ash Ratio	1,10	1,10
C massecuite crystal content	29,3	29,3
C massecuite brix	95,8	95,8
Molasses TPD	5,0	5,0
Sugar Pol	98,88	98,88

Two simultaneous calculations

Calculated quantities

Final Molasses Purity	38,5	38,5
Molasses Loss % in MJ	10,36	10,36
Overall Recovery	81,46	81,46
Tons cane per hour (factors)	108	108
Tons cane per hour (target)	150	250
Molasses % Cane	4,2	4,2
Cane / Sugar Ratio	9,34	9,34
Tons cane / season	605 399	1 006 918
Tons cane / week ave	20 180	33 564
Tons raw sugar/season (tgt)	64 838	107 841
Tons molasses/season (tgt)	25 542	42 482

Monthly thru puts	Max	Ave	Max	Ave
Evaporation	138	129	225	215
TFH in bagasse / hr	20,2	18,5	33,1	250,4
m³ C Mcte / hr	6,6	6,5	11,5	10,9
Ton MJ brix/hr	23,0	21,8	37,7	36,2
Ton MJ nonpol/hr	3,4	3,3	6,0	5,6
Ton cane fibre/hr	22,1	20,3	36,3	33,8
Tons cane per hour	160	150	261	250

Balance 1: Phase 1a	Tons cane: 605 399
Balance 2: Phase 2	Tons cane: 1 006 918

Input quality?	Y	Fibre %cane	Sucrose %cane	Purity %cane	TCH	TCH	MJ/hr Ton brix Adjusted	MJ/hr Ton brix Adjusted	TFH Cane	TFH Cane	TFH Bagasse	TFH Bagasse	MJ/hr Ton NP	MJ/hr Ton NP
Manual input														
Months	Wks	13,5	13,0	84,0										
April	2,3	15[illegible]	[illegible]	[illegible]	120	212	15,2	26,9	18,3	32,2	16,8	29,6	3,4	6,0
May	3,4	14[illegible]	[illegible]	[illegible]	133	234	17,7	31,2	18,8	33,1	17,1	30,2	3,4	6,0
June	4,4	12[illegible]	[illegible]	[illegible]	155	256	22,8	37,7	19,6	32,3	17,7	29,2	3,4	5,6
July	4,0	12[illegible]	[illegible]	[illegible]	157	257	23,0	37,7	20,2	33,0	18,2	29,8	3,3	5,5
August	4,4	13,9	13,1	84,9	160	261	23,0	37,7	22,1	36,3	20,2	33,1	3,3	5,5
September	4,8	13,7	13,4	85,4	157	256	23,0	37,7	21,4	35,0	19,5	31,9	3,2	5,3
October	3,7	13,6	13,4	85,2	157	256	23,0	37,7	21[illegible]	[illegible]	[illegible]	[illegible]	3,3	5,4
November	2,5	12,8	14,0	84,1	143	242	22,2	37,7	18[illegible]	[illegible]	[illegible]	[illegible]	3,4	5,8
December	0,6	14,1	13,8	84,4	149	247	22,6	37,7	21[illegible]	[illegible]	[illegible]	[illegible]	3,4	5,7
January	0,0	13,5	13,0	84,0	0	0	0,0	0,0	0,0	0,0	-1,8	0,0	0,0	0,0
	30,00	13,6	13,0	84,0	150	250	21,8	36,2	20,3	33,8	18,5	30,8	3,3	5,6
Maximum obtained					160	261	23,0	37,7	22,1	36,3			3,4	6,0
Maximum allowed							23,0	37,7	22,1	36,3			3,4	6,0
Maximum from factors					131	131	18,6	18,6	20,9	20,9			3,6	3,6

Monthly crush up to constraint

Input constraints

Figure 1. Dual Mass and Colour Balance – monthly constraint calculations.

Figure 2 depicts a section of the basic factory mass balance showing the two side-by-side mass balances, one for the peak brix period for the proposed design and the second showing the conditions for a planned future expansion of the factory (also at the peak brix loading period).

Figure 3 shows the calculated sizes of some of the major equipment.

The Dual Mass and Colour Balance model is quite flexible, and various processing subtleties can be handled. Examples of these are:

- Mud recycle to the diffuser station
- Double curing of C-massecuites
- Dispatch of B-sugar (either all or part) with the A-sugar
- A-sugar affination within the rawhouse.

These options have been built into the model and are easily activated through the setting of ‘flags’.

