Development and Verification of an Aspen Plus® Model of a Sugarcane Biorefinery

by

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Prof. A. Stark

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DECLARATION

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The research reported in this dissertation/thesis, except where otherwise indicated, is my original work.

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ABSTRACT

This dissertation describes a steady-state mass and energy balance model of a raw sugar mill which has been developed using the Aspen Plus® software.

The Aspen Plus® model was designed to replicate an existing MATLAB™ model of a 'generic' South African sugar mill with the aim of facilitating future expansion to different products which cannot be easily handled in MATLAB™.

The first step was to create the entire sugar mill flowsheet in Aspen Plus®. This involved deciding how best to model the complex processes such as multiple-effect evaporators and pans. Key factors in simulating sugar mills are modelling the boiling point elevation and crystallisation. The UNIQUAC thermodynamic model with coefficients regressed by Starzak (2015) was used in Aspen Plus® to predict the vapour-liquid equilibria in sugarcane juice solutions. Also, a solid-liquid equilibria model was developed in order to accurately handle the crystallisation and dissolution of sucrose.

A dynamic tool (proportional integral controller) was used to solve the material balance of the evaporator station. Microsoft Excel® has been incorporated into the Aspen Plus® model to iteratively solve this dynamic tool.

Initially, the Aspen Plus® model was verified for a cane throughput of 244 t/h against the results of the existing MATLAB™ model. The stream results showed a good comparison between Aspen Plus®, MATLAB™ and real sugar mill data.

Different scenarios have been tested in Aspen Plus® and compared to the MATLAB™ model. Cane throughputs of 230 and 270 t/h were simulated and the results compared favourably between the two models. The energy requirement of the sugar mill was calculated for different flow rates of imbibition (water used for juice extraction). A portion of an intermediate stream (a potential biorefinery feedstock) was diverted after the clarifiers and the effects on the rest of the sugar mill were quantified. Finally, different cane purities were simulated in order to assess the effects on syrup purity, molasses purity and boiling house recovery.

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NOMENCLATURE

Symbol	Description	Units
A	heat transfer area	m^2
$Brix_{ref}$	refractometer brix	%
DS	dry substance	%
F^i	total mass flow rate of stream i	kg/h
F_j^i	mass flow rate of component j in stream i	kg/h
NSW	(non-sucrose)-to-water ratio	
Pol	a measure of sucrose by polarimetry	%
PU	stream true purity	%
PU_{app}	stream apparent purity	%
Q	heat duty	MJ/h
SC	solubility coefficient	
Suc	sucrose concentration	%
SW	sucrose-to-water ratio	
T^i	temperature of stream i	°C
U^i	overall heat transfer coefficient, $i =$ evaporation effect number 1-5	$W/(m^2K)$

Abbreviations	Description
wat	Water
suc	sucrose (dissolved)
nsuc	non-sucrose
fib	Fibre
lim	Lime
cry	crystal (sucrose)

CHAPTER 1: INTRODUCTION

1.1 Rationale for the research

Sugar has been produced in South Africa since the mid-19th century. It is an essential ingredient in cakes, desserts, sweets and cooldrinks. According to the Sugar Milling Research Institute NPC (SMRI), 2.36 million tons of sugar were produced by 14 sugar mills during the 2013/2014 milling season (SMRI, 2015). Most of the sugar mills in South Africa are in the Kwa-Zulu Natal region. The South African sugar industry helps to grow the economy by generating an estimated R8 billion annually. According to the South African Sugar Association (SASA, 2017) this industry provides 79 000 direct jobs, as well as an estimated 350 000 indirect jobs, which consist of growing and harvesting sugar cane. Assuming that each job-holder has a number of dependents, close to one million people are supported by the sugar industry in South Africa.

Unfortunately, the South African sugar industry is battling economically. This is due to increasing costs of production and a low international market price for sugar. The South African sugar industry needs to find a competitive advantage in order to increase its profits. There are many products which could be made from the sucrose or hemi-/cellulose found in sugarcane. Currently in South Africa, sugar is the main source of income generated from sugarcane. Countries such as Brazil and India have other sources of income from sugarcane. Brazil produces large quantities of bio-ethanol (Amorim and Lopes, 2005), whereas sugar mills in India provide large amounts of electricity to the national grid (Natu, 2012).

Sugar mills produce a large quantity of sugarcane biomass (bagasse) which could be turned into high value products. Also, the sucrose in sugarcane could be turned into higher value products than sugar. Examples of potential products are chemicals such as levulinic acid and energy products such as bio-butanol. One of the goals of the Sugarcane Technology Enabling Programme for Bioenergy (STEP-Bio) is to identify which new products would be profitable in the South African context. In order to make these products in existing sugar mills, an accurate process model is required of the various unit operations of a sugar mill. Mass and energy balances across all unit operations are needed to determine how changes in operation would affect stream flow rates and compositions. Once this is known, economic studies may be undertaken in order to assess the changes in profitability.

1.2 Existing model

Recently, the MATLAB™ software was successfully used to model a sugarcane biorefinery (Starzak, 2015). The definition of a biorefinery is a facility which sustainably processes biomass into a wide variety of marketable products and energy (de Jong, 2015). MATLAB™ is a programming language which was created in order to solve mathematical equations easily. The MATLAB™ model was developed at the SMRI by Prof. M. Starzak (2015). This model has been verified by the Sugars® software and validated by actual sugar mill data. However, MATLAB™ has no built-in sugar knowledge and cannot easily be applied to chemical processes.

1.3 Aspen Plus® software

Aspen Plus® is a market-leading chemical process optimization software (www.aspentech.com). It is generally described as the most powerful process modelling tool available. Aspen Plus® has a very large chemical database. The newest Aspen Properties® includes over 9,000 sets of binary parameters from the NIST physical property databanks. This provides unrivalled thermophysical properties modelling which is ideal for the products (n-butanol, lysine, levulinic acid etc.) and processes (distillation, extraction etc.) of a biorefinery. However, sugar streams are a complicated mixture of chemicals and thus require customised thermodynamic models for use in Aspen Plus®.

1.4 Purpose of the study

The aim of this project was to model a 'typical' South African sugarcane biorefinery with the requirement that the Aspen Plus® software needed to be used. The Aspen Plus® model was to be based on the existing MATLAB™ model. This new model would be valuable to the Biorefinery Techno-Economic Modelling (BRTEM) sub-section of the STEP-Bio programme. One of the desired outcomes of the BRTEM group was to develop a verified model of a South African sugar mill using the Aspen Plus® platform. This model would be used in future for research to assess new product viability. The Aspen Plus® model will allow expansion to additional processes and products in a sugarcane biorefinery. These processes cannot easily be programmed in MATLAB™ and chemical databanks would need to be programmed in for MATLAB™ simulations.

The Aspen Plus® model needed to have the same level of detail and flexibility as the existing MATLAB™ model, showing all streams and their properties. The main requirement was to show a sugarcane processing factory with raw sugar as a product. After this, a model of cogeneration and ethanol production could be attempted. The goal was to simulate various operating conditions of sugar mills and quantify the results.

1.5 MATLAB™ model development

A 'generic' South African sugar mill has been modelled using the MATLAB™ software (Starzak, 2015 & 2016a). The model includes mud filtration and a 3-boiling partial 'remelt' boiling house configuration. Mass and energy balances were undertaken for the unit operations of the sugar mills. These unit operations were placed in six modular blocks, namely: extraction, clarification, evaporation, crystallisation, sugar drying and utilities. Degree of Freedom (DOF) analyses were performed on the modular blocks. Technological and design choices were made in order to reduce the DOF. For example, input streams were specified and requirements were imposed on various intermediate and product streams. Separation coefficients were specified for the diffuser, clarifier, vacuum filter and syrup clarifier. This ensured that the DOF around each module was zero and thus a unique solution could be calculated.

In the next chapter the MATLAB™ model, as well as other computer programs and software which have been used to model sugar mills, are reviewed.

CHAPTER 2: LITERATURE REVIEW

2.1 Sugar mill layout

Sugarcane arrives at sugar mills and is offloaded onto large feeder tables. The flow diagram of a typical sugar mill is shown in Figure 2.1. The cane goes through a preparation stage in which the cane is chopped and shredded in order to rupture sucrose-containing juice cells. The prepared cane is sent to an extraction process in which the sucrose is washed out from the cane biomass. The biomass is then pressed by mills before being sent to the boilers for use as fuel.

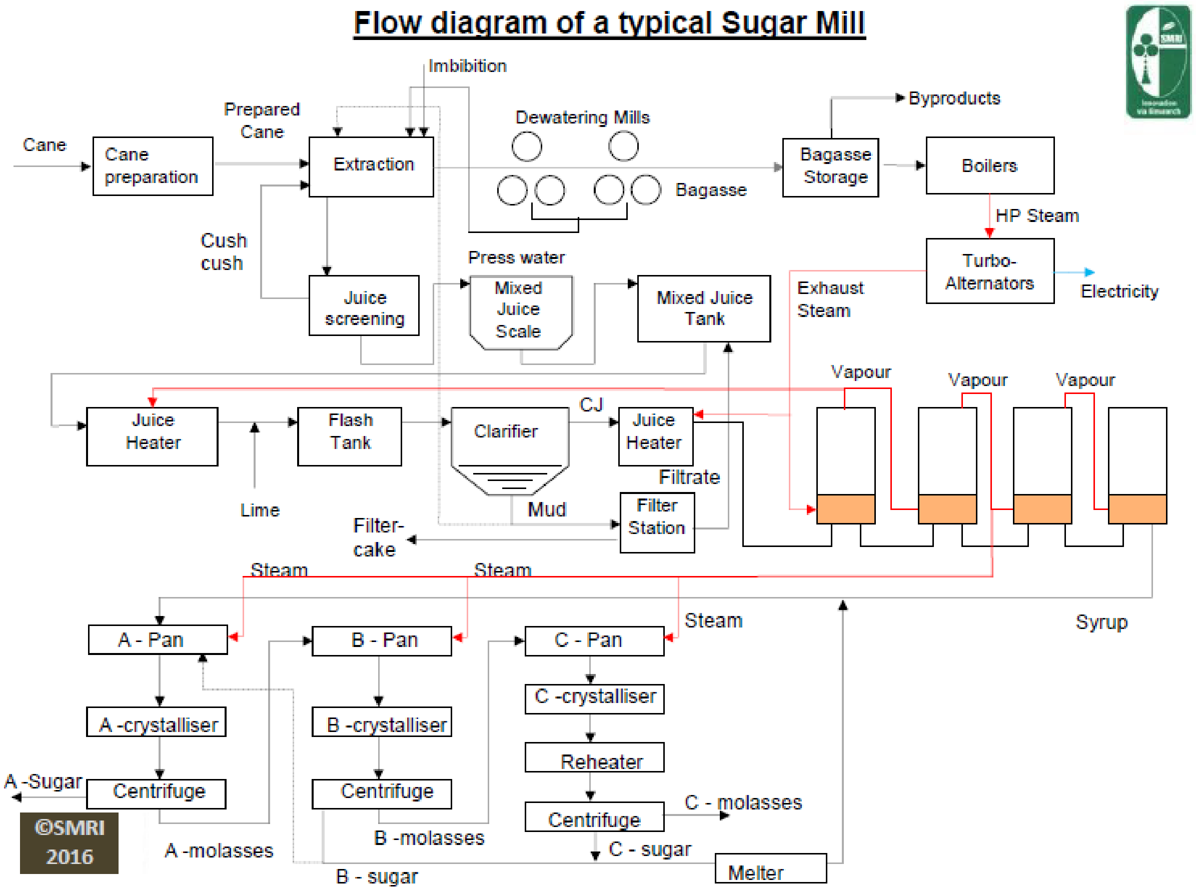

Figure 2.1 Sugar mill layout (SMRI, 2016)

The sucrose-rich juice is screened, weighed and then sent to the mixed juice tank. The juice from this tank is sent to a series of heaters and thereafter flashed to remove trapped air particles. The next step is to clarify the juice in order to remove suspended solids impurities and adjust the pH to neutral. Mud from the clarifiers is sent to a vacuum filtration unit in order to recover juice. This juice is then recycled to the mixed juice tank.

Clear juice from the clarifiers is sent to a series of evaporators in order to concentrate the juice. After the evaporation process the sucrose solution is called syrup. The syrup is sent to large vessels called pans where sucrose is crystallised under vacuum. Cooling crystallisers are then used to crystallise more sucrose at ambient pressure. The sugar crystals are separated from the molasses in centrifuges. The raw sugar is dried in order to prevent caking.

The utilities section of a sugar mill consists of boilers and cooling towers. The boiler produces high pressure (HP) steam for use in the mill. Turbo-alternators generate electricity from HP steam. Exhaust steam from the turbo-alternators is used to provide the heating duty in the evaporation station. Cooling water is used to maintain a vacuum in some of the evaporators and all the vacuum pans. Cooling towers are used to lower the water temperature, allowing the cooled water to be returned back to the process.

2.2 Process modelling of sugar mills

There are a number of models which have been designed to simulate sugar mills. Process modelling has been undertaken by experts using a variety of different techniques. Specialised computer programs and spreadsheets have been developed using most high level programming languages (such as FORTRAN, PASCAL, BASIC, C, etc.) (Peacock, 2002).

2.3 Computer programs

Guthrie (1972) created a computer program to solve the material balance of a raw sugar mill. However, the model did not have the capability to solve the energy balance. Due to the presence of recycle streams the model was solved in an iterative manner. The outputs of the model showed the various streams and their compositions. The absence of an energy balance makes it inadequate to handle the current challenges in the South African sugar industry since energy usage is a crucial factor.

Hoekstra (1981) focussed on the simulation of evaporator stations in sugar mills and developed the Program for Evaporation Simulation and Testing (PEST). The computer program had the capability to simulate condensate flash vapour return, vapour bleeding and throttling between effects. Heat losses were taken into account and heat transfer coefficients could also be predicted. He designed the computer program using FORTRAN. This means that users of the model need to have an understanding of computer programming. One of the aims for future development of the PEST model was for it to be incorporated into an overall material and heat balance model of a sugar mill. However, to be able to use this program effectively, a thorough understanding of the principles of multiple effect evaporators and the relevant calculations would be required (Love, 1999).

Mass and energy balances of a new sugar mill were undertaken by Reid and Rein (1983). It was a purpose-built model which focussed on steam economy in order to supply bagasse to a downstream paper plant. The model was based on a crushing rate of 600 tonnes cane per hour. The standard three-boiling system was solved using a computer program. Hoekstra (1983) described the boiling house model, which was coded in the PL/1 programming language. At a later stage the model was translated into spreadsheet format. In subsequent developments, Hoekstra (1985, 1986) modelled a continuous vacuum pan. The pan was assumed to be made up of stirred tank reactors. However, the 'generic' South African layout consists of batch and continuous vacuum pans.

2.4 Spreadsheet models

Spreadsheet packages are the most common means of simulation in the sugar industry. These packages are easy to use and are readily available. Complete sugar mill mass and energy balances have been developed using spreadsheets.

Radford (1996) designed a spreadsheet for the mass, energy and colour balances of a sugar mill, which included a model of crystallisation. However, no details of the mass and energy balances were given, with only two equations listed which relate colour transfer to crystallisation rates.

Using spreadsheet software, Hubbard and Love (1998) were able to fit a mass balance to an over-specified system. The numerical analyses of a continuous centrifuge were performed by the optimisation utility of the spreadsheet. Love (2017) was able to explain how an over-specified mass balance could be solved in Aspen Plus® when linked to Microsoft Excel®.

2.5 Commercial flowsheeting packages

ChemCad and PRO/II are flowsheeting packages which have been designed for the chemical and petrochemical industries. However, they do not have models for the equipment commonly found in sugar mills. Since sucrose is a non-volatile soluble component these flowsheeting packages struggle with the prediction of vapour-liquid equilibria for sugar streams (Peacock, 2002).

2.6 SUGARS™ software

SUGARS™ is a commercial flowsheeting package which was specifically developed in order to simulate the behaviour of sugar mills and ethanol refineries (www.sugarsonline.com). The software has been used in many parts of the world to assist with process modifications. It was used successfully to model two South African sugar mills, Malelane and Komati (Stolz and Weiss, 1997). The results from the mass and energy balances calculated by the SUGARS™ program were similar to the performance data from the mills. It is relatively easy to use, and models may be transferred between developers. However, SUGARS™ cannot be used to model any other biorefinery products besides ethanol.

2.7 Existing Aspen Plus® models

Various Brazilian authors have used Aspen Plus® to model sugarcane biorefineries. However, the focus of these models has been on ethanol production.

Bonomi et. al. (2016) developed a Virtual Sugarcane Biorefinery (VSB) in order to simulate the entire sugarcane production chain. The aim was to evaluate new alternatives from sugarcane and different technologies for biofuel and biochemical production. Results from the VSB were validated against existing mills. Aspen Plus® was used to model the biorefinery. Sugar mills producing sugar, ethanol and electricity have been classified as a phase II biorefinery. This means that they have the flexibility to produce multiple products from one feedstock, depending on economic factors like product demand and market prices. Sugarcane juice can either be sent to produce sugar or bioethanol in the same facility.

The most common configuration considered in the Brazilian report was using half the sugarcane juice to produce sugar while the other half, along with all the molasses, is used to produce bioethanol. However, no flowsheet was given for the sugar mill process. The physical properties for the feedstock were shown, but no details were provided on how Aspen Plus® was used to model the sugar mill. Presentations given by the same author describe the various model building blocks used in Aspen Plus® but offer only basic flowsheets.

Palacios-Bereche et. al. (2013) used Aspen Plus® to develop process simulations of a sugar mill with the aim of investigating a new technology for ethanol production. The UNIQUAC model was selected because of its accuracy in predicting boiling point elevation. A FORTRAN subroutine was developed in order to handle the enthalpy calculations of sugarcane juice. Although a detailed flowsheet was shown, no crystallisation processes were modelled. In this model, syrup from the evaporators was sent to fermentation in order to produce ethanol. This is different from the South African sugar mill configuration where syrup is sent to crystallisation.

2.8 The MATLAB™ sugar mill model

Starzak and Zizhou (2015) describe the mathematical framework of a 'generic' South African sugar mill modelled using the MATLAB™ software. Their report deals with five modular blocks of a sugar mill model, namely: extraction, clarification and filtration, evaporation, crystallisation and utilities. The steady-state mass and energy balances were listed for each modular block. The evaporation module was solved for two cases: either assuming equal heat transfer areas or assuming a pressure distribution in the different evaporation effects. At this stage of development of the MATLAB™ model, only a four-effect evaporator station was modelled. Sucrose inversion and entrainment were not taken into account.

Starzak (2016b) continued the development of the MATLAB™ sugar mill model with the aim of validation. The evaporation module was modified to include five effects. Sucrose inversion and entrainment were modelled for evaporation and pan boiling processes. A new solubility coefficient equation was proposed in order to adequately model the sugar crystallisation process. The model was also extended to include a sugar drying module.

Comprehensive validation of the model was achieved using 51 sugar mill performance indices taken from the 90th Annual Review of the 2014/2015 Milling Season in Southern Africa (Smith et. al., 2015). Some examples of the performance indices are sucrose extraction, bagasse pol, imbibition usage, limestone usage, filter cake pol, syrup refractometer brix, boiling house recovery, cane-to-sugar ratio and steam-to-cane ratio. Data was selected from seven South African sugar mills which all have mud filtration and a three-boiling partial 'remelt' configuration. Operating parameters were optimized in the MATLAB™ model in order to produce results which matched the average values of the 51 performance indices of the sugar mills.