Dual Boiling House Balances - Rawhouse Three Boiling System														
XINAVANE					Balance 1: XN Phase 1a			150tch 1				Thursday, 24 March 2005		6,10
Cane quality estimated, XN 2002 and SM 2002					Balance 2: XN Phase 2			250 tch 3			Ver 1,00			
Calculate Suc/DS? N	150tch	250 tch	150tch	250 tch	150tch	250 tch		150tch	250 tch		150tch	250 tch	150tch	250 tch
CANE/JUICES	Cane (Suc)		MJ (Suc) Adjust		Filtrate			Filter Cake			CJ (Pol)		Syrup (Pol)	
Tons per hour	157	256	163	267	15	24		6	10		166	272	36	58
m^3/h			156	255	15	24					165	269	28	45
Brix %	15,7	[illegible]	14,1	14,1	9,9	9,9					13,7	13,7	63,6	63,6
Brix t/h	24,6	[illegible]	23,0	37,7	1,5	2,4					22,8	37,3	22,7	37,1
Inversion loss											0,2	0,2	0,2	0,2
Undetermined loss													0,1	0,1
Sucrose t/h	21,0	34,4	19,7	32,3										
Pol t/h			19,7	32,3				0,1	0,2		19,6	32,1	19,5	32,0
Purity rise			0,6	0,6							0,1	0,1	0,1	0,1
Purity (Sucrose/Brix)	85,2	85,2	85,8	85,8										
Purity (Pol/Brix)	85,2	85,2	85,8	85,8							85,9	85,9	86,0	86,0
Sucrose %	13,4	13,4	12,1	12,1				1,7	1,7					
Pol/Sucrose ratio			1,000	1,000										
Non-suc t/h	3,6	6,0	3,3	5,4										
Non-pol t/h			3,3	5,4							3,2	5,3	3,2	5,2
Colour			25 000	25 000							24 500	24 500	25 725	25 725
Colour gain											-2	-2	5	5
% Cane					9,5	9,5	8,8	4,0	4,0	8,6				
Extraction	94,0	94,0												
Water added / lost (tons)					3	6							-127	-208
Mud Recycle ? / Filter wash index					N	N		103	103					
SEEDS/MAGMAS/REMELT	B Magma to A's				B Seed			C Seed			-		Remelt	
Magma for seed footing? (Y or N)					N	N		N	N					
Bx[DS] magma % massecuite	5,0	5,0												
Massecuite/seed ratio	20,0	20,0	3,5	3,5	3,0	3,0		2,5	2,5					
Brix %	85,0	85,0	90,0	90,0	90,0	90,0		90,0	90,0				65,0	65,0
Brix t/h	1,6	2,6	8,9	14,6	3,9	6,4		3,1	5,1				7,1	11,6
Pol t/h	1,4	2,4	7,8	12,8	2,8	4,5		2,0	3,3				6,3	10,3
Purity target (Pol/Brix)					71,0	71,0		63,8	63,8					
Purity obtained (Pol/Brix)	92,0	92,0	87,8	87,8	71,0	71,0		63,8	63,8				88,9	88,9
m^3/h	1,3	2,1	6,8	11,1	3,0	4,9		2,4	3,9				8,5	13,9
Product t/h	1,8	3,0	9,9	16,2	4,3	7,1		3,5	5,7				10,9	17,8
Crystal content	70,0	70,0	45,0	45,0	45,0	45,0		40,0	40,0				30,0	30,0
Crystal growth rate (um/hr)			100	100	35	35		25	25					
Crystal size (um)	156	156	426	426	112	112		118	118					
t/h syrup			11,6	19,0	0,0	0,0		0,0	0,0					
t/h A mol			-	-	4,9	8,0		2,6	4,3					
t/h B mol			-	-	-	-		1,3	2,1					
t brix/h syr			7,4	12,1	0,0	0,0		0,0	0,0					
t brix/h A mol			-	-	3,9	6,4		2,1	3,4					
t brix/h B mol			-	-	-	-		1,0	1,7					
Colour	25 720	25 720	25 724	25 724									39 415	39 415
Magma-aff%whole colour	7,0	7,0												
MASSECUITES			A Massecuite		B Massecuite			C Massecuite			-		Return Syrup	
Double curing? (Y or N)			N	N	N	N		N	N					
Brix %			91,3	91,3	94,1	94,1		95,8	95,8				75,0	75,0