2.9 Aspentech® software

The modelling of sugar streams in Aspen Plus® is challenging due to the complex physical property database required. In order to use Aspen Plus® for modelling a sugarcane biorefinery, online courses by Aspentech® training (https://esupport.aspentech.com/t_homepage) were attended to gain a better understanding of the Aspen Plus® software. Topics which were covered include: optimization, sensitivity analyses, advanced flowsheeting and troubleshooting.

Aspen Custom Modeller® was considered for the creation of custom sugar mill unit operations. The AspenTech Jumpstart guide for Aspen Custom Modeller® (www.aspentech.com) was studied in order to create a custom model of the diffuser and dewatering mills. Equations may be entered implicitly in Aspen Custom Modeller®. The software has built-in algorithms for solving the equations.

2.10 Conclusions

A variety of methods for simulating sugar mills have been reviewed in this chapter. To summarise:

- Computer programs require understanding of programming languages (e.g. FORTRAN and PL/1).
- Spreadsheet models are the easiest to use, however, they lack flowsheeting capabilities.
- Commercial flowsheeting packages struggle with modelling sugarcane juice streams and the unique processes which occur in sugar mills.
- The SUGARS™ software has been designed to model sugar mill operations. However, it is not flexible enough to allow detailed in-house process knowledge to be implemented. It also does not support expansion to other biorefinery products besides ethanol.
- Aspen Plus® has been used extensively in Brazil for the purpose of modelling sugar mills with annexed distilleries.
- The MATLAB™ model has been meticulously developed in order to model most aspects of South African sugar mills. However, it has been programmed using low-level language and extensive programming would need to be done for each additional biorefinery product.
- The Aspentech® software has all the capabilities available in order to successfully model sugarcane biorefineries.

2.11 Aims, hypothesis and objectives of this project

The aim of this project was to provide a modelling tool for future research to assess new product viability of sugarcane biorefineries in South Africa. From the literature review, it can be concluded that none of the existing packages can be readily used for the aim of this project. It was proposed that the Aspen Plus® software, with its large chemical database, and comprehensive unit operation building blocks, would be ideally suited to modelling future biorefinery operations (including fermentation and distillation). Hence, the hypothesis of this project was that Aspen Plus® may be used to model a sugar mill with the same level of accuracy as the MATLAB™ model and with more flexibility (Graphical User Interface and easier to adjust). Therefore, a raw sugar mill model was built with the following objectives:

- To develop a model of a 'generic' South African sugar mill using the Aspen Plus® software.
- To verify the Aspen Plus® model against the results of the existing MATLAB™ model.
- To test different sugar mill operating scenarios which may be useful for biorefinery applications.

CHAPTER 3: ASPEN PLUS® MODEL DEVELOPMENT

3.1 Conversion of the MATLAB™ model into Aspen Plus®

Firstly, a detailed study of a sugarcane mill process flowsheet was undertaken. Thereafter, the MATLAB™ model was studied to understand the particular unit operations that would require unit operation development in Aspen Plus®. The MATLAB™ model uses 194 process parameters and low-level code in order to specify all the operations of the sugar mill. The next step was programming the unit operations and the overall flowsheet in Aspen Plus®.

The Aspen Custom Modeller® software was considered for programming the complex unit operations. Initially this approach was discarded due to a lack of software (Microsoft Visual C+) which was required to export the custom models into Aspen Plus®. Later, the idea of making custom models was rejected due to the lack of flexibility: operating parameters and governing equations would not be easily accessible.

Aspen Plus® is a high-level programmable software. Customised thermodynamic models were needed in order to model sugar juice streams. Experimental data for boiling point elevation in impure sucrose solutions was regressed to the UNIQUAC thermodynamic model by Starzak (2015) for use in the Aspen Plus® model. An exponential equation for the solubility coefficient was developed by Starzak (2016b) and correlates the effect of the non-sucrose to water ratio on the solubility coefficient. This equation was used with the Vavrinecz equation (solubility of pure sucrose in water) in order to predict the solid-liquid equilibria in the Aspen Plus® model.

3.2 Setup

Aspen Plus® V8.8 was used to develop the model. The components which were used to construct the model are water, sucrose, non-sucrose, fibre, lime and air. Non-sucrose refers to any soluble components which are not sucrose while fibre specifies any insoluble components. Properties of D-fructose with a modified molecular weight were chosen as a suitable representation of the non-sucrose component. Similarly, cellulose was used to describe fibre properties. Table 3.1 shows the physical properties of the components in the Aspen Plus® model.

Table 3.1 Molecular weights of components used in the Aspen Plus® model

Component	Molecular weight (g/mol)
Sucrose	342.30
Water	18.02
Non-sucrose (D-Fructose)	204.00
Fibre (cellulose)	162.14
Lime	56.08

3.3 Thermodynamic model

In order to model the physical properties (e.g. boiling point elevation) of sugar streams in Aspen Plus®, regression on experimental data was required. The UNIQUAC model was chosen and regressed as it gave a better prediction of the boiling point elevation than the NRTL model. Table 3.2 shows the results of the regression on experimental data to the UNIQUAC thermodynamic model (Starzak, 2015), using several literature sources (Anon., 1955; Batterham and Norgate, 1975; Thieme, 1927; Saska, 2002).

Table 3.2 UNIQUAC parameters for the ternary system of water-sucrose-(non-sucrose)

Component I	WATER	WATER	SUCROSE
Component J	SUCROSE	NON-SUC	NON-SUC
AIJ	0.20278	4.7097	4.4965
AJI	-0.37225	-4.0332	-1.1452
BIJ	32.6532	-44.4349	218.7819
BJI	23.7752	-173.741	173.9059

3.4 Overall flowsheet

The overall flowsheet (Figure 3.1) shows the seven major processes in a sugar mill (extraction, clarification, evaporation, crystallisation, boiler, drying and cooling tower). Each process (module) was put in a separate flowsheet (called a hierarchy in Aspen Plus®). The lines connecting the hierarchies show the stream connections between the different modules. A key to the stream styles and colours is given in Table 3.3. The process of developing each module is explained in Chapter 4: ASPEN PLUS® MODEL OVERVIEW. A list of stream names may be found in Appendix F.

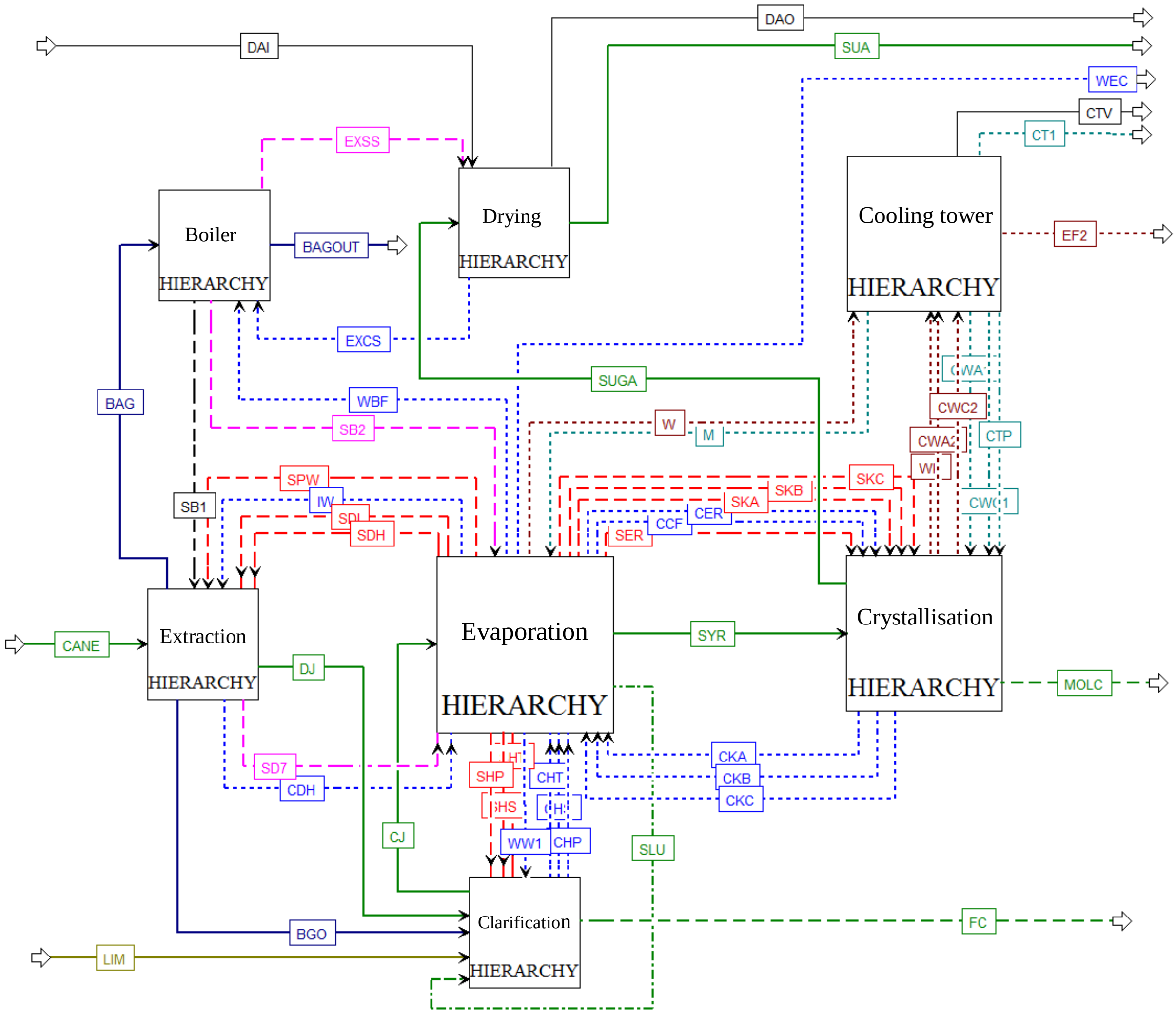

Figure 3.1 Overall flowsheet of the raw sugar mill model in Aspen Plus®

Table 3.3 Stream style key

Type of line	Description
	High pressure (live) steam – 31 bar absolute (bara) and 390 °C
	Exhaust steam – 2 bara and 121 °C
	Vapour bleeds from evaporators (effects 1-3)– V1 at 1.6 bara and 113.8 °C; V2 at 1.25 bara and 106.6 °C; V3 at 0.6 bara and 86.7 °C
	Process streams (containing sucrose)
	By-product process streams (final molasses and filter cake)
	Recycle process streams (e.g. filtrate juice and 'remelt')
	Bagasse and bagacillo
	Condensate
	Cold cooling water
	Warm cooling water
	Other streams (flash vents, air etc.)

3.5 Building blocks

The unit operations of a raw sugar mill can be modelled in Aspen Plus® using the following building blocks:

3.5.1 Mixers

Mixers take multiple streams and join them into one as shown in Figure 3.2. An example of this is the mixed juice tank.

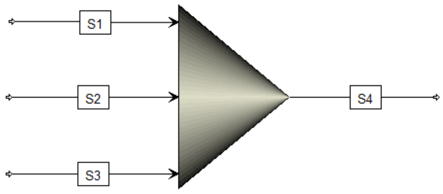

Figure 3.2 Aspen Plus® representation of a mixer.

3.5.2 Distributors

Distributors take one stream and split it into many streams as shown in Figure 3.3. e.g. Vapour bleed splitters.

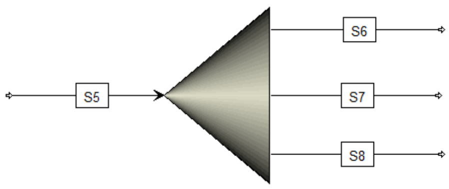

Figure 3.3 Depiction of a distributor in Aspen Plus®.

3.5.3 Separators

Separators are a kind of distributor in that they also take one feed and split it into many streams as shown in Figure 3.4. However, the outlet streams differ in composition. Separation coefficients govern the splits of the components to the different outlets. e.g. Diffuser.

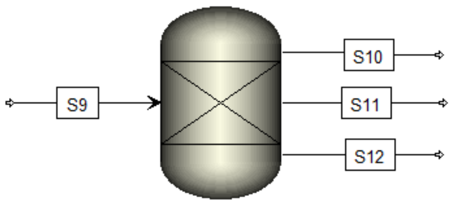

Figure 3.4 Representation of a separator in Aspen Plus®.

3.5.4 Flash vessels

Flash vessels in sugar mills involve a pressure change. When a mixture undergoes a sudden drop in pressure the associated boiling temperature drops too. Thus the mixture experiences a rapid boiling and a vapour-liquid equilibrium is reached. Separation into vapour and liquid streams is governed by the heat balance and vapour-liquid equilibria which depend on the thermodynamic model. Thus, a flash vessel has one inlet and a vapour and liquid outlet as shown in Figure 3.5. e.g. Mixed juice flash.

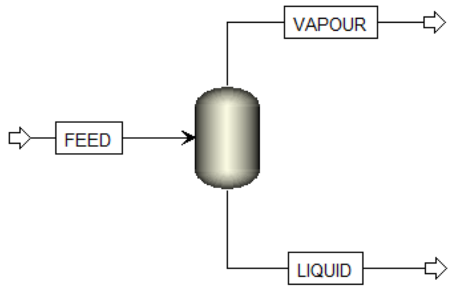

Figure 3.5 Aspen Plus® illustration of a flash vessel.

3.5.5 Pumps

Pumps are responsible for moving liquid mixtures along pipes by using impeller blades to propel the mixture forward with a resultant increase in pressure as shown in Figure 3.6. e.g. Clear juice pump. In the Aspen Plus® model the discharge (outlet stream) pressures were specified.

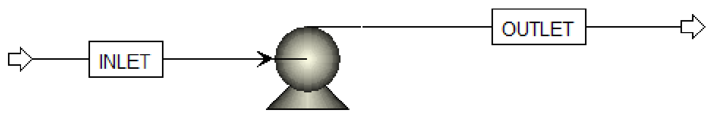

Figure 3.6 Illustration of a pump in Aspen Plus®.

3.5.6 Heat exchangers

Heat exchangers facilitate heat transfer between hot and cold streams as shown in Figure 3.7. The hot stream loses heat and the cold stream gains heat. The heat transferred is proportional to heat transfer coefficients, heat exchange area and temperature differences between the hot and cold streams. e.g. Mixed juice heater.

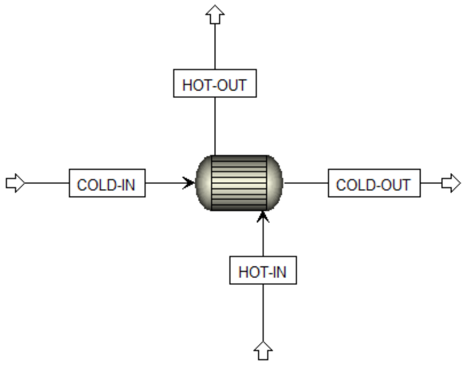

Figure 3.7 Aspen Plus® representation of a heat exchanger.

3.5.7 Heaters (including coolers and condensers)

In heaters, coolers and condensers, an external factor (e.g. heat losses to the environment) alters the temperature of the process streams as shown in Figure 3.8. Heat transfer duties or temperature changes may be specified.

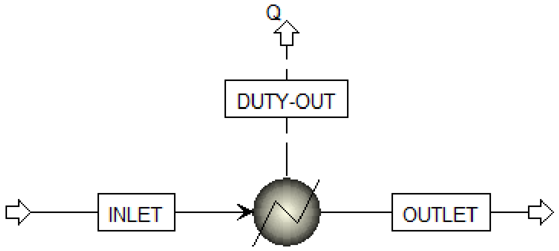

Figure 3.8 Aspen Plus® representation of a heater.

3.5.8 Reactors

Reactors are vessels in which chemical reactions take place as shown in Figure 3.9. Sucrose crystallisation or inversion may be modelled in a reactor (even though crystallisation is not a reaction). The extent of the reaction is governed by reaction kinetics and subsidiary relationships. e.g. solid-liquid equilibria.

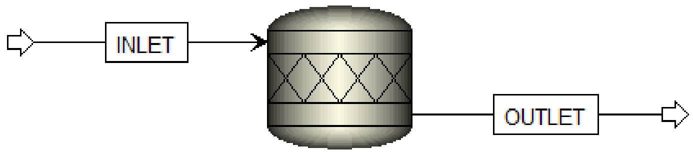

Figure 3.9 Representation of a reactor in Aspen Plus®.

3.6 Cane feed composition

When sugarcane enters the sugar mill, it generally consists of (Rein, 2007, page 37):

- 70 % water
- 15 % dissolved matter
- 15 % fibre (insoluble matter)

The valuable portion of the dissolved matter is sucrose which makes up 13 of the 15 %. The sucrose is crystallised in the final stages of a raw sugar mill. Sucrose is the chemical name for pure sugar and has the formula $C_{12}H_{22}O_{11}$.

The other 2 % of dissolved matter is mainly glucose and fructose but also small amounts of acids, salts and starch (Rein, 2007, page 38). Most of this portion, which is called non-sucrose, leaves the mill in the molasses (the liquid portion remaining after crystal sucrose has been separated), and is a by-product of sugar mills.

The 15 % fibre consists mostly of biomass (bagasse) which is burnt in boilers to provide the sugar mill with electricity and process steam (Rein, 2007, page 603). A small amount of the fibre leaves the mill as filter cake which is another by-product (Jenkins, 1966, pages 210-218).

For every 100 tonnes of cane processed, around 12 tonnes of sugar are produced. Typical quantities of the by-products formed are 30 tonnes of fibrous residue (bagasse), 4 tonnes of molasses and 1 tonne of filter cake (Anon., 2012, page 2).

3.7 Terminology

3.7.1 Refractometer brix

The term brix applies to all dissolved matter and is a percentage of mass or an actual mass. If 100 grams (g) of sucrose solution has 30 g of dissolved matter, then the solution has 30 % brix or 30 g of brix.

Brix is measured with a refractometer. This is an optical instrument which measures how much light is refracted when it enters a solution. The amount by which the light is refracted at the surface of the solution is related to the quantity of dissolved matter (Rein, 2007, page 578).

3.7.2 Pol

Pol is a measure of the amount of sucrose in a substance.

Sucrose is an optically active chemical. This means that it has the ability to rotate the plane of polarisation when polarised light is passed through a solution containing sucrose (Rein, 2007, page 577).

The amount which the plane of polarisation rotates is related to the concentration of sucrose. However, glucose and fructose are also optically active. So pol only gives the apparent sucrose concentration in a solution (Rein, 2007, page 577).

3.7.3 Purity

True purity is a ratio of the sucrose content to the dry solids. Apparent purity is the ratio of pol to refractometer brix (Rein, 2007, page 29).

An apparent purity of 60 % means that there are 60 g of pol (apparent sucrose) per 100 g of refractometer brix (all dissolved solids). The following formula is used to calculate apparent purity:

$$PU_{app} = \frac{Pol}{Brix_{ref}} \times 100$$

3.7.4 Non-pol

The portion of the soluble matter in a substance which is not pol is referred to as non-pol (Rouillard, 1979). In the above example, there would be 40 g of non-pol per 100 g of brix. The relationship between brix, pol, non-pol and purity can be seen in figure 3.10 (adapted from SMRI's Essential Cane Sugar Technology booklet, Anon., 2012, page 9).