Brix t/h			31,3	51,2	11,7	19,1		7,8	12,8				0,0	0,0
Inversion loss			0,1	0,1	0,1	0,1		0,1	0,1					
Undetermined loss			0,2	0,2	0,2	0,2		0,2	0,2					
Pol t/h			27,2	44,5	8,3	13,5		4,3	7,1				0,0	0,0
Purity (Pol/Brix)			86,9	86,9	70,9	70,9		55,0	55,0				0,0	0,0
m^3/h			23,3	38,2	8,3	13,7		5,4	8,9				0,0	0,0
m^3/t Solids in Syrup			1,0	1,0	0,4	0,4		0,2	0,2					
Product t/h			34,2	56,0	12,4	20,3		8,2	13,4				0,0	0,0
Exhaustion			63,2	63,2	60,3	60,3		46,4	46,4					
m^3/t NP in Syrup								1,7	1,7					
Crystal % Tot Mass			50,2	50,2	40,2	40,2		24,4	24,4					
Crystal size			670	670	156	156		135	135					
Crystal deposition rate			253	253	77	77		17	17					
Water, t/h			3,0	4,9	0,7	1,2		0,3	0,6					
Colour			31 078	31 078	93 637	93 637		173 060	173 062				0	0
Colour increase on boiling %			8,0	8,0	15,0	15,0		25,0	25,0					
Colour to sugar			3,8	3,8	27,5	27,5		35,5	35,5					
Ton Bx % ton syr bx													0,0	0,0
SUGAR / MELT to REF	Crystal Product		A Sugar		B Sugar			C Sugar			-		Melt to Refinery	
Brix %	99,9	99,9	99,9	99,9	97,0	97,0		96,0	96,0				65,0	65,0
Pol t/h	17,4	28,4	17,4	28,4	5,4	8,9		2,3	3,7				0,0	0,0
Brix t/h	17,5	28,6	17,5	28,6	5,9	9,7		2,7	4,4				0,0	0,0
Pol	99,4	99,4	99,4	99,4	89,2	89,2		80,6	80,6				0,0	0,0
Purity (Pol/Brix)	99,5	99,5	99,5	99,5	92,0	92,0		84,0	84,0				0,0	0,0
Moisture	0,13	0,13	0,1	0,1										
Product t/h	17,5	28,6	17,5	28,6	6,1	10,0		2,8	4,6				0,0	0,0
Colour	1194	1194	1 194	1 194	25 720	25 720		61 440	61 441				1194	1194
Safety factor	0,22	0,22												
Water, t/h	Includes magma colour occluded in crystal		0,02	0,04	0,18	0,30		0,11	0,18					
Dispatch as product %			100,0	100,0	0,0	0,0								
MOLASSES			A Molasses		B Molasses			C Molasses			-			
Brix %			80,0	80,0	80,0	80,0		84,0	84,0					
Brix t/h			13,8	22,6	5,8	9,4		5,1	8,4					
Purity (Pol/Brix)			71,0	71,0	49,1	49,1		39,6	39,6					
Pol t/h			9,8	16,0	2,8	4,6		2,0	3,3					
Pol/Sucrose ratio								1,000	1,000					
Sucrose t/hr								2,0	3,3					
Purity (Sucrose/Brix)								39,59	39,59					
Product t/h			17,3	28,2	7,2	11,8		6,1	10,0					
m^3/h			12,5	20,4	5,2	8,5		4,3	7,1					
Colour			68 870	68 870	163 523	163 523		231 973	231 977					
Water, t/h (by diff.)			3,0	4,8	0,5	0,9		0,2	0,4					
Molasses Solids%Mass			82,4	82,4	91,3	91,3		95,7	95,7					
Molasse Factor								0,62	0,62					
Molasses % cane			11,0	11,0	4,6	4,6		3,9	3,9					
EVAPORATION (pans and centrifugals)			A Station		B Station			C Station			Total Panfloor			
Feed Solids% (Brix)			67,3	67,3	80,0	80,0		80,0	80,0					
Feed t/h			46,4	76,0	14,6	23,9		9,8	16,0					
Evaporation t/h (pans - no water)			12,2	19,9	2,2	3,6		1,6	2,6		16,0	26,2		
Pan factor			1,3	1,3	1,3	1,3		1,3	1,3					
Total t/h (pans and centrifugals)			15,8	25,9	2,8	4,7		2,1	3,4		20,8	34,0		

Sucrose Balance		
Total factory balance		
Bag	6,0	6,0
FC	0,5	0,5
Mols	9,7	9,7
UDL	1,06	1,06
OR	82,8	82,8
BHR	88,0	88,0
Ext	94,0	94,0
Split refinery balance		
Bag	0,0	0,0
FC	0,0	0,0
Mols	0,0	0,0
UDL	0,0	0,0
OR	0,0	0,0
BHR	0,0	0,0
Ext	0,0	0,0

Figure 2. Dual Mass and Colour Balance – factory mass and colour balance calculations.

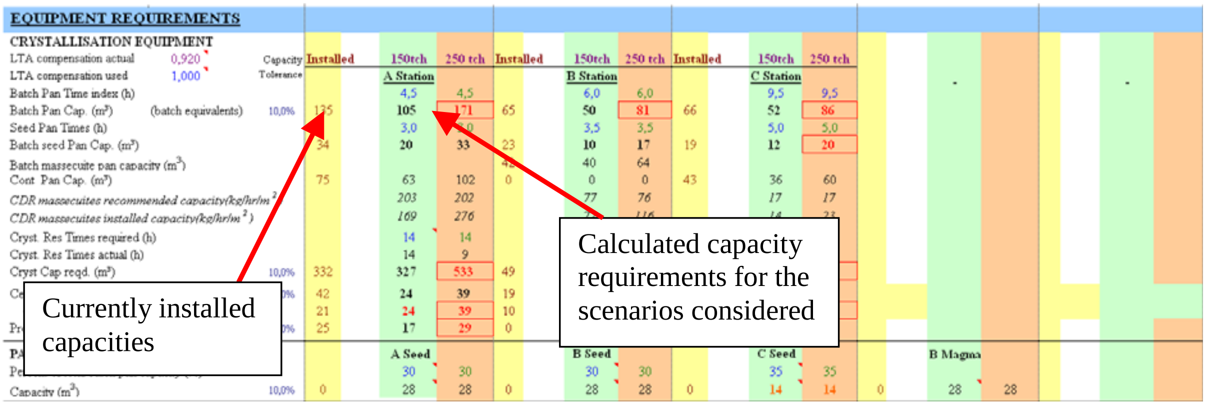

Figure 3. Dual Mass and Colour Balance – extract of equipment sizing section.

Overall Steam and Energy Balance

The design of a factory and the type of equipment installed will place a limit on the maximum level of energy efficiency that can be achieved. The purpose of carrying out an overall energy balance of the proposed mill is to optimise the process design of the factory so as to match its energy requirements, as far as possible, to the available fuel supply. At this stage, a number of available options may be evaluated for their effect on the energy balance of the factory, allowing the basic design to be refined and improved before carrying out more detailed modelling of the individual unit operations.

The model used for this application has been developed and improved over a number of years. It is currently in Excel spreadsheet format, although an earlier *PL/1* implementation of the model was described by Rein and Hoekstra (1994). The model itself is quite complex and therefore, for reporting or presentation purposes, a summary document is usually generated. An example of one page of the summary document, representing the clarification operation in a raw sugar mill, is presented as Figure 4.

Apart from carrying out a mass and energy balance over the entire sugar factory, some further features of the energy balance spreadsheet are as follows:

- The production of high pressure steam in the boilers is modelled.
- The generation of electrical power for in-house consumption or for cogeneration purposes may be optimised using the methods described by Hoekstra (2000).
- An estimate of the steam consumption of the mill during periods of unsteady-state operation can be made, based on the time account of the mill.
- The injection water circuit of the factory, including the cooling towers or cooling pond, is modelled. A full water balance of the factory is also available if required.
- An estimate of the operating costs of the factory, including labour, can be obtained and compared against the projected revenue earned by the sale of the various product streams produced.

In addition, various processing options may be readily incorporated in the balance, including mud recycle to the diffuser, double-curing of A- or C-sugars, alternative boiling house schemes, sulphitation in the rawhouse, and various refinery decolourisation methods and boiling schemes.

As there is a fair degree of overlap between the dual mass and colour balance of the mill and the overall energy balance. A number of iterations between these two models may be necessary to ensure that consistency is achieved in terms of the process configuration selected and the expected process performance.

Stream Number	1	2	3	4	5	6	7	8	9
Product	Mixed Juice	Lime Slurry	Flashed Juice	Clear Juice	Clarifier Mud	Bagacillo	Filter Wash Water	Filtrate Return	Filter Cake
Mass Flow Rate [t/h]	213.2	1.6	213.7	166.9	46.5	1.6	14.7	47.4	12.5
Volumetric Flow Rate [m^3/h]	211.7	1.5	211.8	164.7	45.7	-	14.9	46.5	-
Dissolved Solids Content [%]	13.0	-	13.0	13.5	13.5	-	-	9.5	1.3
Insoluble Solids Content [%]	0.9	8.0	1.0	-	2.3	50.0	-	0.7	25.4
Density [kg/m^3]	1 007	1 030	1 009	1 013	1 016	-	988	1 020	-
Moisture [%]	86.0	92.0	86.0	86.5	84.2	50.0	-	89.9	73.3
Pressure [kPa(a)]	-	-	-	-	-	-	-	-	-
Temperature [deg C]	103	50	100	96	92	53	50	60	60