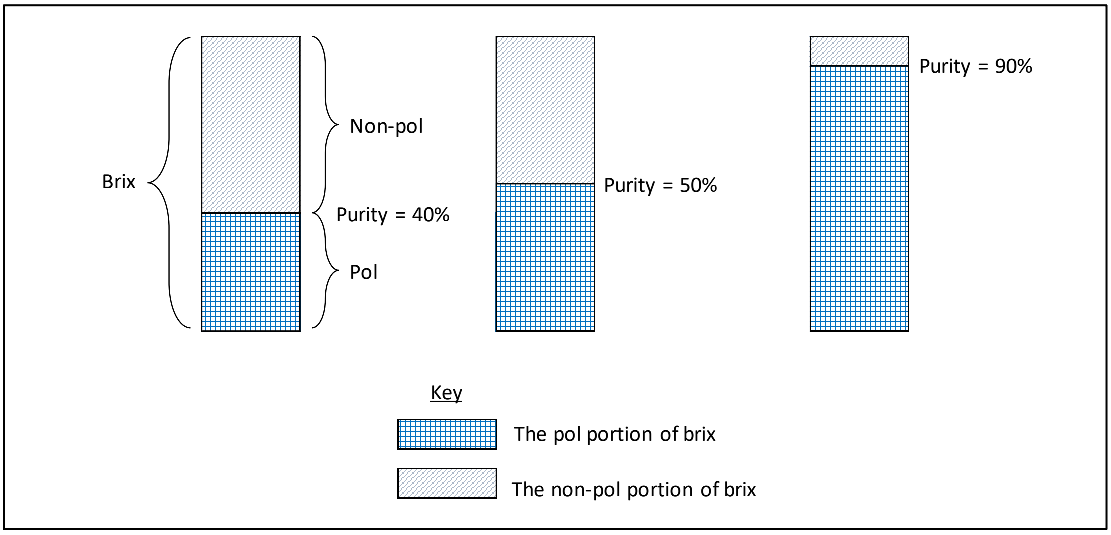

Figure 3.10 Brix, pol and purity relationships

3.7.5 Supersaturation

A solution is termed saturated when no more sucrose will dissolve in it, under constant conditions. If the temperature is raised, more sucrose will dissolve. If the temperature is lowered without allowing crystallisation to occur, the solution will become supersaturated. This creates a state of tension in which the excess dissolved sucrose will precipitate out of solution. If small crystals are added to a supersaturated solution, the crystals will grow due to sucrose being deposited on them (Rein, 2007, page 354). The solubility coefficient is the ratio of the concentration of sucrose in an impure solution to the concentration in a pure solution, both saturated at the same temperature. The concentration is given as a ratio of sucrose to water (Rein, 2007, page 29 and 30). The equation relating the solubility coefficient of sucrose to the (non-sucrose)-to-water ratio in impure solutions is shown in figure 3.11 (Starzak, 2016b). The coefficients for the equation were regressed during the validation of the MATLAB™ model. This equation was used in the Aspen Plus® model in the crystallisation and drying modules.

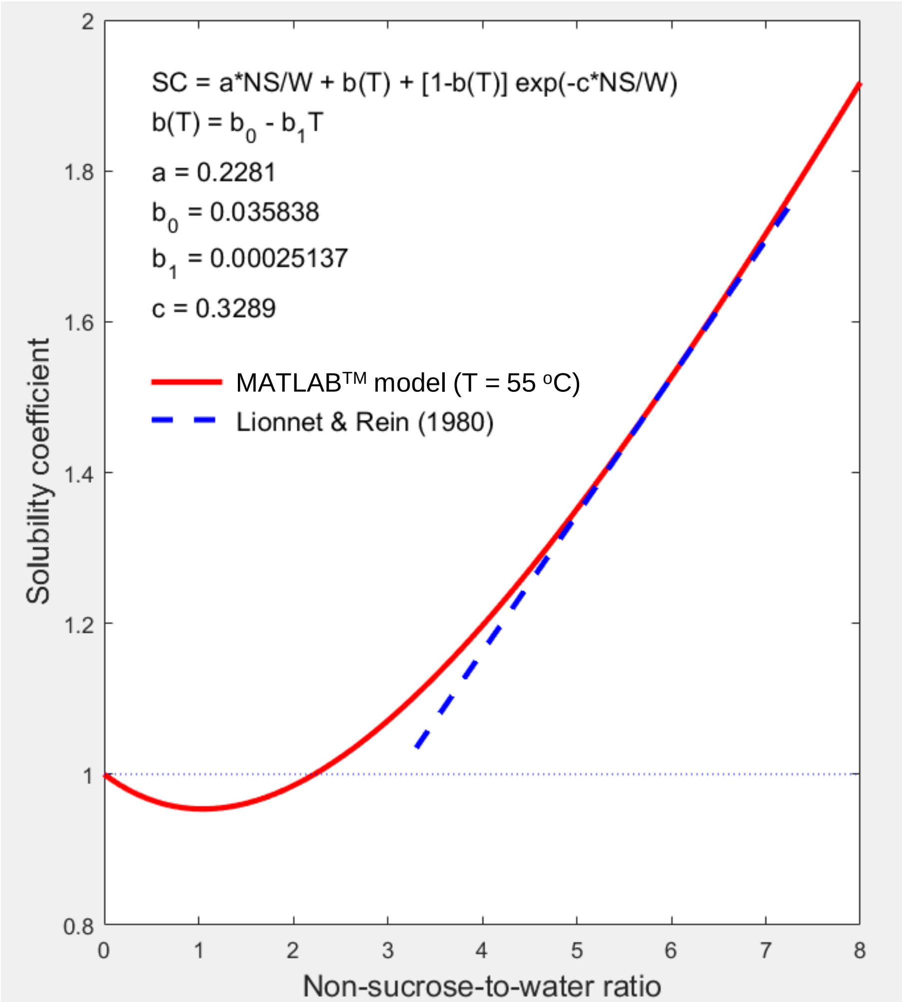

Figure 3.11 Solubility coefficient versus Non-sucrose-to-water ratio (Starzak, 2016b)

3.8 Predicting sugar mill measurements (pol, refractometer brix and apparent purity)

The following data was required in order to compare the Aspen Plus® model to the MATLAB™ model and sugar mill data:

- Refractometer brix
- Pol
- Apparent purity

From the results of the Aspen Plus® model the true sucrose content Suc and the true concentration of dry solids, DS, (and hence the true purity PU) can be calculated. These are generally not reported by sugar mills.

Correlations are thus needed to convert the true values (Suc and DS) to the sugar mill measurements (Pol and $Brix_{ref}$). Two empirical correlations were used in the verification of the MATLAB™ model (Starzak, 2016a). These equations were developed by Hoekstra (Tongaat-Hulett, unpublished results) and verified by Love (2002). Because the Aspen Plus® model was based on the MATLAB™ model, the following equations were needed.

The correlation used to predict DS from $Brix_{ref}$ measurements is:

$$DS = Brix_{ref}\left[1 - 0.00066\left(Brix_{ref} - Pol\right)\right]$$

The correlation used to relate pol readings to Suc (% by weight) is:

$$Suc = Pol + (DS - Pol)H \qquad \text{Where H is Hoekstra's factor.}$$

Hoekstra's correlations assume that the factor H stays constant for any impure sucrose stream. Independent measurements of sucrose content, pol, and refractometer brix data for C-molasses were used to determine the H factor.

These two equations were then rearranged to give the sugar mill measurements (Pol and $Brix_{ref}$) from the true values (Suc and DS):

$$Pol = \frac{Suc - DS \times H}{1 - H}$$

$$Brix_{ref} = \frac{1 + 0.00066 \times Pol - \sqrt{\Delta}}{2 \times 0.00066}$$

Where: $\Delta = (1 + 0.00066 \times Pol)^2 - 4 \times 0.00066 \times DS$

And $H = \frac{Suc_{MOLC} - Pol_{MOLC}}{DS_{MOLC} - Pol_{MOLC}}$

Where: $Suc_{MOLC} = \frac{Sucrose\ \%\ Brix_{ref}}{100} \times Brix_{ref}$

$$Pol_{MOLC} = \frac{PU_{app}}{100} \times Brix_{ref}$$

$$DS_{MOLC} = Brix_{ref} \times (1 - 0.00066 \times \left(Brix_{ref} - Pol_{MOLC}\right)$$

Since there are almost no insoluble solids in 'C' molasses, these equations may be used for streams containing fibre (e.g. draft juice and filter cake) only if the properties are calculated on an insoluble solid-free (fibre-free) basis.

Using data from the MATLAB™ model (table 3.4), the H factor was then calculated.

Table 3.4 Mill data for 'C' molasses

Measured variable	Mill 1	Mill 2	Mill 3	Mill 4	Mill 5	Mill 6	Mill 7	Mean
Refractometer Brix ($Brix_{ref}$)	81.78	84.64	83.25	78.59	82.51	82.41	80.27	81.92
Apparent Purity	36.06	33.99	36.97	36.12	35.58	34.87	37.40	35.86
Sucrose % $Brix_{ref}$	40.14	37.57	39.73	40.96	40.15	39.22	40.24	39.72

3.9 Definition of factory performance indices

In order to verify the Aspen Plus® model, the sugar mill performance indicators were calculated. Most of the indicators were calculated directly from model variables. However, the following performance indices need to be clarified (Starzak, 2016b):

$$\text{Sugar extraction} = \frac{F_{suc}^{DJ}}{F_{suc}^{CANE}} \times 100$$

Where F_{suc}^{DJ} refers to the sucrose content in draft juice (DJ) and F_{suc}^{CANE} refers to sucrose content in the sugar cane feed (CANE). A full list of streams and their abbreviations may be found in Appendix F.

$$\text{Extraction pol factor} = \frac{Pol^{DJ}F^{DJ} + Pol^{BAG}F^{BAG}}{Pol^{CANE}F^{CANE}} \times 100$$

$$\text{Extraction brix factor} = \frac{Brix_{ref}^{DJ} \times F^{DJ} + Brix^{BAG} \times F^{BAG}}{Brix_{ref}^{CANE} \times F^{CANE}} \times 100$$

$$\text{Limestone, tonne/1000 tonnes dry sugar} = 1000 \times \frac{F_{lim}^{LIM}}{F^{SUA}}$$

$$\text{Filter wash index} = \frac{Brix_{ref}^{CJ}}{Brix_{ref}^{FJ}} \times 100$$

A, B, C massecuite (pan), m^3/tonne brix $= \frac{F^i}{\rho^i \times F^{DJ}} \times \frac{100}{Brix_{ref}^{DJ}}$

$\rho\ [t/m^3]$ – Massecuite density

$i =$ PANA, PANB, PANC

The following formula was used for the C-massecuite % crystal content:

C-massecuite % crystal content $= \frac{PU_{app}^{PANC} - PU_{app}^{MOLC}}{100 - PU_{app}^{MOLC}} \times Brix_{ref}^{PANC}$

C-molasses @85 brix % on cane $= \frac{F^{MOLC} \times Brix_{ref}^{MOLC}}{85 \times F^{CANE}} \times 100$

A, B, C-exhaustion index $= \frac{100}{PU_{app}^{i}} \times \frac{PU_{app}^{i} - PU_{app}^{j}}{100 - PU_{app}^{j}} \times 100$

PU_{app} – apparent purity

$i =$ PANA, PANB, PANC

$j =$ MOLA, MOLB, MOLC

Boiling house recovery (BHR) $= \frac{F_{suc}^{SUGA} + F_{cry}^{SUGA}}{F_{suc}^{DJ}} \times 100$

Cane-to-sugar ratio $= \frac{F^{cane}}{F_{suc}^{SUGA} + F_{cry}^{SUGA}}$

The following formula was used for calculating the steam-to-cane ratio:

Steam-to-cane $= \frac{F^{SBF}}{F^{CANE}}$

Stream SBF is the high-pressure steam from the boiler after blowdown losses have been considered.

CHAPTER 4: ASPEN PLUS® MODEL OVERVIEW

In this chapter each module is described and an explanation of how it was modelled in Aspen Plus® is given. The Aspen Plus® flowsheets for each module may be found after the descriptions. The bullet points refer directly to the Aspen Plus® model. Each building block of the model is described in Appendix A.

4.1 Extraction module

4.1.1 Description of module

4.1.1.1 Introduction

Sugarcane needs to be processed in order for the valuable component (sucrose) to be extracted. Cane preparation refers to the process of taking whole stick cane and turning it into a fine mulch. This process is done in two parts: Firstly, the cane stalks are roughly broken up by knives; secondly the cane is shredded into a fine mulch. Juice (a mixture of sucrose and water) can now easily be extracted from the fibrous matter (Rein, 2007, page 79).

4.1.1.2 Cane preparation

Cane arrives at sugar mills (in sticks of about 1 metre length) and is offloaded onto feeder or spiller tables. The cane is then conveyed to one or two sets of cane knives. Steam is used to power a turbine which drives the knives. The cane is roughly broken up by the action of the knives.

- The cane knives are modelled by a duplicator block in Aspen Plus®. Only a single stream leaves the duplicator block (i.e. the inlet stream is copied to the outlet stream). The block is shown because of its corresponding turbine.

The chopped cane is conveyed to a set of hammers which are called shredders. The hammers expose the juice bearing cells by smashing the cane (Rein, 2007, page 79 and 86).

- The shredders are modelled in the same way as the cane knives.
- High pressure steam demands in the cane knives and shredders are related to the cane throughput (described in Appendix B.1.1 and B.1.3).

The cane mulch is now sent to a diffuser for juice extraction.

4.1.1.3 Juice extraction

A diffuser is a solid-liquid separator in which the sucrose is leached out of the cane through a washing process. The cane bed is moved along the length of the diffuser by a conveyor system. The floor of the diffuser has perforations in order for liquid to pass through.

- The diffuser is modelled by a perfectly mixed tank with separation coefficients.
- The separation coefficient for water is determined by a calculator block (described in Appendix B.1.5)

This is a rough approximation of the actual 12-stage counter-current diffuser which exists in real sugar mills. Each stage of the diffuser involves the pumping of juice collected in that stage to the previous stage. This is poured onto the cane bed and increases in sucrose content as it percolates through (Schmidt and Wise, 1956).

The fibre is conveyed all the way through the diffuser and exits after being pressed down by a roller. This exit stream is called megasse.

Steam is injected into the diffuser in order to maintain a high temperature to increase sucrose recovery (Rein, 2007, page 150). Heat also minimizes bacterial action (Ravnö and Purchase, 2005).

- Direct steam injection to the diffuser is proportional to the cane throughput (described in Appendix B.1.7). The steam used to maintain a high temperature in the diffuser is a portion of the steam produced in the first effect evaporator (vapour bleed V1).

Hot water, called imbibition, is added in a counter-current fashion. This washing process helps to displace the juice from the fibre (leaching). The megasse and juice streams leave at opposite ends of the diffuser.

- Imbibition flow rate is proportional to the flow rate of fibre in the megasse stream (described in Appendix B.1.4).

Some of the juice, called scalding juice, is heated by condensing steam. It is then recycled into the diffuser. The scalding juice is poured onto the cane feed to the diffuser in order to raise the temperature of the cane quickly. The heat increases the permeability of unbroken juice cells (Rein, 2007, page 150).

- The scalding juice heater is modelled by a shell-and-tube heat exchanger.
- A portion of the steam produced in the second effect evaporator (Vapour bleed V2) is sent to the scalding juice heater in proportion to the flow rate of cane.

The residence time of the cane in the diffuser is around one hour.

- Heat losses in the diffuser are accounted for by a cooler block placed on the draft juice stream.

The rest of the juice which is called draft juice, is sent to the mixed juice tank for further processing. In practice, the megasse and draft juice leave the diffuser at different temperatures.

The imbibition is pumped backwards through 12 stages losing heat. The scalding juice heater helps to increase the temperature in the front end. However, the draft juice leaves at a temperature of about 60 °C and the megasse leaves at about 64.5 °C.

- To account for the different temperatures, a heater and cooler block where placed after the diffuser on the megasse and draft juice streams, respectively.
- A calculator block handles the heat transfer from the draft juice to the megasse (described in Appendix B.1.8).

4.1.1.4 Megasse dewatering

The megasse is saturated with dilute juice which is pressed out in a series of dewatering mills.

- The dewatering mills are modelled by a separator with split coefficients.
- High pressure steam demand to the dewatering mill is proportional to the flow rate of fibre in the megasse stream.

The dilute juice which is removed from the fibre is called press water. This water, which is recovered from the bottom of the mills, still contains some sucrose and is recycled back to the diffuser.

The removal of moisture from the megasse increases the calorific value of the biomass and thus it burns better in the boilers. The 'dry' megasse is now called bagasse and still contains approximately 50 % moisture (Starzak, 2015).

- A calculator block manipulates the split coefficient in order to maintain a specified bagasse moisture content (described in Appendix B.1.9).

Most of the bagasse is sent to the boiler. A small portion (about 1%) is sent to the clarification module where it is used as a filter aid in the mud filters.

- Heat losses in the dewatering mills are accounted for by a cooler block placed on the press water stream directly after the mills.

4.1.1.5 <u>Press water recycle</u>

The press water from the dewatering mills is pumped into a temporary holdup tank in order to increase its temperature before putting it back into the diffuser. This tank is kept at a constant temperature by directly injecting steam.

- The tank is modelled by a mixer in Aspen Plus®.
- A calculator block determines the amount of steam which is required to maintain the temperature in the tank (described in Appendix B.1.10). Vapour bleed V1 (a portion of the steam produced in the first effect evaporator) is sent to the press water tank. The flow rate of steam is manipulated by the specified temperature of the hot press water.

4.1.1.6 <u>Extraction module mechanical drives</u>

The cane knives, shredders and dewatering mill mechanical drives are modelled by turbines in Aspen Plus®. These turbines are driven by high pressure steam. The boilers supply steam at 31 bar absolute (bara) from the boilers (Starzak, 2016a). The exhaust steam which exits the turbines is at 2 bara (Starzak, 2016a). This exhaust steam is sent to the evaporation module.

- Calculator blocks manipulate the flow rate of steam to the cane knives turbine and shredder turbine based on the flow rate of cane (described in Appendix B.1.1 and B.1.3).
- A calculator block manipulates the flow rate of steam to the dewatering mills turbine based on the flow rate of fibre in the megasse stream (described in Appendix B.1.2).
- A calculator block determines the amount of steam needed by the extraction module mechanical drives by summing the requirements of the three turbines (described in Appendix B.1.3).

4.1.2 Flowsheet

The details of the extraction module are shown in the following flowsheet (Figure 4.1).