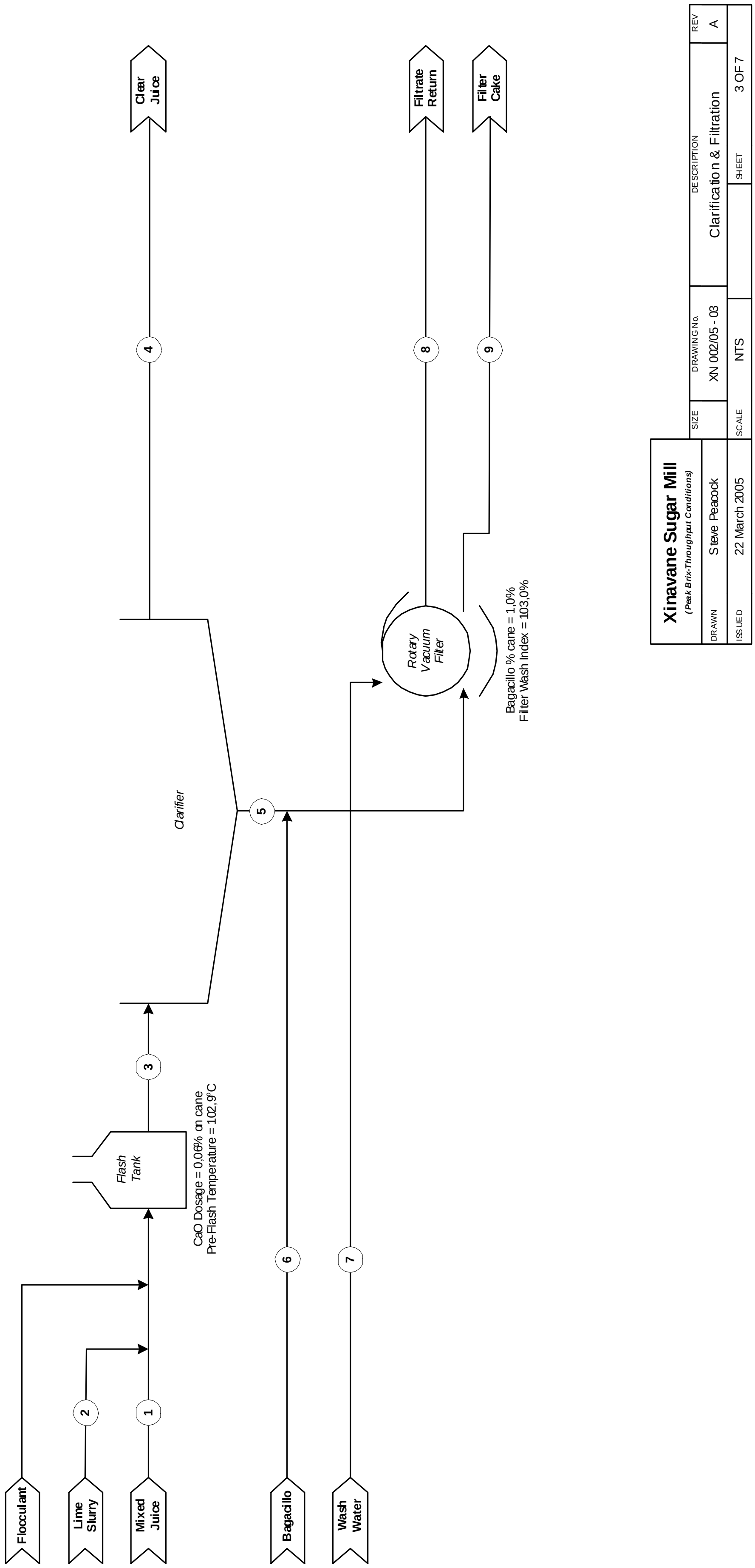

Figure 4. Example page from an overall energy balance summary document.

A primary area of focus in terms of energy efficiency is the evaporator station, with its arrangement of vapour bleeds (Love *et al.*, 1999). Of particular importance in this regard is the use of bleed vapour streams for juice heating purposes. A preliminary design for the evaporator station and the juice heaters should thus be developed in conjunction with the construction of the overall energy balance of the factory. This minimises the need for any later iterations of the energy balance model or the dual mass and colour balance model that would arise if the evaporator station and / or juice heaters were to be designed in isolation.

Detailed Evaporator Balance

The evaporator station, which is at the heart of the complex steam and vapour system of the factory, is designed using a detailed heat and mass balance model originally developed by Hoekstra (1981). The model itself is currently implemented as a *DOS*-based *Pascal* program and can be easily customised to describe a great variety of vessel configurations. The balance is comprehensive and even takes into account the recovery of flash vapour from condensates, which can become a significant factor in the latter vessels of the station. Some of the major inputs to this evaporator balance, such as the clear juice flow rate and the vapour bleed requirements of the factory, are supplied by the overall energy balance model of the factory.

Making use of the inherent flexibility of the computer model, the design of any proposed evaporator station can be optimised in terms of heating surface requirements, steam economy and operability. The sensitivity of the design to changes in its operating conditions may also be investigated (for example, see Love *et al.*, 1999). Any changes to the configuration of the evaporator station which result from the modelling investigations need to be incorporated in the overall energy balance of the factory, with iteration of the two models as necessary, so that the effect of such changes on the vapour bleed requirements from the station can be accounted for properly. An example print-out from the evaporator model is presented in Figure 5.

Heater calculations

Because the amount of sensible heat required for juice heating in a sugar factory is substantial, the vapour bleed requirements for mixed juice heating and for clear juice heating (if carried out) are of critical importance to the overall energy balance of the facility. Juice heating operations may be modelled using heat transfer and fluid flow fundamentals, as shown in the model output for a mixed juice heater simulation which is presented in Figure 6.

A cost-effective design for a juice heating station may be achieved by appropriate consideration of the approach temperatures achieved in each heater. Proper specification of the heater geometry (e.g. the number of tube passes) is also of critical importance in obtaining a design that is robust (for example, being resistant to fouling during routine operation), while still minimising the overall pressure drop across the units.

As with the detailed balance of the evaporator station, any change to the configuration of the juice heaters at this stage of the design process needs to be incorporated in the overall energy balance of the factory. Iteration of the models (the energy balance, evaporator balance and the heater balances) may be necessary in order to ensure that the effects of such changes on the steam balance of the mill are fully accounted for.

Finalisation of the design

As has been indicated in the text above, many runs of the heat and mass balances are usually necessary before the design can be finalised. Different runs are also necessary to simulate both the average conditions and the various peak conditions that occur at different times during the season (as defined by the throughput strategy). It may also be necessary to further iterate the design in order to meet the customer requirements for both the factory as a whole and for individual sections of the mill, considering standard process performance parameters (such as extraction and boiling house recovery), the steam economy of the facility and the cost of the installed plant.

EVAPORATOR UNITS (VESSELS):

Effect No. :		1	2	3	4	5
VAPOUR FLOWS:						
Flow rate (t/h)	: In :	63.0	46.5	4.0	5.7	9.1
	: Out :	63.2	47.1	5.7	6.4	9.9
Pressure(kPa abs)	:In :	200.0	150.9	68.9	56.4	36.5
	:Out :	150.9	124.7	56.4	36.5	16.0
Temperature(°C)	: In :	120.2	111.5	89.5	84.3	73.7
	: Out :	111.5	105.9	84.3	73.7	55.3
JUICE FLOWS:						
Flow rate (t/h)	: In :	165.0	101.8	54.8	49.0	42.6
	: Out :	101.8	54.8	49.0	42.6	32.8
Brix (°Bx)	: In :	13.5	21.9	40.7	45.4	52.2
	: Out :	21.9	40.7	45.4	52.2	68.0
Temperature(°C)	: In :	115.0	112.2	107.7	86.2	76.1
	: Out :	112.2	107.7	86.2	76.1	59.5

PERFORMANCE OF UNITS:

Hydrostatic Head (m) :	0.3	0.3	0.3	0.3	0.3
Hydrostatic Temp.Rise(K):	0.6	0.8	1.5	2.2	4.7
Boiling Point Elev. (K) :	0.7	1.8	1.9	2.3	4.1
Overall dT (K) :	7.4	3.1	1.8	6.1	9.6
Calandria Area (m^2) :	2000	4023	703	402	877
Calandria HTC (kW/m^2C) :	2.6	2.3	2.0	1.5	0.7
Calandria Heat Flux (MW):	38.5	28.7	2.5	3.7	5.9
$Area^2$/(Flux/HTC) (m^2/K):	270	1296	392	66	92

EXHAUST AND VAPOUR BLEED STREAMS:

Exh./Vapour Bleed No. :	Exh.	1	2
Flow rate (t/h) :	63.0	16.7	44.5
Pressure (kPa abs) :	200.0	150.9	124.7
Temperature (°C) :	120.2	111.5	105.9
Vapour Bleed/CJ (%) :	38.2	10.1	27.0

Throttling after Bleed? Yes dP = 55.8 kPa across Throttling Valve.

CONDENSATE VAPOUR FLASH:

Ref.No. of Cond. Flash	1	2
Flash into Effect No. :	3	5
Flash flow rate (t/h) :	1.4	2.7
Cascade after Flash(t/h)	45.1	62.1

Total Vapour Feed = 63.0 t/h Total Evaporation = 132.2 t/h
Total Vapour Bleed = 61.2 t/h Total C.J.Feed = 165.0 t/h
Total Syrup Make = 32.8 t/h Syrup Brix = 68.0 °Bx
Juice Purity = 86.4 %

Figure 5. Example print-out from the evaporator design model.