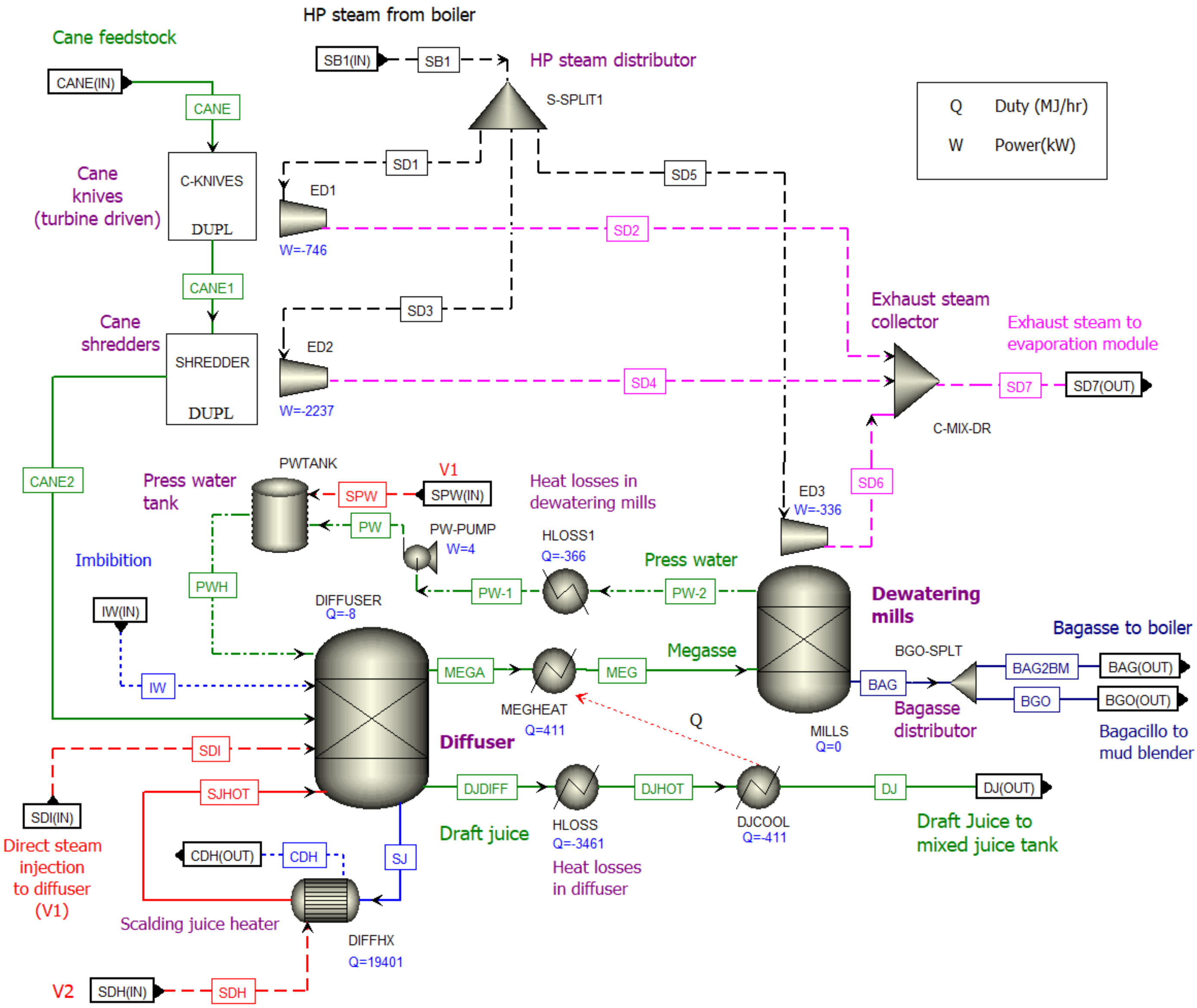

Figure 4.1 Flowsheet of extraction module in Aspen Plus®

IN connectors, eg. CANE(IN), show where streams are coming from an external source into a flowsheet/hierarchy. OUT connectors, eg. DJ(OUT), show where streams leave the flowsheet/hierarchy.

4.2 Clarification module

4.2.1 Description of module

4.2.1.1 Introduction

The juice which is extracted in the diffuser needs to be clarified so as to be free from suspended particles which slip through the screen of the diffuser. Also, impurities in solution are precipitated and the pH is adjusted. The process of clarification forms a mud layer of impurities and suspended solids.

4.2.1.2 Mixed juice tank

Draft juice from the diffuser is weighed before entering a storage tank. Two other recycle streams also flow into this tank: juice recovered from the mud filter and sludge from the syrup clarifier in the evaporation module. The juice in this tank is referred to as mixed juice.

- The mixed juice tank is modelled by a mixer in Aspen Plus®.

4.2.1.3 Mixed juice heating

The mixed juice is heated in order to raise the temperature above the saturation temperature and also to sterilise the juice. The saturated temperature of the mixture at atmospheric pressure is around 100.3 °C. The heating process occurs in a series of heat exchangers, called juice heaters. There are three sets of heaters: primary, secondary and tertiary.

- The mixed juice heaters are modelled by three shell and tube heat exchangers.

Different pressures of saturated steam from the evaporators are sent to these heat exchangers. The highest pressure (with corresponding highest temperature) is sent to the last heat exchanger. The outlet temperature of the mixed juice is specified from each juice heater in the Aspen Plus® model. After the tertiary heater, the mixed juice is at a temperature of 103.9 °C.

- Calculator blocks determine the required flow rate of steam (vapour bleeds V3, V2 and V1) in order to achieve the specified outlet temperatures from each mixed juice heater (described in Appendix B.2.1 – B.2.3).

4.2.1.4 Lime addition

Milk-of-lime, which is a mixture of suspended lime particles in water, is added to the mixed juice after the primary juice heater. The heating process helps to speed up the reaction between lime and juice. Lime is added for two reasons. Firstly, to raise the pH of the mixed juice to limit sucrose inversion. Secondly, to form a precipitate with insoluble particles (Rein, 2007, page 220).

- The addition of lime to the juice is modelled by a mixer.
- The flow rate of lime is manipulated by a specified lime content in the mixed juice. A calculator block determines how much lime to add (described in Appendix B.2.4).

4.2.1.5 Sucrose inversion

Sucrose inversion is the reaction of sucrose to form glucose and fructose in an acidic medium and at high temperatures. The reaction has the following stoichiometry:

$$Sucrose + Water \rightarrow Glucose + Fructose$$

Cane juice naturally has a pH of 4.5 - 5.5, which is less than neutral. Sucrose inversion is of concern at these acidic pH levels. The pH of the mixed juice after the lime is added would be about 7.2 (Davis, 2018), which helps to stop inversion. Adding the lime at this early stage also gives the lime sufficient time to evenly disperse throughout the juice before it reaches the clarifier.

4.2.1.6 Mixed juice pump

The lime-juice mixture is pumped at 3.5 bara to the secondary heater. Pressure isn't taken into account in the Aspen Plus® model until it impacts the simulation. Therefore, up until this point the pressure of the process stream has been assumed to be 1.013 bara (atmospheric).

4.2.1.7 Mixed juice flash

The mixed juice is flashed after the tertiary heater in order to remove gas. The flash vessel is kept at atmospheric pressure. As the juice enters the vessel, the pressure drops rapidly. Since the mixed juice enters at 103.9 °C, which is above saturation, a portion of the water in the juice evaporates quickly (called flashing). Air bubbles and dissolved air get caught up in the vapour stream and are released from the juice (Rein, 2007, page 220).

- The mixed juice flashing process is modelled by a flash vessel with the temperature and pressure specified.

4.2.1.8 Clarifier

The mixed juice is then sent to a clarifier which is basically a settling tank. The lime in the mixed juice reacts with phosphates present in sugar cane. This causes an amorphous calcium phosphate precipitate to form. Some dissolved solids and particulate matter coagulate with the precipitate. Flocculant is added at between 4 and 7 ppm (relative to mixed juice) before entering the clarifier (Blom and Munsamy, 1991). This aids the settling of the precipitate, which is then called mud. A clear juice which is now mainly free from insoluble matter flows out of the clarifier.

- The clarifier is modelled by a separator with split coefficients.
- The clarifier separation coefficient for water is manipulated based on a specification for the mass fraction of water in the mud stream (described in Appendix B.2.5).
- Heat losses in the clarifier are modelled by coolers with a specified temperature drop on the exit streams of mud and clear juice.

The clear juice is pumped from the clarifier at about 2.4 bara to the preheater in the evaporation module.

Over 90 % of the mud stream leaving the clarifier consists of clear juice (Anon., 2012, page 41). This needs to be recovered due to its sucrose content.

4.2.1.9 Mud treatment

Some mills recycle mud to the diffuser. This saves the loss of sucrose in filter cake but also causes filtration problems in the diffuser. The screen of the diffuser may become blocked. The mills which were chosen for the validation of the MATLAB™ model all have mud filtration. A vacuum filtration step recovers as much juice as possible while the cake which forms on the filter is generally sold to farmers or used in animal feed factories.

The mud stream from the clarifier is sent to a blender in which fine bagasse particles (bagacillo) are added. These bagacillo particles aid filtration in the vacuum filter.

- The mud-bagacillo blender is modelled by a mixer.
- A calculator block controls the flow rate of bagacillo to the blender (described in Appendix B.2.6).

The mud-bagacillo mixture is then sent to a vacuum filter. Vacuum draws the liquid portion through a fine mesh screen. The solid portion is deposited on the mesh as a mat of fibre and mud.

- The vacuum filter is modelled by a separator with split coefficients.

Wash water is added to help displace the juice trapped in the fibrous layer. A portion of the hot condensate from the evaporation module is used as wash water. The high temperature minimises microbiological activity in the filter station (Rein, 2007, page 265).

- Calculator blocks determine the wash water flow and water separation coefficient in the filter (described in Appendix B.2.7 and B.2.8). The wash water flow rate to the vacuum filter is proportional to the fibre content of the filter cake. The water separation coefficient is manipulated based on a specified mass fraction of water in the filter cake.

The fibrous mat is scraped off the filter at the end of the vacuum cycle. This is called filter cake and is a by-product of the sugar milling process.

Filtrate juice which is recovered from the vacuum filter is then recycled to the mixed juice tank. Provision has been made for some of the filtrate juice to be diverted for further processing.

4.2.2 Flowsheet

The details of the clarification module are shown in the following flowsheet (Figure 4.2).

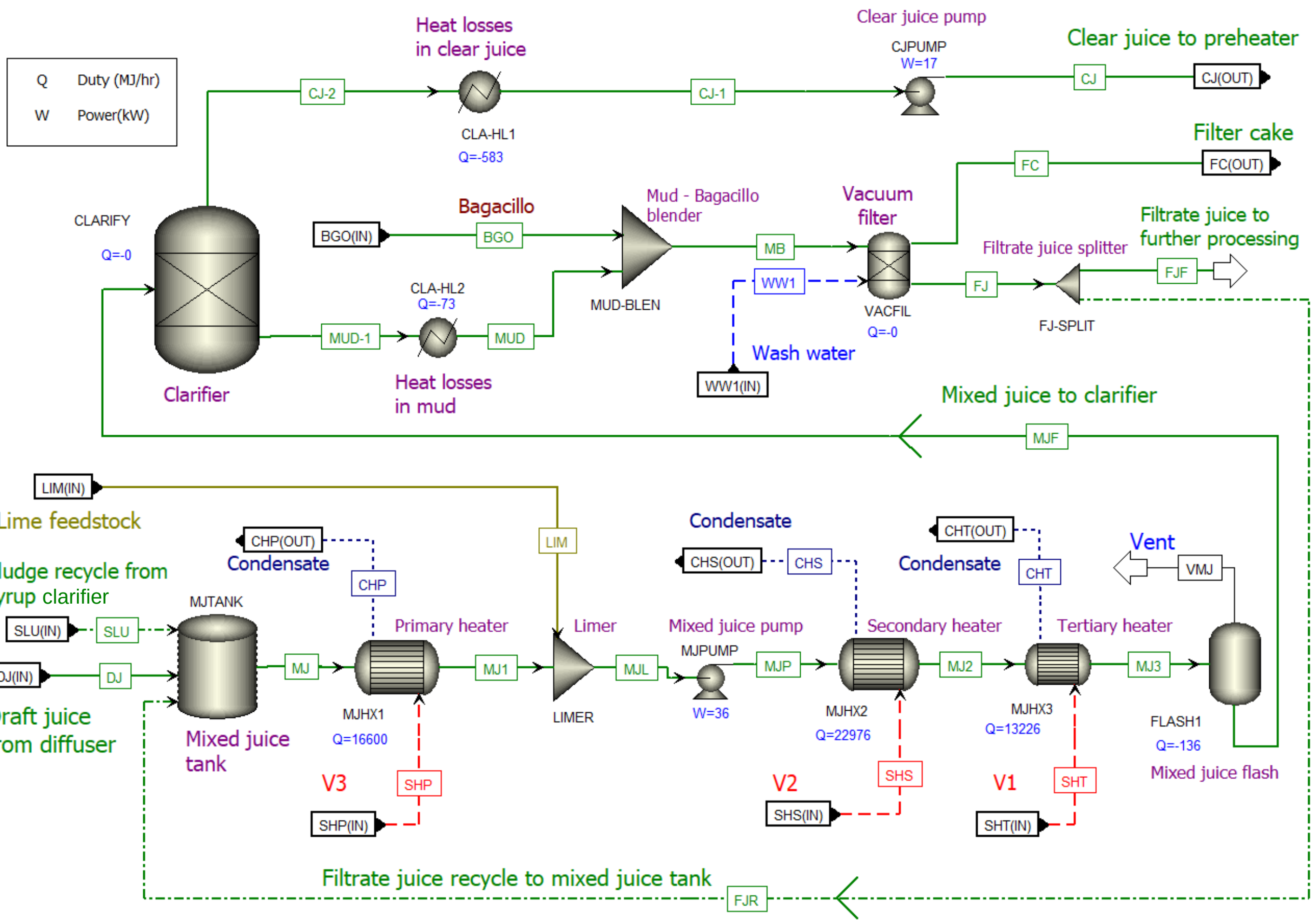

Figure 4.2 Flowsheet of clarification module in Aspen Plus®

4.3 Evaporation module

4.3.1 Description of module

4.3.1.1 <u>Introduction</u>

In order to crystallise sucrose a supersaturated solution is required. This is accomplished in two stages: evaporation and pan boiling. In the evaporation process, water is boiled off from the clear juice. In the pans more water is boiled off to form a supersaturated solution. Small crystals are added and these grow as sucrose is deposited (Rein, 2007, page 354).

4.3.1.2 Evaporation

Clear juice from the clarifier has a very low concentration of dissolved matter (11 % brix). Sucrose is the major component of the dissolved matter in clear juice (85 % apparent purity).

Water is removed by boiling the juice in a series of 4 or 5 evaporator vessels. The Aspen Plus® model assumes 5 effects. The remaining juice after the evaporators, called syrup, has between 62 and 68 % brix (Davis, 2018).

Clear juice is first preheated by exhaust steam in order to get it closer to the boiling point.

- The clear juice preheater is modelled by a shell-and-tube heat exchanger.

Exhaust steam from the turbo-alternator (boiler module) and motor drive turbines (extraction module) is used to heat the clear juice in the preheater.

- A calculator block which uses an Excel® spreadsheet calculation method determines the amount of steam required by the preheater in order for the clear juice to reach a specified temperature before entering the evaporators (described in Appendix B.3.1.1).

The clear juice is then sent to the first effect evaporator. Exhaust steam provides the heat for the juice to boil.

- A calculator block which uses an Excel® spreadsheet calculation method (PI controller) iterates through the amount of exhaust steam which is sent to the first effect in order to ensure sufficient vapour bleed V3 is available to the primary mixed juice heater (described in Appendix B.3.1.3).

The two processes which occur in an evaporator are: steam condensation in the calandria and water evaporation in the tubes of the evaporator.

- The condensing steam is modelled by a cooler block.
- The evaporation of water is modelled by a flash block.

The energy released by the condensing steam is sent to the flash block. Heat losses are accounted for by applying a reduction factor to the energy which goes to the flash vessel.

- A calculator block applies the reduction factor to model the heat lost.

When the water evaporates, some juice droplets get carried up (entrained) in the vapour stream (Rein, 2007, page 304).

- Droplet entrainment is taken into account by specifying a liquid fraction which exits in the vapour stream.

Some of the sucrose is inverted in the evaporators. This occurs due to high temperatures and long residence times (Rein, 2007, page 303).

- Sucrose inversion is modelled by a reactor.
- The following reaction is modelled:

$$Sucrose\ +\ Water\ \rightarrow\ Non-sucrose$$

Note: Fructose and glucose are treated collectively as non-sucrose in the Aspen Plus® model. Both entrainment and inversion are proportional to the flow rate in the various evaporation effects.

4.3.1.3 Multiple-effect evaporation

Juice is boiled in a series of five evaporators. Exhaust steam is only fed to the first evaporator (first effect). Vapour evaporated in the first effect still has heat energy. This vapour is condensed in the second evaporator (second effect). The heat of vapourisation given off by condensation causes the juice in the second effect to boil. This process is repeated in the remaining three evaporators (Rein, 2007, page 273-275).

- The 5 evaporators are modelled by 5 coolers and 5 flash vessels. The pressure distribution in the evaporation effects were assumed. Note: fixing the pressure will cause unrealistic behaviour should scenarios be tested which are considerably different from the tuned operating point.
- Boiling point elevation is correctly predicted by the UNIQUAC thermodynamic model.

4.3.1.4 Vapour bleeding

Some of the vapour produced in the first three effects is tapped (‘bled’) off to be used in other sections of the mill. The vapour which is not ‘bled’ off is sent to the next effect.

Vapour bleeds are sent to the following modules:

- ❖ Extraction module (scalding juice heater, direct steam injection to diffuser, press water tank).
- ❖ Clarification module (mixed juice heaters).
- ❖ Crystallisation module (vacuum pans and "remelter").

The vapour bleed flow rates are manipulated in order to meet desired specifications in these modules (e.g. required temperatures, brix).

- Vapour bleeds are modelled by splitters.
- Calculator blocks manipulate the vapour bleed splitters to provide the required steam.

Valves were used in order to account for hydraulic temperature losses in the steam lines. These valves were placed on the steam feeds (2^{nd} to 5^{th} effects). A fixed pressure drop of 0.02 bara across the valves was assumed (Starzak, 2015).

The vapour to the 4^{th} effect is throttled in order to control the brix of the liquid from the 5^{th} effect (syrup). A 0.15 bara pressure drop was assumed across this valve.

- In Aspen Plus® a design specification handles the throttle process. See Appendix C.1 for details.

4.3.1.5 Vapour recovery from condensates

The vapour bleeds (V1, V2 and V3) condense in other sections of the mill. These condensates still contain some heat energy which can be recovered. This is done by dropping the pressure slightly and thus 'flashing' off some new steam.

- Vapour recoveries from condensates are modelled in flash vessels.

The steam which was recovered is then added to the vapour streams from the evaporators before the vapour bleed splitters.

4.3.1.6 Final condensate uses

The condensate from the steam in the final effect evaporator is sent to a distributor.

- All distributors are modelled by flow splitters in Aspen Plus®.

Portions are sent to the crystallisation module (centrifuges and 'remelter') while the remainder is joined with the liquid stream from the fourth effect vapour recovery flash vessel.

This joined condensate stream is sent to the fifth effect vapour recovery flash vessel. The liquid from this flash vessel is distributed to the following areas:

- ❖ Clarification module (vacuum filter),
- ❖ Extraction module (imbibition to diffuser),
- ❖ The remainder is an effluent stream from the mill and would be sent to an effluent treatment plant which is not modelled.

The vapour from this flash vessel is joined with the vapour from the fifth effect evaporator before being sent to the barometric condenser.

4.3.1.7 Barometric condenser

The vapour from the final effect evaporator is condensed in a barometric condenser. 1 kg of steam occupies 1673 litres, when this is condensed the resulting water only occupies 1 litre (Anon., 2012, page 48). This space contraction produces a vacuum in the last effect vapour space. This vacuum draws through the previous effects leading to a pressure profile. The pressure distribution modelled in Aspen Plus® is 1.6, 1.25, 0.6, 0.4 and 0.16 (bara) in the five effects respectively. As the pressure drops, so too does the saturated temperature of the juice. For this reason, vapour which has been evaporated in the first effect has a higher temperature than the juice in the second effect.

- The barometric condenser after the final stage evaporator is modelled as a condenser with a specified temperature and pressure.

Cold water from the cooling tower is sprayed inside the barometric condenser. This water causes the steam to condense as it comes into contact with the cold droplets. The condenser has a barometric leg in order to maintain a vacuum while still allowing the liquid to pass out (Rein, 2007, page 329).

4.3.1.8 Syrup clarifier

The syrup from the final effect evaporator is clarified before being sent to the crystallisation module. The pressure is changed in the Aspen Plus® model to 1.013 bara before entering the syrup clarifier.

- The syrup clarifier is modelled by a separator with split coefficients.

A small scum layer called sludge is scraped off and recycled to the mixed juice tank.

4.3.2 Flowsheet

The details of the evaporation module are shown in the flowsheet on the next page (Figure 4.3).

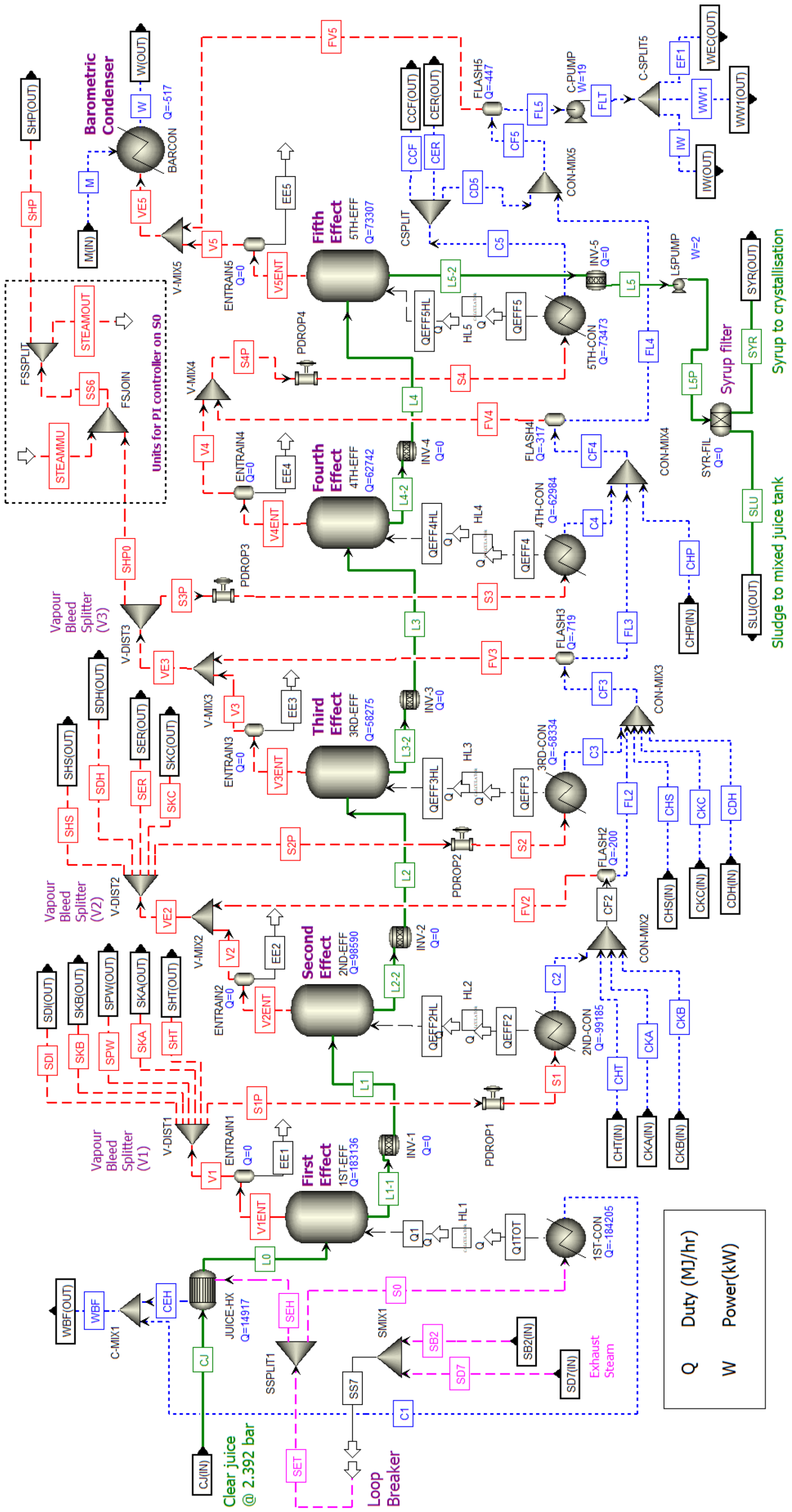

Figure 4.3 Flowsheet of evaporation module in Aspen Plus® (A3 version in Appendix E)

4.4 Crystallisation module

4.4.1 Description of module

4.4.1.1 Introduction

Syrup enters the crystallisation module from the syrup clarifier in the evaporation module. The goal of crystallisation is to recover as much of the sucrose at as high a purity as possible from this syrup stream.

Sucrose is crystallised in large vessels called 'vacuum pans'. The boiling house configuration consists of 3-boilings in pans referred to as 'A', 'B' and 'C' pans.

After boiling the syrup in the 'A' pans, 'A' massecuite is formed. This is a mixture of sugar crystals and mother liquor (the contents of the pan besides sugar crystals) (Rein, 2007, page 28).

This massecuite is cooled in large crystallisers, where crystal growth continues. The massecuite is then sent to centrifuges. This is where the separation of sugar crystals from mother liquor occurs. The mother liquor passes through small perforations and is now called 'A' molasses (Rein, 2007, page 29).

This molasses still has a large quantity of sucrose in it. It is sent to the 'B' pans where some of this sucrose is recovered (by crystallisation). The 'B' massecuite goes through crystallisers and centrifuges as well to form a 'B' sugar and 'B' molasses.

The 'B' molasses is sent to the 'C' pans for a final boiling. A final molasses is obtained after the 'C' massecuite has gone through crystallisers and centrifuges. This molasses is a by-product of sugar mills.

In the following sections, the individual unit operations of the crystallisation module are described.

4.4.1.2 Syrup distributor

After evaporation, syrup is sent to the 'A' pans and the magma mingler. The purpose of the magma mingler is to provide seed crystals for the 'A' pans (Rein, 2007, page 364).

- The process of distributing the syrup is modelled by a flow splitter.

4.4.1.3 Vacuum pans

The process of crystallisation usually occurs in both batch and continuous pans, however in Aspen Plus® it is modelled as a continuous process. The pans are operated under vacuum in the same manner as the final effect evaporator. Each pan has its own barometric condenser which condenses the evaporated vapour. However, in Aspen Plus® a centralised condenser is modelled as it does not affect the mass and energy balances.

- The 'A', 'B' and 'C' vacuum pans are modelled by five units each, namely:
    1. Condenser to model the steam condensation process in the calandria.
    2. Flash vessel to model the evaporation of water from the syrup in the pans.
    3. Stoichiometric reactor to model the crystallisation process.
    4. Separator to model droplet entrainment.
    5. Stoichiometric reactor to model sucrose inversion.

Two calculator blocks govern the following processes in the vacuum pans:

1. Droplet entrainment (described in Appendix B.4.1).
2. Crystallisation (described in Appendix B.4.2). The extent of crystallisation for the batch pans and cooling crystallisers is determined by the SLE model – Vavrinecz equation (Vavrinecz, 1962; 1965) and solubility coefficient equation (van der Poel, 1998; Rein, 2007).

- Vapour bleed flow rates (V1 and V2) to the pans are manipulated by specified dry solids content in the exit massecuites.

4.4.1.4 Cooling crystallisers

Massecuites from the pans are sent to cooling crystallisers. The 'A' and 'C' crystallisers are water cooled whilst the 'B' crystalliser is aircooled. This adequately represents the current sugar mill configurations in South Africa (Starzak 2016a). The temperature, pressure and supersaturation are controlled in the exit massecuite stream from each crystalliser. Sucrose continues to crystallise as the temperature drops.

The 'A' and 'C' cooling crystallisers are modelled by:

- Shell-and tube heat exchangers for the cooling process.
- Stoichiometric reactors to model the crystallisation of sucrose (described in Appendix B.4.3).

The air-cooled 'B' crystallisers are modelled by:

- Two units: a cooler and stoichiometric reactor.

The air stream to the 'B' crystallisers is not modelled.

4.4.1.5 Centrifuges

After the crystallisers, the massecuites are sent to centrifuges. Wash water is added to remove the film of molasses which covers the sugar. Adding water causes some dissolution of crystals. A few small crystals go with the molasses through the perforations (Anon., 2012, page 62).

- Crystal dissolution and loss is modelled by a stoichiometric reactor.
- Separation of sugar from molasses is modelled by a separator with split coefficients.

4.4.1.6 Magma mingler

A portion of the 'B' sugar is mixed with some syrup in a magma mingler. Syrup which is unsaturated is added to the 'B' sugar in the mingler and this causes crystals to partially dissolve.

- Crystal loss is modelled by a stoichiometric reactor (described in Appendix B.4.4).
- Syrup addition to 'B' sugar is modelled by a mixer. Syrup flow rate to the magma mingler is manipulated based on a specified moisture content in the magma (described in Appendix C.5).

4.4.1.7 Partial 'remelt'

All of the sugar from the 'C' pans and the rest of the 'B' sugar which was not sent to the mingler is 'remelted'. Steam and water are added to the remelter causing all crystals to dissolve. This remelt stream is recycled to the 'A' pans in order to recover more sucrose since 'C' sugar has a purity which is too low to be marketed.

- Crystal dissolution is modelled by a stoichiometric reactor with a 100 % conversion.

- The stream mixing of steam, condensate, 'B' sugar and 'C' sugar is modelled by a mixer.

4.4.2 *Flowsheet*

The details of the crystallisation module are shown in the following flowsheet (Figure 4.4).

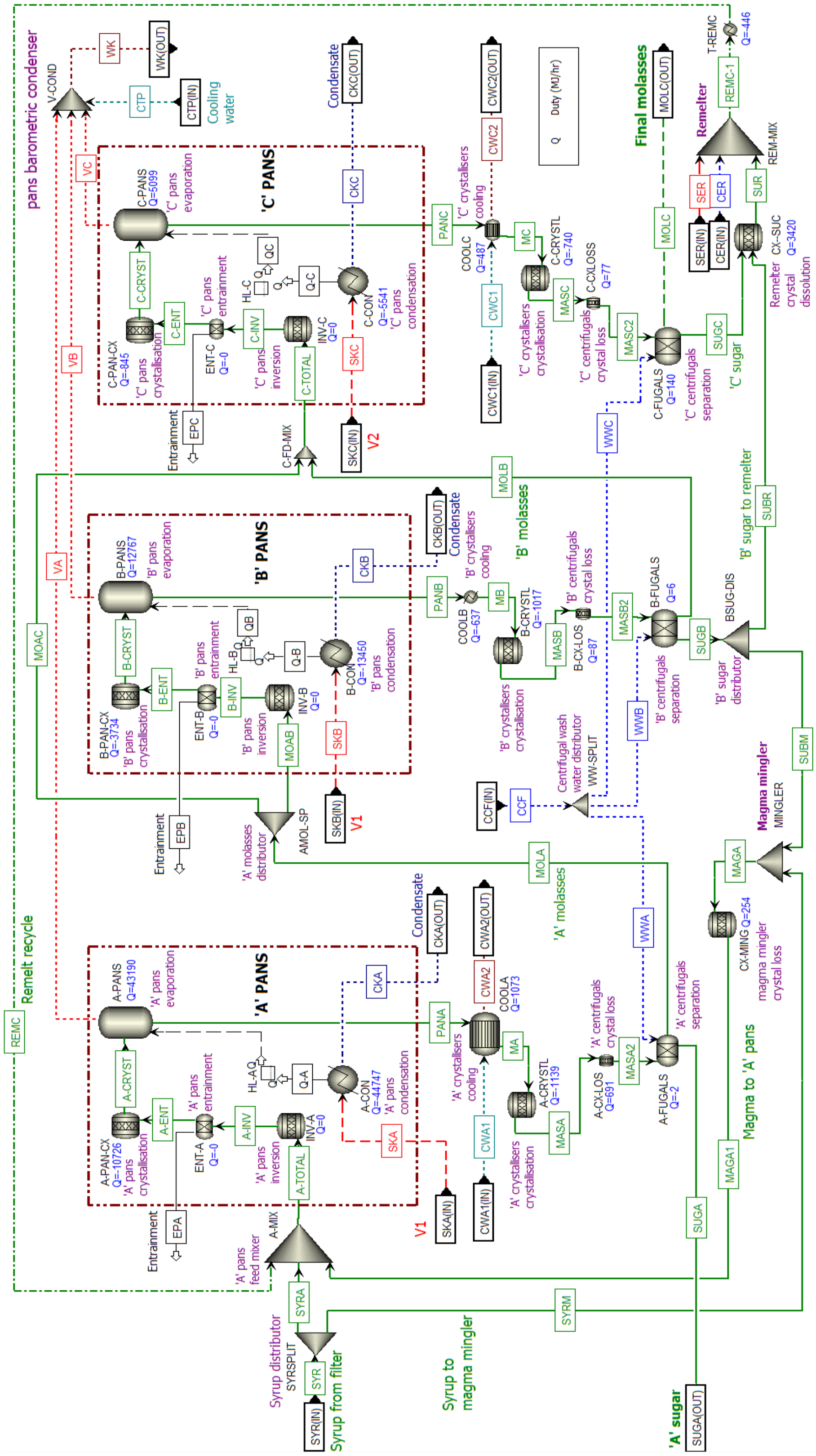

Figure 4.4 Flowsheet of crystallisation module in Aspen Plus® (A3 version in Appendix E)

4.5 Drying module

4.5.1 Description of module

4.5.1.1 Introduction

Raw sugar from the 'A' centrifuges has to be dried in order to improve its keeping quality. Even though it has a low moisture content (0.5 – 2 %), deterioration occurs in the film of molasses which covers each crystal (Anon., 2012, page 66).

Heated air is forced around the sugar crystals in a drier to remove the moisture. Crystallisation of sucrose occurs at the surface. This forms a 'skin' which prevents bound moisture from being released (Anon., 2012, page 66).

4.5.1.2 Sugar driers

The type of drier which is modelled is a rotary louvre drier. This is a horizontal drum which has a slight downward tilt towards the sugar outlet. As the drum rotates, louvres pick up the sugar and drop it into the air flow (Anon., 2012, page 66).The drier is separated into two sections: the first section of the drier uses heated air and the second section uses ambient air to cool the sugar.

- The sugar dryer is modelled by two shell and tube heat exchangers (for heating and cooling), two separators (for moisture separation) and two mixers (for moisture mixing).
- Calculator blocks determine the extent of drying (described in Appendix B.5.2 and B.5.3). Moisture separation in the sugar dryer is calculated based on specifications for moisture contents after the different sections of the dryer. The specification for the moisture content after the cooling section was optimised in the regression of the MATLAB^{TM} model. The resulting value was higher than the moisture content in the heating section (see Appendix D, parameters 508 and 510).
- A stoichiometric reactor unit is used to model crystallisation which occurs in the drying process (described in Appendix B.5.4). Crystallisation in the dryer is determined by the SLE model.

Sugar leaving the dryer has a moisture content below 0.1 %.

4.5.1.3 Air heater

The air is heated by exhaust steam before entering the dryer.

- The air heater is modelled by a shell and tube heat exchanger (described in Appendix B.5.1). The exhaust steam flow rate to the dry air heater is calculated based on a specified exit temperature of the air.

4.5.2 Flowsheet

The details of the drying module are shown in the following flowsheet (Figure 4.5).

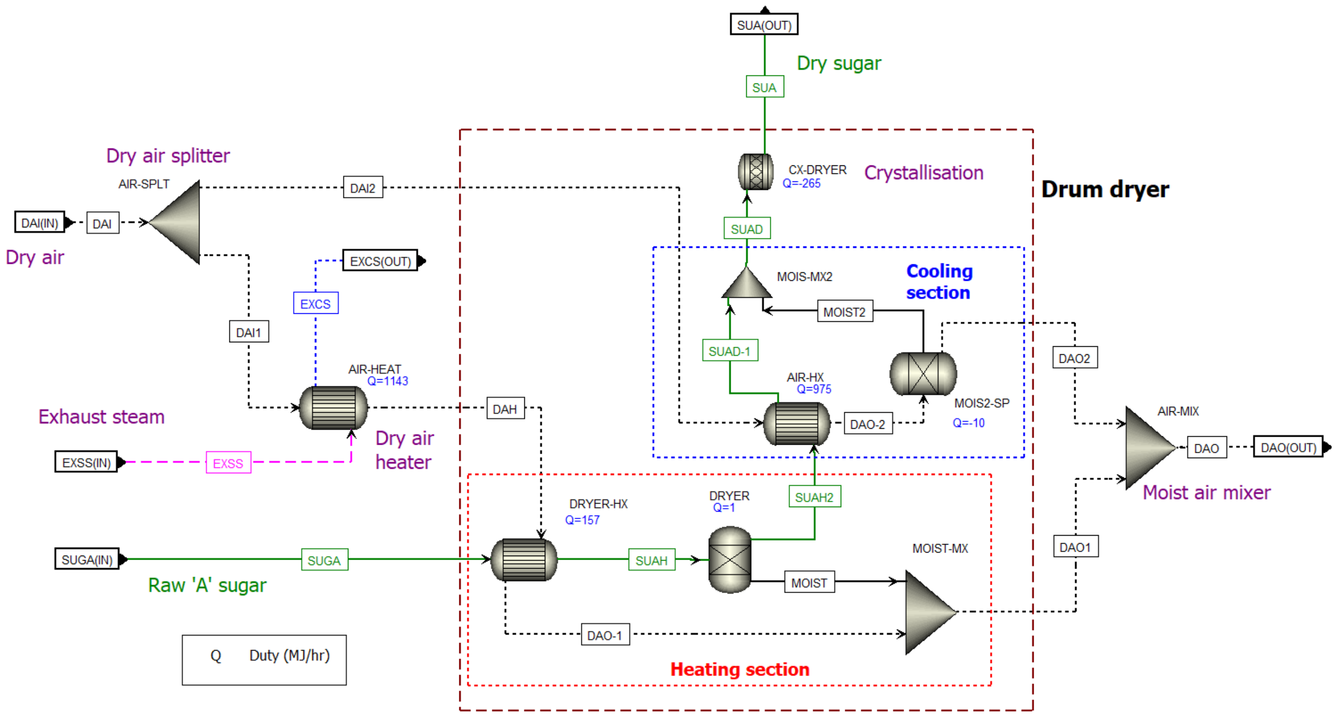

Figure 4.5 Flowsheet of drying module in Aspen Plus®

4.6 Boiler module

4.6.1 Description of module

4.6.1.1 <u>Boiler</u>

Bagasse is burnt in large boilers in order to generate superheated steam at 31 bara and 390 °C. This steam, called 'live' steam (Anon., 2012, page 79), is used in the turbo-alternators to generate electricity and in the extraction module mechanical drives (cane knives, shredders and mills).

The outlet from these turbines is called exhaust steam and is at a pressure of 2 bara and 121 °C. The exhaust steam is used in the dryer module (air heater), evaporation module (preheater and first effect evaporator) and boiler module (deaerator). The condensate of the exhaust steam is then sent back to the boiler in order to make high pressure steam again.
An assumed amount of bagasse per kg of steam generated (0.45 kg/kg steam) was specified (Starzak 2016a). Combustion reactions in the boiler are not considered.

- The boiler is modelled by a heater with a fixed outlet pressure and temperature.
- A calculator block determines the amount of bagasse required by the boiler (described in Appendix B.6.1). The bagasse required by the boiler is proportional to the demand for high pressure (31 bara) steam.