SHELL-AND-TUBE HEAT EXCHANGER 26-Jun-03
Assumes steam/vapour on shell-side and no phase change on tube-side

FACTORY	Xinavane	DUTY	MJ Heating - 150 TCH (2 Primary Heaters in Series)

"LOOK-UP" DATA			
Material			Mild Steel
Thermal conductivity	kW/m°C	kw	0.0429
Roughness	m	R	0.0000457
Fouling factor shell-side	kWm2°C	hs	11.63
Velocity heads/pass		Nh	2.5
Fouling factor tube-side	kWm2°C	ht	2.1

SHELL SIDE (Vapour)			V2	TUBE SIDE (Liquid)			Mixed Juice
Vapour pressure	kPa(a)		118.0	Mass flowrate	t/hr	Ml	215.0
Vapour satd. temp.	°C	Tv	104.0	Inlet temp.	°C	Ti	72.5
Vapour mass rate	t/hr	Mv	7.03	Brix	deg	Bx	14.0
OHTC	kW/m2°C	U	0.98	Outlet temp.	°C	To	91.2

CONSTRUCTION				TUBES			
Horizontal/Vertical (H/V)			V	Effective tube length	m	L	3.960
No.of tube passes		Np	14	Tube OD	m	D	0.04
Entry pipe ID	m	Di	0.200	Wall thickness	m	t	0.0015
Exit pipe ID	m	Do	0.200	Total no. of tubes		N	434
Heat transfer area	m2	A	216.0	Velocity	m/s	u	1.76
				Pressure drop	kPa	Pt	115.10

Figure 6. Example of the model output from the juice heater design model.

Sizing of major equipment

As mentioned previously, the Dual Mass and Colour Balance calculates preliminary sizes for major items of plant, based on their required capacities. These preliminary results can be supplied to equipment suppliers/manufacturers for the obtaining of budget quotes.

As the final design is honed, more accurate vessel designs are obtained from equipment specific models.

Incorporation of long term capacity plans

In some cases it is necessary to take account of future capacity plans when carrying out a process design, such as when carrying out an intermediate phase in a stage-wise factory expansion or when installing a back-end refinery at a mill where a future expansion of capacity is planned. Under these conditions, the design process may be constrained by the need for the installed equipment to be compatible with the future expansion plans. It may even be necessary to undertake a preliminary design for the expanded case in order to ensure that the right decisions are made at the intermediate stage of the project.

A brief example of a process design to take account of future expansion plans is the intermediate expansion of the evaporator station shown in Figure 7 to a factory throughput of 150 tons of cane per hour. Under normal circumstances, the existing second effect Kestner vessel would have been transferred to first effect duty and the first effect Robert vessel transferred to second effect duty. Additional surface area of around 2000 m^2 would have been necessary in the second effect of the station in order to provide the required capacity for the expanded case.

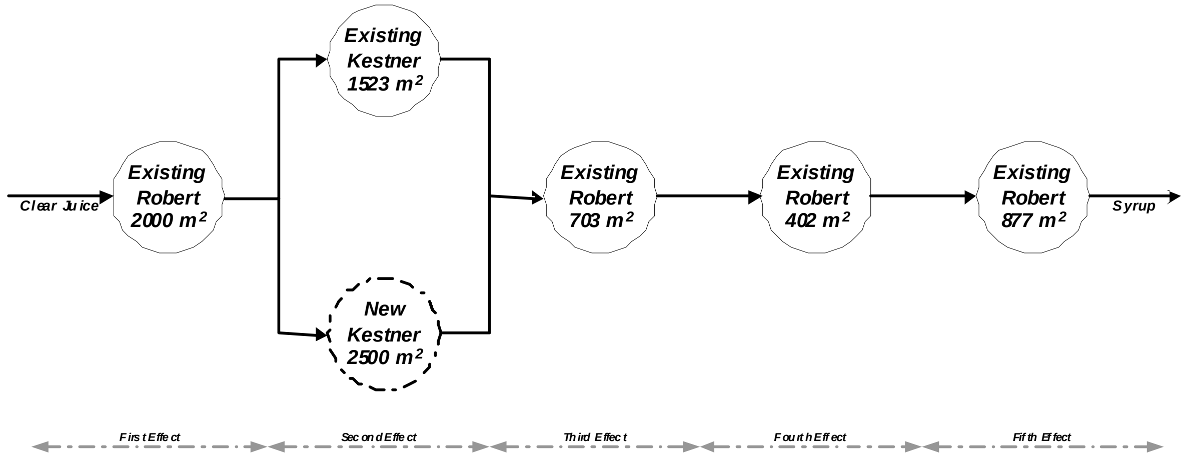

Figure 7. Evaporator configuration for the intermediate expansion stage (150 TCH).

However, taking into account the planned expansion of the mill to 250 tons of cane per hour in the longer term, this process design was modified as is shown in Figure 7. A new Kestner vessel of 2500 m^2 was installed in the second effect, with the existing vessels retaining their positions in the layout. This configuration is somewhat unconventional, in that it has a Robert vessel as a first effect and two Kestner vessels as second effects. However, this intermediate layout is more compatible with the long-term design of the station (which is more conventional in configuration), as illustrated in Figure 8.