Some boiler water is lost as blowdown (0.2 %). This is an important part of boiler maintenance in order to purge some of the suspended solids from the system (Rein, 2007, page 662-663).

- A distributor models the boiler water blowdown. The flow rate of boiler blowdown is proportional to the flow rate of boiler feed water.

4.6.1.2 Live steam usage

Live steam is sent to the turbo-alternator and turbines for cane knives, shredders and dewatering mills in the extraction module. A small portion is also lost due to leaks, venting, cleaning and start-up demands. The following equation is used to estimate the amount of steam lost (Starzak, 2016a):

$$F^{SBL} = 0.1(F^{CANE})^{0.67}$$

Where F^{SBL} is the flow rate of steam lost and F^{CANE} is the flow rate of cane feed. The units are tonnes per hour.

- The live steam splitter (which accounts for steam losses) is modelled as a distributor.
- A calculator block determines the steam loss (described in Appendix B.6.3). Live steam losses are calculated based on the flow rate of cane.

A turbo-alternator generates electricity for the sugar mill.

- The turbo-alternator is modelled by a turbine with a fixed discharge steam pressure (2 bara) and specified efficiency (0.856).

4.6.1.3 Turbo-alternator exhaust steam distribution

The exhaust steam from the turbo-alternator is distributed to the boiler water deaerator (not modelled) as well to the dryer module and evaporation module.

- The turbo-alternator exhaust steam distributor is modelled by a flow splitter.

It was assumed that the flow rate of exhaust steam to the deaerator is proportional to the flow rate of live steam consumption (The ratio of deaerator flow to live steam is 2%) (Starzak, 2016a).

- Make-up water to the boiler is calculated as a summation of live steam losses, boiler blowdown and the exhaust steam flow to the deaerator.

4.6.2 Flowsheet

The details of the boiler module are shown in the following flowsheet (Figure 4.6).

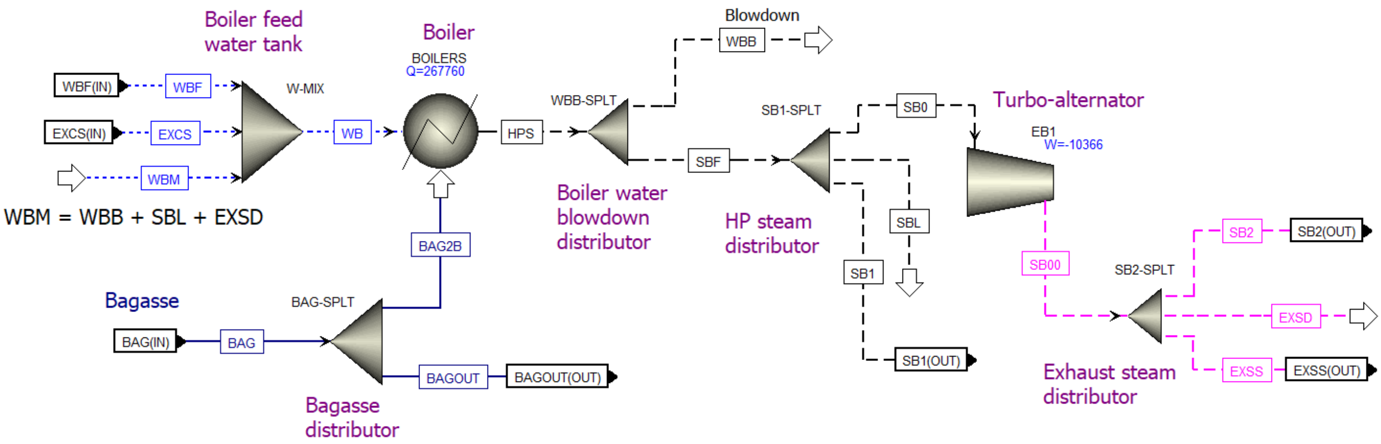

Figure 4.6 Flowsheet of boiler module in Aspen Plus®

4.7 Cooling tower module

4.7.1 Description of module

Cold water is required in various parts of the sugar mill. This water is used to condense steam and to cool down massecuites.

The following units require cold water:

- Barometric condenser of the evaporation module
- Barometric condenser of pans
- Cooling crystallisers ('A' and 'C')

The warm water which returns from these units is sent to a cooling tower. Spray pond cooling towers are generally used in sugar mills. Warm water is sprayed into the air causing some of the water to evaporate. Due to evaporation the remaining water is cooled. An effluent stream is taken off before the cooling tower in order to maintain a constant flow rate in the cooling water cycle.

- The cooling tower is modelled by a flash vessel with a fixed temperature.
- Heat loss to the environment is modelled by a cooler block.

The cool water is then distributed to where it is needed.

4.7.2 Flowsheet

The details of the cooling tower module are shown in the following flowsheet (Figure 4.7).

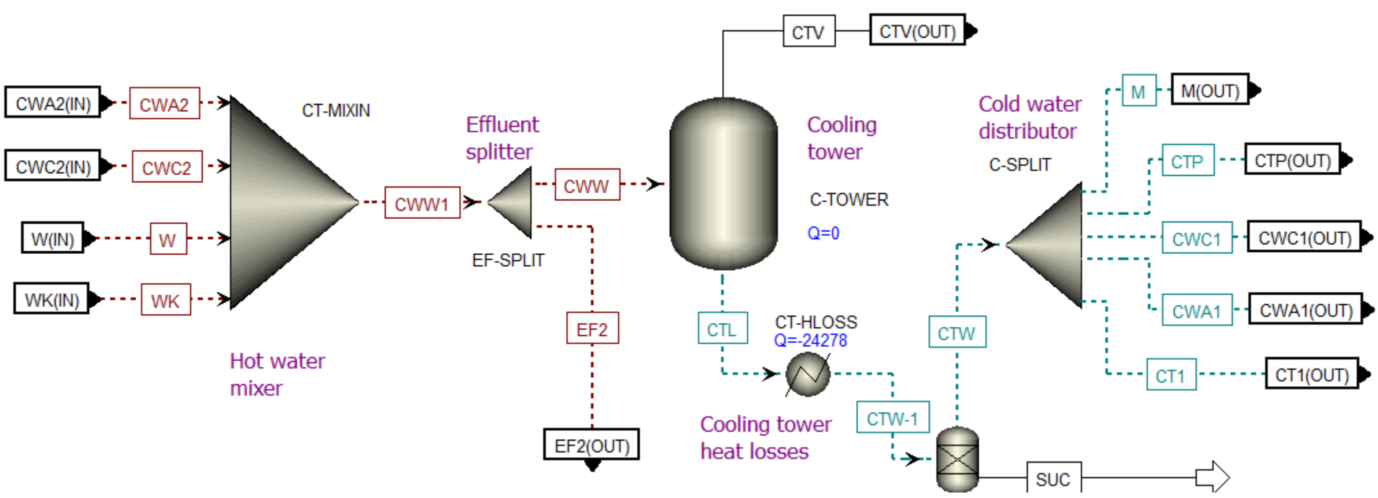

Figure 4.7 Flowsheet of cooling tower module in Aspen Plus®

CHAPTER 5: RESULTS AND DISCUSSION

5.1 Introduction

In this chapter the Aspen Plus® model results are presented and discussed. Firstly, there is a direct comparison of the stream properties between the Aspen Plus® model and the MATLAB™ model. This was done for a throughput of 244.18 t/h (the same conditions as the validation of the MATLAB™ model). Various points from the development of the Aspen Plus® model are discussed after the results.

In section 5.3 there is a comparison of the sugar mill performance indices. The Aspen Plus® model results were compared with the MATLAB™ model and factory data. A brief look at the heat transfer areas of the evaporator station may be found in section 5.4.

In section 5.5 the predictive capabilities of the Aspen Plus® model are presented. Discrete scenarios with potential biorefinery applications were chosen and the Aspen Plus® results were compared to the MATLAB™ model for the same input conditions.

5.2 Direct comparison of Aspen Plus® model results

The Aspen Plus® model was verified at each stage of development by comparing the results with the MATLAB™ model. Selected stream results are shown for the Aspen Plus® model in tables 5.1 - 5.6. Temperatures, pressures, flow rates and compositions are shown as well as calculated properties (Dry Solids, pol, refractometer brix, apparent purity etc.).

The extraction module results are shown in table 5.1. Volumetric flow rates were presented for the Aspen Plus® model in order to provide data for potential biorefinery feedstocks. The calculated properties which are on a fibre-free basis means that fibre was discarded before the calculations. The true properties (sucrose and dry solids) were calculated directly from model results whereas the mill measurements (pol and refractometer brix) were calculated based on the equations presented in section 3.8.

Table 5.1 Comparison of selected streams of the extraction module showing compositions and properties (MATLAB™ vs. Aspen Plus®)

Model:	MATLAB™	Aspen Plus®	MATLAB™	Aspen Plus®	MATLAB™	Aspen Plus®
Stream:	Cane	Cane	Draft Juice	Draft juice	Bagasse	Bagasse
Stream property						
Temperature (°C)	27.0	27.0	60.0	60.0	64.5	64.5
Pressure (bara)	1.01	1.01	1.01	1.01	1.01	1.01
Mass Flow (kg/h)	244180	244180	279360	279358	75320	75318
Volume Flow (m^3/h)		202.9		278.0		47.1
Mass Flows (kg/h)						
WATER	167334	167334	239440	239448	38383	38382
SUCROSE	34602	34602	33467	33477	1122	1124
NON-SUCROSE	5469	5469	5336	5349	121	119
FIBRE	36776	36776	1090	1083	35694	35693
LIME	0	0	0	0	0	0
CRYSTAL	0	0	0	0	0	0
Mass Fractions						
WATER	0.6853	0.6853	0.8571	0.8571	0.5096	0.5096
SUCROSE	0.1417	0.1417	0.1198	0.1198	0.0149	0.0149
NON-SUCROSE	0.0224	0.0224	0.0191	0.0191	0.0016	0.0016
FIBRE	0.1506	0.1506	0.0039	0.0039	0.4739	0.4739
LIME	--	--	--	--	--	--
CRYSTAL	--	--	--	--	--	--
Calculated property	Cane	Cane	Draft Juice	Draft juice	Bagasse	Bagasse
Sucrose (%)	14.17	14.17	11.98	11.98	1.49	1.49
Sucrose (Fibre-free basis)	16.68	16.68	12.03	12.03	2.83	2.84
Dry Solids (Fibre-free basis)	16.41	16.41	13.89	13.90	1.65	1.65
Pol (%)	14.02	14.02	11.85	11.85	1.48	1.48
Refractometer Brix (Fibre-free basis)	19.36	19.36	13.96	13.97	3.14	3.14
Brix (refractometer)	16.44	16.44	13.91	13.92	1.65	1.65
Apparent purity (%)	85.26	85.26	85.19	85.17	89.62	89.73

As shown in table 5.1 there is excellent agreement between the results of the Aspen Plus® model and the MATLAB™ model in the extraction module. Calculator blocks were used where possible to directly calculate required flow rates of streams (eg. high pressure steam demands to the extraction module mechanical turbines and imbibition water flow rate).

Selected results from the clarification module are shown in table 5.2 on the next page.

Table 5.2 Comparison of selected streams of the clarification module showing compositions and properties (MATLAB™ vs. Aspen Plus®)

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	Lime	Lime	Clear Juice	Clear Juice	Filtrate juice	Filtrate juice	Filter cake	Filter cake
Stream property								
Temperature (°C)	20.0	20.0	99.8	99.9	94.0	93.9	90.7	93.9
Pressure (bara)	1.013	1.01	2.39	2.39	1.01	1.01	1.01	1.01
Mass Flow (kg/h)	7660	7656	281670	281669	33960	33959	9110	9111
Volume Flow (m^3/h)		15.3		294.0		36.7		15.8
Mass Flows (kg/h)								
WATER	6894	6890	243138	243139	28863	28861	6377	6377
SUCROSE	0	0	33322	33333	4140	4139	225	225
NON-SUCROSE	0	0	5211	5198	798	798	174	174
FIBRE	0	0	0	0	0	0	1570	1569
LIME	766	766	0	0	160	160	765	766
CRYSTAL	0	0	0	0	0	0	0	0
Mass Fractions								
WATER	0.9000	0.9000	0.8632	0.8632	0.8499	0.8499	0.7000	0.7000
SUCROSE	--	--	0.1183	0.1183	0.1219	0.1219	0.0247	0.0247
NON-SUCROSE	--	--	0.0185	0.0185	0.0235	0.0235	0.0191	0.0191
FIBRE	--	--	--	--	--	--	0.1723	0.1723
LIME	0.1000	0.1000	--	--	0.0047	0.0047	0.0840	0.0840
CRYSTAL	--	--	--	--	--	--	--	--
Calculated property	Lime	Lime	Clear Juice	Clear Juice	Filtrate juice	Filtrate juice	Filter cake	Filter cake
Sucrose (%)	0	0	11.83	11.83	12.19	12.19	2.47	2.47
Sucrose (Fibre-free basis)	0	0	11.83	11.83	12.19	12.19	2.98	2.98
Dry Solids (Fibre-free basis)	0	0	13.68	13.68	14.54	14.54	4.38	4.37
Pol (%)	0	0	11.70	11.71	12.03	12.02	2.31	2.30
Refractometer Brix (Fibre-free basis)	0	0	13.70	13.70	14.63	14.63	5.90	5.89
Brix (refractometer)	0	0	13.70	13.70	14.63	14.63	4.88	4.88
Apparent purity (%)	--	--	85.45	85.48	82.18	82.17	47.23	47.19

From table 5.2 it can be seen that the results compare well between the two models in the clarification module. Clear juice compositions agree to the fourth decimal place. Filter cake temperature is different between the two models due to a specification of temperature difference between the filter cake and filtrate juice being implemented in the MATLAB™ model.

Selected results from the evaporation module are shown in table 5.3 on the next page.

Table 5.3 Comparison of selected streams of the evaporation module showing compositions and properties (MATLAB™ vs. Aspen Plus®)

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	L1	L1	L2	L2	L3	L3	L4	L4	L5 (syrup)	L5 (syrup)
Stream property										
Temperature (°C)	113.8	113.5	106.6	106.3	86.8	86.6	77.1	77	58.7	58.6
Pressure (bara)	1.6	1.6	1.25	1.25	0.6	0.6	0.4	0.4	0.16	0.16
Mass Flow (kg/h)	198370	198364	151350	151306	120750	120742	91740	91733	58350	58191
Volume Flow (m^3/h)		206.2		152.6		116.2		83.8		46.8
Mass Flows (kg/h)										
WATER	160144	160130	113270	113220	82750	82745	53824	53820	20533	20464
SUCROSE	33227	33034	33055	32873	32977	32790	32898	32714	32629	32563
NON-SUCROSE	5237	5201	5237	5213	5241	5207	5229	5199	5187	5165
FIBRE	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0	0	0	0	0	0	0	0
CRYSTAL	0	0	0	0	0	0	0	0	0	0
Mass Fractions										
WATER	0.8073	0.8073	0.7484	0.7483	0.6853	0.6853	0.5867	0.5867	0.3519	0.3517
SUCROSE	0.1675	0.1665	0.2184	0.2173	0.2731	0.2716	0.3586	0.3566	0.5592	0.5596
NON-SUCROSE	0.0264	0.0262	0.0346	0.0345	0.0434	0.0431	0.057	0.0567	0.0889	0.0888
FIBRE	--	--	--	--	--	--	--	--	--	--
LIME	--	--	--	--	--	--	--	--	--	--
CRYSTAL	--	--	--	--	--	--	--	--	--	--
Calculated property	L1	L1	L2	L2	L3	L3	L4	L4	L5 (syrup)	L5 (syrup)
Sucrose (%)	16.75	16.65	21.84	21.73	27.31	27.16	35.86	35.66	55.92	55.96
Dry Solids (%)	19.37	19.27	25.26	25.17	31.59	31.47	41.46	41.33	64.81	64.83
Pol (%)	16.57	16.47	21.61	21.49	27.02	26.86	35.48	35.28	55.32	55.36
Refractometer Brix (%)	19.40	19.31	25.33	25.23	31.69	31.57	41.63	41.50	65.24	65.26
Apparent purity (%)	85.41	85.31	85.31	85.17	85.26	85.10	85.22	85.01	84.79	84.82

Table 5.3 shows good comparison between the results. Overall stream flow rates are very similar between models. In the Aspen Plus® model a bit less sucrose remains in the liquid from the 1st effect evaporator (L1). The component flow rate of sucrose in the MATLAB™ model is 0.58 % more. This difference continues through the effects. The flow rate of syrup from the final effect evaporator is 0.27 % less in the Aspen Plus® model.

Selected results from the crystallisation module are shown in table 5.4 on the next page.

Table 5.4 Comparison of selected streams of the crystallisation module (Part 1) showing compositions and properties (MATLAB™ vs. Aspen Plus®)

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	'A' pans outlet	'A' pans outlet	'A' sugar	'A' sugar	'A' molasses	'A' molasses
Stream property						
Temperature (°C)	67.1	67.1	56.2	56.2	60.2	60.2
Pressure (bara)	0.16	0.17	1.01	1.01	1.01	1.01
Mass Flow (kg/h)	58430	58483	29080	29081	33280	33324
Volume Flow (m^3/h)		40.4		18.5		25.6
Mass Flows (kg/h)						
WATER	4856	4867	236	236	8543	8553
SUCROSE	19492	19522	628	629	18031	18066
NON-SUCROSE	6842	6843	137	137	6706	6706
FIBRE	0	0	0	0	0	0
LIME	0	0	0	0	0	0
CRYSTAL	27234	27251	28080	28078	0	0
Mass Fractions						
WATER	0.0831	0.0832	0.0081	0.0081	0.2567	0.2567
SUCROSE	0.3336	0.3338	0.0216	0.0216	0.5418	0.5421
NON-SUCROSE	0.1171	0.117	0.0047	0.0047	0.2015	0.2012
FIBRE	--	--	--	--	--	--
LIME	--	--	--	--	--	--
CRYSTAL	0.4661	0.466	0.9656	0.9655	--	--
Calculated property	'A' pans outlet	'A' pans outlet	'A' sugar	'A' sugar	'A' molasses	'A' molasses
Sucrose (%)	79.97	79.98	98.72	98.71	54.18	54.21
Dry Solids (%)	91.69	91.68	99.19	99.19	74.33	74.33
Pol (%)	79.17	79.18	98.69	98.68	52.81	52.84
Refractometer Brix (%)	92.50	92.49	99.23	99.22	75.46	75.46
Apparent purity (%)	85.59	85.61	99.46	99.46	69.99	70.03
Mother liquor (ML) flow (kg/h)	31196	31232	1000	1003	33280	33324
Sucrose in ML (%)	62.48	62.5	62.79	62.74	54.18	54.21
Dry solids in ML (%)	84.42	84.42	76.45	76.44	74.33	74.33
True purity in ML (%)	74.02	74.04	82.13	82.08	72.89	72.93

From table 5.4 it can be seen that the results compare excellently between models in the 'A' station of the crystallisation module. Aspen Plus® calculates that the pressure in the 'A'pans is 0.17 bara as opposed to 0.16 bara in the MATLAB™ model. Table 5.5 on the next page shows some more results from the crystallisation module as well as the final dry sugar stream.

Table 5.5 Comparison of selected streams of the crystallisation module (Part 2) and final dry sugar stream (SUA) showing compositions and properties (MATLAB™ vs. Aspen Plus®)

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	'B' molasses	'B' molasses	'C' molasses	'C' molasses	Final dry sugar	Final dry sugar
Stream property						
Temperature (°C)	50.7	50.7	56.1	56.1	35.9	35.9
Pressure (bara)	1.01	1.01	1.01	1.01	1.01	1.01
Mass Flow (kg/h)	11520	11519	10890	10878	28860	28867
Volume Flow (m^3/h)		8.4		8.1		18.2
Mass Flows (kg/h)						
WATER	1524	1522	2277	2272	23	23
SUCROSE	4306	4304	3534	3529	78	79
NON-SUCROSE	4760	4759	5078	5077	139	137
FIBRE	0	0	0	0	0	0
LIME	0	0	0	0	0	0
CRYSTAL	931	933	0	0	28620	28628
Mass Fractions						
WATER	0.1323	0.1322	0.2091	0.2089	0.0008	0.0008
SUCROSE	0.3738	0.3736	0.3245	0.3244	0.0027	0.0027
NON-SUCROSE	0.4132	0.4132	0.4663	0.4667	0.0048	0.0048
FIBRE	--	--	--	--	--	--
LIME	--	--	--	--	--	--
CRYSTAL	0.0808	0.0810	--	--	0.9917	0.9917
Calculated property	'B' molasses	'B' molasses	'C' molasses	'C' molasses	Final dry sugar	Final dry sugar
Sucrose (%)	45.46	45.47	32.45	32.44	99.44	99.45
Brix (%)	86.77	86.78	79.09	79.11	99.92	99.92
Pol (%)	42.65	42.66	29.28	29.27	99.41	99.41
Refractometer Brix (%)	89.54	89.56	81.94	81.96	99.96	99.96
Apparent purity (%)	47.63	47.64	35.74	35.71	99.45	99.46
Mother liquor (ML) flow (kg/h)	10589	10585	10890	10878	240	239
Sucrose in ML (%)	40.67	40.66	32.45	32.44	32.53	33.07
Dry solids in ML (%)	85.62	85.62	79.08	79.11	90.36	90.49
True purity in ML (%)	47.50	47.49	41.03	41.01	36.00	36.54

Table 5.5 shows excellent comparison between the results in the rest of the crystallisation module. Also, the final dry sugar flow rate has only a difference of 7 kg/h between models.

5.2.1 Convergence complexities

An example of the complexity is illustrated in the interaction between the evaporator and clarification modules. Process steam is sent from the evaporators to the clarification module. This is affected by cane throughput, composition and other process parameters of the clarification module. Condensates from the mixed juice heaters in the clarification module are then returned to the evaporator module where they are flashed at a reduced pressure in order to recover more vapour which joins the vapour produced in the following evaporator effects.

The Wegstein method was used to converge most of the Aspen Plus® simulation. The Newton method was used to initially converge the filtrate juice recycle in the clarification module. The secant method was used to converge the design specifications.

The MATLAB™ model is a purely mathematical model with only one unique solution of a set of simultaneous equations. The tolerance in the design specifications in the Aspen Plus® model means that slightly different values for the steam flow rate may be calculated in successive simulations. This also occurs in different parts of the sugar mill model and this contributes to differences between the MATLAB™ and Aspen Plus® results.

The process parameters are listed in Appendix D: table D.1. Process parameters shown in red are different to the MATLAB™ model. The diffuser water extraction coefficient is different due to the scalding juice stream being shown in Aspen Plus®. Heat losses were implemented in a different manner in Aspen Plus®. Temperature drops were assumed in order to match the MATLAB™ results.

5.2.2 Pressure specifications

Sugar cane is fed to the sugar mill via conveyor belts at atmospheric pressure. The following points list the pressure specifications of the model:

- The cane knives and shredders are both operated at atmospheric pressure.
- Draft juice from the diffuser was assumed to be at atmospheric pressure.
- The press water from the dewatering mills is pumped to the press water tank at 2 bara.
- Mixed juice from the Limer is pumped at 3.5 bara to the secondary mixed juice heater.
- After the tertiary heater the mixed juice is flashed at atmospheric pressure.
- The clear juice and mud from the clarifier are assumed to be at atmospheric pressure.
- Clear juice is pumped at 2.4 bara to the preheater in the evaporation module.

- Pressure drops (0.02 bara) are accounted for in the evaporators to take into account hydraulic losses.
- The pressure distribution in the five evaporator effects was assumed: 1.6; 1.25; 0.6; 0.4 and 0.16 (bara), with the lowest pressure being in the last effect.
- An assumed pressure drop of 0.15 bara was taken into account across the vapour throttle valve. This was accounted for in the vapour to the 3rd effect in order to match the MATLAB^{TM} results, however this should be on the vapour to the 4th effect since that is the final vapour bleed.
- The syrup was assumed to be at atmospheric pressure.
- Temperatures were specified in the batch pans, the pressures were calculated by flash calculations.
- Massecuites from the pans were assumed to be at atmospheric pressure.
- The sugar and molasses streams were assumed to be at atmospheric pressure.
- The exit condensate and cooling water from the barometric condenser in the evaporation module was assumed to be at atmospheric pressure.
- The cooling water from the sump of the barometric condenser in the crystallisation module was assumed to be at 3 bara in order to match the MATLAB^{TM} model results.
- Cooling water is distributed from the cooling tower at 3 bara.
- The boiler was assumed to operate at 31 bara.
- The exhaust steam was specified at 2 bara.

5.2.3 Steam conditions available to biorefinery

The following stream specifications would be available to a potential biorefinery:

- High pressure steam at 31 bara 390 °C.
- Exhaust steam at 2 bara and 121 °C.
- Vapour bleed (V1) at 1.6 bara and 113.8 °C.
- V2 at 1.25 bara and 106.6 °C.
- V3 at 0.6 bara and 86.7 °C.

5.2.4 Proportional-Integral (PI) controller

Direct implementation of the MATLAB^{TM} model calculation procedure led to an over-specified vapour bleed splitter in the evaporation module shown in Figure 5.6. The flow rates to both the primary mixed juice heater and the fourth effect evaporator are controlled by external factors. This leads to a situation where the vapour bleed splitter is overspecified (DOF is -1). The flow rate to the fourth effect is controlled by a design specification (see Appendix C.1) which means it has to be a specified variable of the vapour bleed splitter. Therefore, the flow rate to the primary mixed juice heater may not be specified in Aspen Plus®.

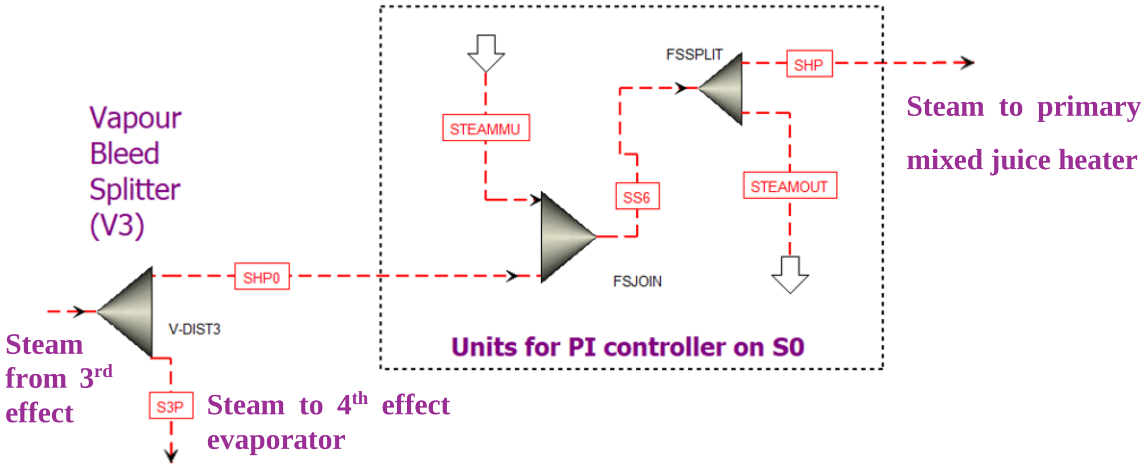

Figure 5.1 Over-specified vapour bleed splitter and fictitious streams (STEAMMU and STEAMOUT) in Aspen Plus®

To overcome this problem, fictitious streams were created in order to ensure the correct steam flow rate to the primary mixed juice heater. To illustrate how this works, if insufficient steam is available from the vapour bleed splitter for the primary mixed juice heater then 'make-up steam' is added by a fictitious stream (STEAMMU). However, if there is too much steam going to the primary mixed juice heater then steam is removed in fictitious stream (STEAMOUT).

A Microsoft Excel® spreadsheet with a PI controller (Love, 2017) was linked to Aspen Plus®. The spreadsheet reads the flow rate of stream STEAMOUT on every iteration. An error integral is then calculated. The controller solves for the exhaust steam flow to the 1st effect evaporator to minimize the error between the fictitious streams (STEAMMU and STEAMOUT). i.e. They cancel out.

5.2.5 *Flash vessels*

The mixed juice flash vessel has pressure and temperature specified in order to match the MATLAB™ results for the vent stream.

The evaporators are modelled as a combination of units with a flash vessel to model the separation of vapour and liquid phases. These flash vessels only have the pressure specified. The heating duty to the flash vessels is provided by condensing steam.

The vapour recovery flash vessels have specified operating pressures. The duty was set to take into account heat losses and match the MATLAB™ results for the vapour flow rates.

The flash vessels of the pans have the temperature specified as parameters of the model. The heat duty is provided by condensing steam and thus the pressure in the flash vessel is calculated by Aspen Plus®.

5.2.6 *Heat losses*

Temperature drops were specified in the diffuser, dewatering mills and clarifier. The heat loss in the diffuser was chosen to give the correct megasse temperature. In the evaporators and batch pans a reduction factor was applied to the energy transferred from the condensing steam in order to take into account heat losses.

5.2.7 *Biorefinery applications*

This model of a sugar mill may be expanded to a sugarcane biorefinery, where the viability of add-on downstream processes to alternative or additional products can be investigated. The filtrate juice splitter provides the option to divert some filtrate juice to downstream processing. Currently, all the filtrate juice is recycled to the mixed juice tank. This filtrate juice could be sent to a fermentor to aid production of high value products. One of the goals for the Aspen Plus® model, was to see what happens to the raw sugar mill when intermediate streams are diverted.

5.2.8 *Pan boiling*

The process of crystallisation usually occurs in a batch and continuous mode, however in Aspen Plus® it is modelled as a continuous process. This approximation doesn't compromise the validity of the mass and energy balances.

5.2.9 Inversion reaction

In the Aspen Plus® model the extent of inversion was calculated based on the MATLAB™ results for the evaporators and pans. The coefficients of the reaction were calculated using a molecular weight of 204 g/mol for non-sucrose. Thus, the reaction becomes:

$$Sucrose + Water \rightarrow 1.766\, Non - sucrose$$

5.2.10 Solid-Liquid equilibria

Starzak (2016b) determined that the calculation of the crystallisation module was mainly affected by the solubility coefficient equation. Replicating the calculations of the crystallisation module in Aspen Plus® was done by means of calculator blocks. For the sake of simplicity and understanding, the calculation method was chosen to be Excel®. An example of the calculation procedure may be seen in Appendix B.4.2.

5.2.11 Modelling of centrifuges

A simplified version of the MATLAB™ centrifuge model was adopted for use in Aspen Plus®. Separation coefficients were determined from the results of the MATLAB™ model. This was deemed to be an adequate representation for the purpose of the Aspen Plus® model since changes in configuration would have little effect on the centrifuges. I.e. diverting half the clear juice stream for use in a biorefinery would have negligible changes on the separation of molasses from crystal sugar in the centrifuges.

5.2.12 Sugar drying module

The cooling stage of the rotary drum dryer is currently modelled as increasing the moisture in the sugar. This is due to the process parameters in the MATLAB™ model. The hot 'A' sugar moisture content, was specified as being 0.065 % (wet basis). The cold 'A' sugar moisture content, was optimised during validation to be 0.0788 % (wet basis). This was subsequently fixed in the MATLAB™ model in 2017, however a decision was made in September 2016 to use the latest version of the MATLAB™ model available for verification of the Aspen Plus® model. This was due to the MATLAB™ code being changed on a weekly basis as errors were fixed. This illustrates that the process models need to be carefully checked to ensure the integrity of the results.

5.2.13 Recycle convergence

Break points (F Chikava and K Foxon, 2017) were added in order to aid recycle convergence in Aspen Plus®. A break point is when a stream is split into two, and some external means of solving the balance is considered. A break point was placed in the filtrate juice recycle (between the vacuum filter and mixed juice tank). Individual spreadsheets with PI controllers were created to converge the individual components in the stream. A break point was also added on the exhaust steam distributor in the evaporation module. Since the steam is in a closed cycle (condensation, boiling, cooling) the inlet and outlet converge naturally to the same value.

5.2.14 Calculator blocks versus design specifications

Calculator blocks can accomplish many advanced tasks in Aspen Plus®. For example: Temperatures after the mixed juice heaters are specified in the model. Calculator blocks determine how much steam is needed to reach the required temperature. However, the calculation fails when the results of the steam and condensate enthalpies are not available.

Design specifications use successive substitution to manipulate the flow rate of steam within boundaries in order to sample the temperature, until the temperatures match the desired outcome. The more design specifications are added the longer it takes for simulations to converge.

5.3 Comparison with factory data and MATLAB™ results

The cane characteristics and performance indices calculated from the results of the Aspen Plus® model are shown in Table 5.6. They have been compared to the MATLAB™ model, with the factory data shown for reference. The standard deviations shown are for the data from the seven South African sugar mills with the average value shown.

Table 5.6 Performance indices (Factory data vs. MATLAB™ vs. Aspen Plus®)

Cane characteristics	Factory	Standard deviation	Matlab™ model	Aspen® model
Cane flow rate t/h	244.18	70.41	244.18	244.18
Cane sucrose %	14.19	0.35	14.17	14.17
Cane pol	14.04	0.36	14.02	14.02
Cane refractometer brix	16.67	0.34	16.44	16.44
Cane apparent purity (DAC)	85.31	0.55	85.27	85.27
Cane fibre %	15.12	1.24	15.06	15.06

	Performance index	Factory	Standard deviation	Matlab™ model	Aspen® model
1	Sugar extraction	96.77	0.44	96.75	96.75
2	Bagasse pol	1.47	0.2	1.48	1.48
3	Bagasse moisture %	50.86	1.54	50.96	50.96
4	Bagasse fibre %	46.75	1.79	47.39	47.39
5	Imbibition % on fibre	307.71	37.25	295.38	295.38
6	Extraction pol factor	98.76	0.82	100	100
7	Extraction brix factor	100.23	0.73	99.94	99.94
8	Draft juice % on cane	113.79	6.13	114.41	114.41
9	Draft juice refractometer brix	14.02	0.69	13.92	13.92
10	Draft juice apparent purity	85.13	0.46	85.17	85.17
11	Draft juice true purity	86.21	0.42	86.22	86.22
12	Draft juice suspended solids, % DJ	0.45	0.37	0.39	0.39
13	Limestone, tonne/1000 tonnes dry sugar	26.65	2	26.53	26.52
14	Clear juice refractometer brix	13.76	0.73	13.70	13.70
15	Clear juice apparent purity	85.13	0.82	85.49	85.48
16	Filtrate apparent purity	82.73	2.59	82.63	82.17
17	Filter cake % on cane	3.25	2.65	3.73	3.73
18	Filter cake pol	2.33	0.87	2.34	2.30
19	Filter cake moisture %	70.37	4.11	70.00	70.00
20	Filter wash index	101.96	3.35	98.42	93.60
21	Syrup refractometer brix	65.34	4.13	65.25	65.26
22	Syrup apparent purity	85.1	0.78	84.83	84.82
23	A-massecuite (pan), m^3/tonne DJ brix	1.05	0.09	0.99	1.04
24	A-massecuite refractometer brix (pan)	92.64	0.26	92.5	92.49
25	A-massecuite apparent purity (pan)	85.68	0.72	85.6	85.61
26	A-molasses apparent purity	69.55	1.89	70	70.03
27	B-massecuite (pan), m^3/tonne DJ brix	0.4	0.05	0.37	0.40
28	B-massecuite refractometer brix (pan)	94.64	0.46	94.65	94.67
29	B-massecuite apparent purity (pan)	69.74	1.16	69.58	69.62
30	B-molasses apparent purity	47.24	1.44	47.65	47.64
31	C-massecuite (pan), m^3/tonne DJ brix	0.27	0.03	0.26	0.29
32	C-massecuite refractometer brix (pan)	96.96	0.61	96.93	96.96
33	C-massecuite apparent purity (pan)	54.2	1.33	53.97	53.97
34	C-massecuite crystal content % (pan)	27.71	1.98	27.47	27.54
35	C-molasses @ 85 brix % on cane	4.34	0.3	4.3	4.30
36	C-molasses refractometer brix	81.92	1.98	81.93	81.96
37	C-molasses apparent purity	35.86	1.17	35.76	35.71
38	Remelt apparent purity	85.49	0.64	85.54	85.56
39	A molasses-massecuite ML true purity diff.(cr)	1	0.5	1.15	1.15
40	B molasses-massecuite ML true purity diff.(cr)	1	0.5	0.98	0.98
41	C molasses-massecuite ML true purity diff.(cr)	1	0.5	1.02	1.02
42	A-pan massecuite temperature, °C	67	3	67.12	67.12
43	B-pan massecuite temperature, °C	67	3	66.88	66.88
44	C-pan massecuite temperature, °C	67	3	66.91	66.91

	Performance index (continued)	Factory	Standard deviation	Matlab™ model	Aspen® model
45	A-crystalliser massecuite temperature, °C	55	3	56.23	56.23
46	B-crystalliser massecuite temperature, °C	50	3	49.86	49.86
47	C-crystalliser massecuite temperature, °C	45	3	45.22	45.22
48	A-exhaustion index	61.66	4.01	60.77	60.73
49	B-exhaustion index	61.14	1.95	60.21	60.29
50	C-exhaustion index	52.7	3.03	52.52	52.62
51	Dry A-sugar pol	99.41	0.08	99.41	99.41
52	Dry A-sugar moisture %	0.08	0.02	0.08	0.08
53	Boiling house recovery	85.65	1.9	85.74	85.75
54	Cane-to-sugar ratio	8.48	0.31	8.4	8.51
55	Steam-to-cane ratio	0.43	0.04	0.4	0.40

As shown in Table 5.7, the performance indices calculated from the results of the Aspen Plus® model compare favourably well with both the MATLAB™ model and factory data. Most of the performance indices produced by the Aspen Plus® model fall within the standard deviation of the factory data. The filter wash index was calculated wrongly in the MATLAB™ model. Instead of using the brix of the filtrate juice, the brix of draft juice was used. When the correct value was used the filter wash index becomes 93.61 for the MATLAB™ model which is almost identical to the value of 93.61 calculated for the Aspen Plus® model.

5.4 Energy related considerations

The area required for heat transfer in the five effects of the evaporators may be calculated by rearranging the following equation:

$$Q = UA\Delta T \qquad \text{where:} \quad \Delta T = T_{condensing\ steam} - T_{flash\ vessel}$$

The heat duty (Q) of the condensing steam was used. The following table (5.7) shows the overall heat transfer coefficients, proposed by Love (1999), which were assumed in the five effects.

Table 5.7 Overall heat transfer coefficients for evaporators (Love et al. 1999)

Effect	Overall heat transfer coefficients ($\frac{W}{m^2.K}$)
1	2500
2	2500
3	2000
4	1500
5	700

The heat transfer areas for the five effects of the evaporator station are shown in table 5.8.