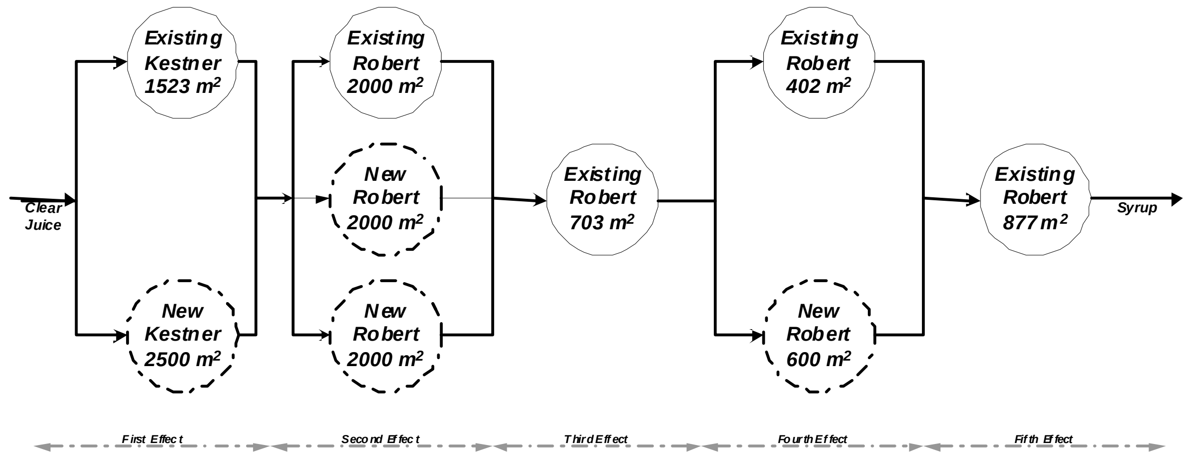

Figure 8. Evaporator configuration for the final expansion stage (250 TCH).

As of the writing of this paper, the intermediate expansion phase of 150 tons cane per hour has been in operation for one full season and the evaporator station has operated completely trouble-free at the required throughput.

Preliminary plant layouts

Using the preliminary plan sizes, a preliminary plant layout can be developed, taking into account both site and operability requirements. This plant layout will give preliminary requirements for factory buildings and steelwork. The plant layout will also define much of the requirements for piping and pumping. In carrying out expansion projects, a consideration of the existing factory layout can often have a substantial impact on the design which is finally implemented.

Preliminary costing

At this stage there is sufficient information to proceed with preliminary costing based on the factory layout and a list of all the major plant items required.

In some cases, the design process may result in two or more feasible options for the installation which need to be evaluated to determine which is the most cost effective. An example of such a scenario is the installation of clear juice heaters in a new or expanded sugar mill, which may ultimately be more expensive than installing a larger first effect evaporator vessel which accommodates the extra surface area required for juice heating within the evaporator itself (Peacock and Love, 2003). At the preliminary costing stage, the various available options may be evaluated based on their total installed cost and a decision made as to which of the alternatives to install.

Conclusions

Because of the complex interactions between the various operations in a sugar factory, the procedure used for the design of a greenfields factory or for an expansion project cannot be a simple ‘once-through’ process. In order to accommodate the interactions, the procedure used by the design team needs to be iterative in nature and requires the use of engineering judgement to make appropriate assumptions, thereby avoiding excessive iterations of the final design.

A properly structured and comprehensive design procedure has been presented which allows a well thought-out design to be finalised early in the project, thus meeting the standard requirements of cost, performance and timing. The design is based on an agreed-upon set of basic design data, a cost-effective cane supply and throughput strategy, the appropriate selection of process technology to be used and the detailed modelling of the various unit operations in the sugar production process, as well as of the overall factory design which is proposed.

REFERENCES

Archibald RD and Smith IA (1975). The effect of low juice purities at Darnall on boiling house capacity. *Proc S Afr Sug Technol Ass* 49: 63-73.

Hoekstra RG (1981). A computer program for simulating and evaluating multiple effect evaporators in the sugar industry. *Proc S Afr Sug Technol Ass* 55: 43-50.

Hoekstra RG (2000). The use of linear programming in power generation decisions. *Proc S Afr Sug Technol Ass* 74: 267-273.

Love DJ, Meadows DM and Hoekstra RG (1999). Robust design of an evaporator station as applied to the Xinavane rehabilitation project. *Proc S Afr Sug Technol Ass* 73: 211-218.

Peacock SD and Love DJ (2003). Clear juice heaters - Do we need them? *Proc S Afr Sug Technol Ass* 77: 452-462.

Rein PW and Hoekstra RG (1994). Implications of cane diffusion for energy economy in a sugar mill. Proceedings of an ISSCT Combined Workshop on Efficient Production and Utilisation of Steam in Sugar Factories, Pune, India, pp 347-348.

APPENDIX 1
Summary of the process of factory design