Table 5.8 Overall heat transfer areas for evaporators

Effect	U ($\frac{W}{m^2.K}$)	Q (MJ/h)	T_1 (°C)	T_2 (°C)	ΔT (°C)	A (m^2)
1	2500	184196	120.2	113.8	6.4	3192
2	2500	99188	112.9	106.6	6.3	1749
3	2000	58337	101.8	86.7	15.1	536.6
4	1500	62984	85.1	77.1	8.0	1458
5	700	73473	74.6	58.6	16.0	1822

5.5 Predictive model capabilities

One of the key objectives for building the Aspen Plus® model was to investigate different discrete scenarios of a sugar mill.

The different scenarios which were tested were:

- Increasing imbibition flow rates (section 5.5.1).
- Simulating a higher and lower cane throughput (section 5.5.2).
- Varying cane purity (section 5.5.3).
- Diverting a portion of the clear juice to a potential biorefinery (section 5.5.4).

5.5.1 The effects of changing imbibition flow rate

The effect of increased imbibition flow rates on bagasse usage was considered. As was expected, the higher the flow rate of imbibition water the greater the usage of bagasse by the boilers. The results are shown in Figure 5.1 on the next page.

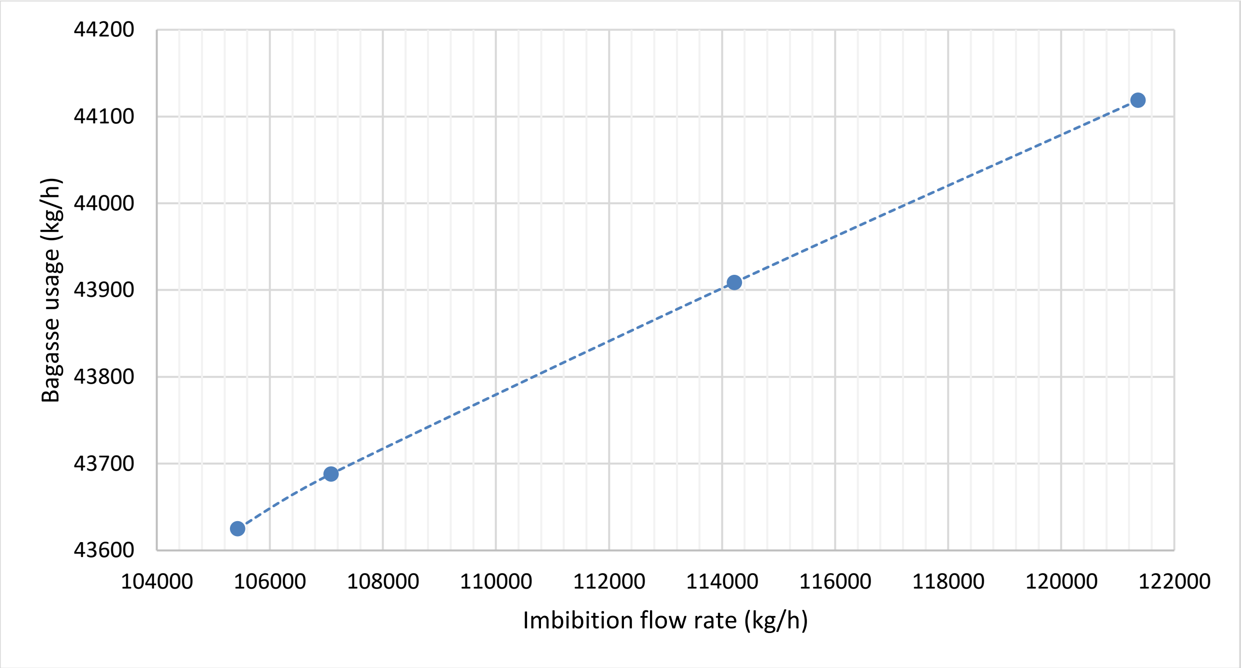

Figure 5.2 Model predictions of bagasse usage for increasing imbibition flow rates

There is no relationship between imbibition flow rates and extraction programmed into the model. So the only effect which was considered was how the bagasse usage by the boiler increases as the imbibition flow rate increases. This demonstrates the principle that the more water added to the system, the more water has to be evaporated in order to reach supersaturation.

5.5.2 Different cane throughputs

Comparison of results between Aspen Plus® and MATLAB™ for a cane throughput of 230 t/h are shown in Table 5.9 on the next page.

Table 5.9 Aspen Plus® and MATLAB™ results for a cane throughput of 230 t/h

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	Cane	Cane	Draft juice	Draft juice	Clear juice	Clear juice	Filter cake	Filter cake
Pressure (bara)	1.01	1.01	1.01	1.01	2.39	2.39	1.01	1.01
Temperature (°C)	27	27	59.4	60	99.8	99.9	90.7	93.9
Flow rate (tonnes/h)	230	230	262.98	263.14	265.16	265.32	8.58	8.58
	|		|		|		|	
Water (%)	68.53	68.53	85.71	85.71	86.31	86.32	70	70.00
Sucrose (%)	14.17	14.17	11.99	11.98	11.84	11.83	2.47	2.47
Non-sucrose (%)	2.24	2.24	1.92	1.91	1.85	1.85	1.91	1.91
Fibre (%)	15.06	15.06	0.39	0.39	0	0	17.23	17.23
Lime (%)	0	0	0	0	0	0	8.4	8.40
Crystal (%)	0	0	0	0	0	0	0	0

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	Syrup	Syrup	Dry sugar	Dry sugar	Final molasses	Final molasses	Boiler bagasse	Boiler bagasse
Pressure (bara)	0.160	0.16	1.01	1.01	1.01	1.01	1.01	1.01
Temperature (°C)	58.7	58.6	35.9	35.9	56.1	56.1	65.5	64.3
Flow rate (tonnes/h)	54.82	54.91	27.19	27.18	10.25	10.35	41.11	41.30
	|		|		|		|	
Water (%)	35.18	35.28	0.08	0.08	20.91	21.60	50.96	50.96
Sucrose (%)	55.95	55.86	0.27	0.27	32.45	32.21	1.49	1.49
Non-sucrose (%)	8.87	8.86	0.48	0.48	46.63	46.18	0.16	0.16
Fibre (%)	0	0	0	0	0	0	47.39	47.39
Lime (%)	0	0	0	0	0	0	0	0
Crystal (%)	0	0	99.17	99.17	0	0	0	0

Table 5.9 shows that there is good agreement between the Aspen Plus® model and MATLAB™ model results for a cane throughput of 230 t/h.

Comparison of results between Aspen Plus® and MATLAB™ for a cane throughput of 270 t/h are shown in Table 5.10.

Table 5.10 Aspen Plus® and MATLAB™ results for a cane throughput of 270 t/h

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	Cane	Cane	Draft juice	Draft juice	Clear juice	Clear juice	Filter cake	Filter cake
Pressure (bara)	1.01	1.01	1.01	1.01	2.39	2.39	1.01	1.01
Temperature (°C)	27	27	61	60	99.8	99.9	90.7	93.9
Flow rate (tonnes/h)	270.00	270.00	309.20	308.90	311.80	311.44	10.08	10.07
	|		|		|		|	
Water (%)	68.53	68.53	85.73	85.71	86.34	86.32	70	70
Sucrose (%)	14.17	14.17	11.97	11.98	11.82	11.83	2.46	2.46
Non-sucrose (%)	2.24	2.24	1.91	1.91	1.84	1.84	1.91	1.91
Fibre (%)	15.06	15.06	0.39	0	0	0	17.22	17.23
Lime (%)	0	0	0	0	0	0	8.41	8.40
Crystal (%)	0	0	0	0	0	0	0	0

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	Syrup	Syrup	Dry sugar	Dry sugar	Final molasses	Final molasses	Boiler bagasse	Boiler bagasse
Pressure (bara)	0.16	0.16	1.01	1.01	1.01	1.01	1.01	1.01
Temperature (°C)	58.7	58.6	35.9	35.9	56.1	56.1	62.5	64.4
Flow rate (tonnes/h)	64.35	64.24	31.91	31.90	12.03	11.91	48.23	47.00
	|		|		|		|	
Water (%)	35.18	35.10	0.08	0.08	20.91	20.00	50.96	50.96
Sucrose (%)	55.95	56.03	0.27	0.27	32.45	32.89	1.49	1.49
Non-sucrose (%)	8.87	8.89	0.48	0.48	46.63	47.10	0.16	0.16
Fibre (%)	0	0	0	0	0	0	47.39	47.39
Lime (%)	0	0	0	0	0	0	0	0
Crystal (%)	0	0	99.17	99.17	0	0	0	0

Deviations between the results of the models increase as the cane throughput varies further away from 244 t/h. Tolerances potentially introduce a difference between models. Where MATLAB has everything specified to the 4th decimal place, the Aspen Plus® model has some tolerances in the specifications in order to aid convergence.

An example of this is the moisture content of the final syrup. In Aspen Plus® the moisture content has been specified as 0.352 with a 0.001 tolerance meaning that the water mass percentage can vary between 35.1 and 35.3%. This may be seen in the results from tables 5.9 and 5.10: for a cane throughput of 230 t/h the syrup moisture content converged to 35.28 % while for a cane throughput of 270 t/h the moisture content converged to 35.10 %.

5.5.3 The effects of varying cane purity

The effect of varying cane purity on the syrup purity is shown in figure 5.3 and the effect on the boiling house recovery is shown in figure 5.4.

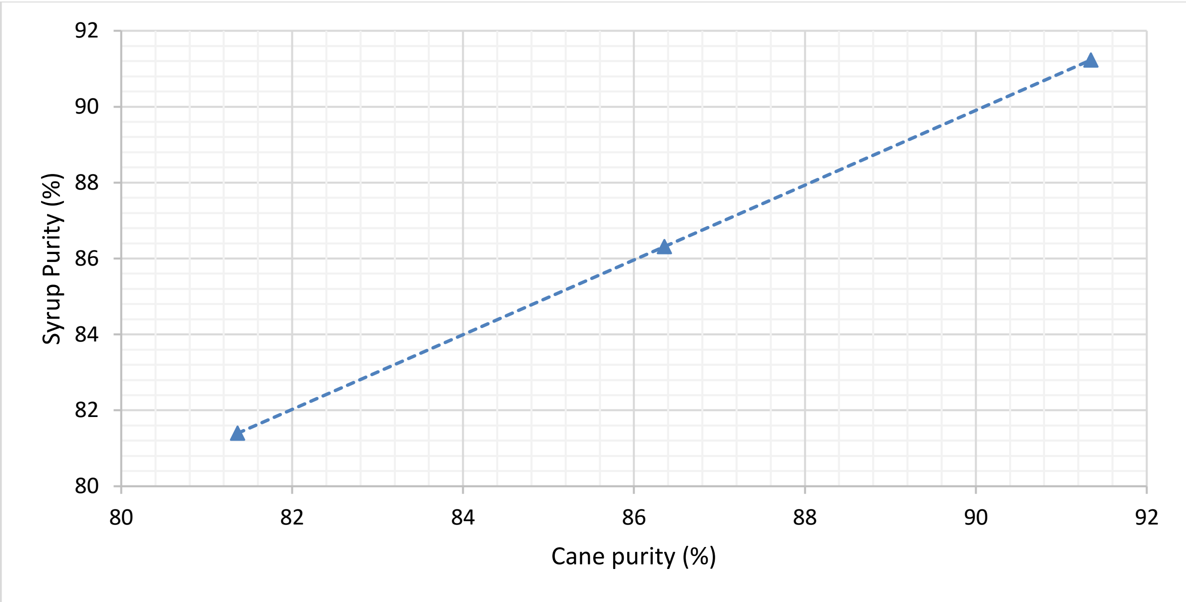

Figure 5.3 Model predictions of the effect of cane purity on syrup purity

From figure 5.3 it may be seen that there is no difference between the MATLAB™ results and the Aspen Plus® results.

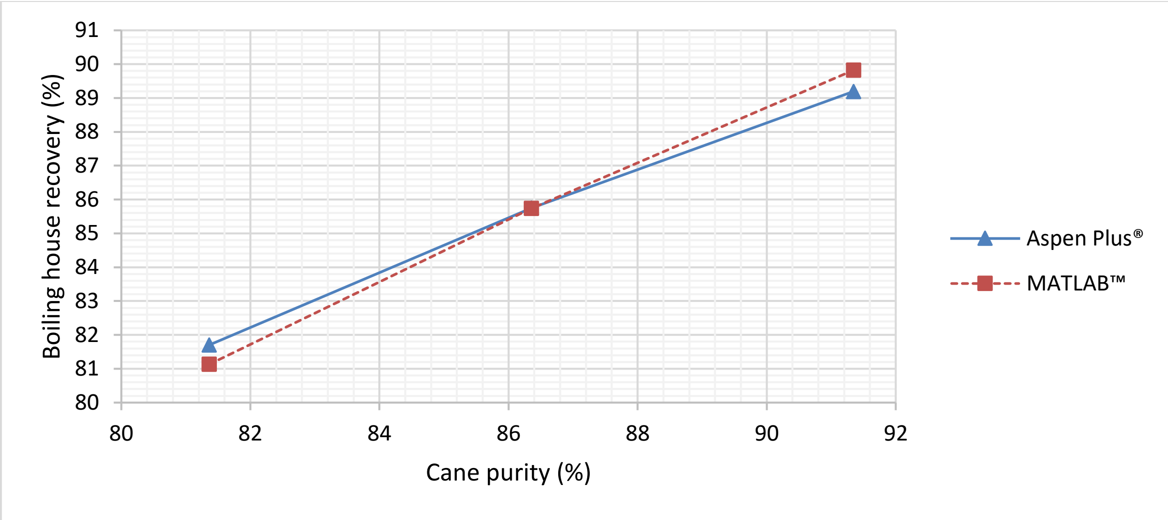

Figure 5.4 Model predictions of the effect of cane purity on boiling house recovery

From figure 5.4 it can be seen that the boiling house recovery for the Aspen Plus® model behaves differently to the MATLAB™ model, however the values are within 1 % of each other.

5.5.4 Effects of diverting an intermediate stream (clear juice)

Comparison between Aspen Plus® and MATLAB™ results for a cane throughput of 244.18 t/h and 10 % clear juice diverted to a biorefinery are shown in table 5.11.

Table 5.11 Results of diverting 10 % of an intermediate stream (clear juice)

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	Clear juice	Clear juice	Syrup	Syrup	Dry sugar	Dry sugar	Final molasses	Final molasses	Boiler bagasse	Boiler bagasse
Pressure (bara)	2.39	2.39	0.16	0.16	1.01	1.01	1.01	1.01	1.01	1.01
Temperature (°C)	99.8	99.9	58.7	58.6	35.9	35.9	56.1	56.1	64.5	64.3
Flow rate (tonnes/h)	253.49	253.49	52.40	52.36	25.99	25.97	9.80	9.97	40.56	40.03
	|		|		|		|		|	
Water (%)	86.32	86.32	35.18	35.17	0.08	0.08	20.91	22.12	50.96	50.96
Sucrose (%)	11.83	11.83	55.95	55.95	0.27	0.27	32.45	32.04	1.49	1.49
Non-sucrose (%)	1.84	1.84	8.87	8.87	0.48	0.48	46.63	45.85	0.16	0.16
Fibre (%)	0	0	0	0	0	0	0	0	47.39	47.39
Lime (%)	0	0	0	0	0	0	0	0	0	0
Crystal (%)	0	0	0	0	99.17	99.17	0	0	0	0

Comparison between Aspen Plus® and MATLAB™ results for a cane throughput of 244.18 t/h and 20 % clear juice diverted to biorefinery are shown in table 5.12.

Table 5.12 Results of diverting 20 % of an intermediate stream (clear juice)

Model:	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®	MAT-LAB™	Aspen Plus®
Stream:	Clear juice	Clear juice	Syrup	Syrup	Dry sugar	Dry sugar	Final molasses	Final molasses	Boiler bagasse	Boiler bagasse
Pressure (bara)	2.39	2.39	0.16	0.16	1.01	1.01	1.01	1.01	1.01	1.01
Temperature (°C)	99.8	99.9	58.7	58.6	35.9	35.9	56.1	56.1	64.5	64.4
Flow rate (tonnes/h)	225.31	225.32	46.6	46.56	23.11	22.15	8.71	9.68	37.48	36.96
	|		|		|		|		|	
Water (%)	86.33	86.32	35.18	35.20	0.08	0.08	20.91	18.86	50.96	50.96
Sucrose (%)	11.83	11.83	55.95	55.94	0.27	0.28	32.45	39.11	1.49	1.49
Non-sucrose (%)	1.84	1.84	8.87	8.87	0.48	0.50	46.63	42.04	0.16	0.16
Fibre (%)	0	0	0	0	0	0	0	0	47.39	47.39
Lime (%)	0	0	0	0	0	0	0	0	0	0
Crystal (%)	0	0	0	0	99.17	99.14	0	0	0	0

From table 5.12 it can be seen that up to the syrup stream the models agree. However, the molasses flow rates are quite different. The Aspen Plus® value is 11.1 % higher than the MATLAB™ model. Also, the dry sugar flow rate is 4.2 % lower in the Aspen Plus® model. However, the composition of the dry sugar is fairly consistent between the two models.

CHAPTER 6: CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusions

A 'generic' South African sugar mill was successfully modelled using Aspen Plus®. The model was initially verified against the MATLAB™ model for a cane throughput of 244.18 t/h. The results compared very well throughout the sugar mill and many improvements to the MATLAB™ model were made as a result of this project. More changes have been made to the MATLAB™ model since September 2016 but it was decided that the Aspen Plus® model should be based on this date so as to focus on the dissertation. One major error picked up since then was the fact that the cooling section of the dryer increases the moisture content of the sugar in the MATLAB™ model.

The Aspen Plus® model was made to be predictive and most of the results compared favourably with the MATLAB™ results. Higher and lower cane throughputs were simulated. The effect of increasing the imbibition flow rate was quantified. Also, the effects on the rest of the mill where analysed when portions of the clear juice were diverted away from the evaporators. And lastly, the effects of varying the cane purity on cane purity and boiling house recovery.

There are certain areas which were simplified, for example the centrifuges. Also, certain parameters were unable to be implemented in Aspen Plus®, for example the 'C' massecuite purity after the crystallisers. A design specification was tested however convergence failed. This is one reason why molasses flow rates and purities start diverging when cane throughputs are changed from 244 t/h.

The process of converting the Aspen Plus® model into an independent, predictive model required the mathematical structure of the model to be handled differently. This is due to differences in the way Degree-Of-Freedom (DOF) analysis is handled in the two software packages. While the MATLAB™ code was designed to satisfy the DOF around entire modules, Aspen Plus® requires the DOF to be satisfied around each unit operation. This meant that certain unit operations had to be solved differently to the MATLAB™ model.

The unit operation which was most difficult to solve was the vapour bleed splitter of the third effect evaporator. Vapour is distributed to the primary mixed juice heater and the fourth effect evaporator. The flow rate of vapour to the primary mixed juice heater is determined by the specified temperature of the exit mixed juice from the heater. Also, the flow rate of vapour to the fourth effect is manipulated by the specified final syrup water concentration. This creates an over-specified situation in Aspen Plus® since only one of the two outputs may be specified. This was solved by using a dynamic tool (Proportional-Integral controller). Microsoft Excel® iterates through flow rates for the exhaust steam to the first effect evaporator in order to produce exactly the right vapour in the third effect evaporator to meet both output specifications.

A steady-state raw sugar mill model is a complex undertaking with numerous recycle streams. To find a solution which is robust and reliable was challenging. Initial guesses for recycle streams can help Aspen Plus® to converge. The Aspen Plus® model has over 200 streams which are completely specified in terms of temperature, pressure and composition.

6.2 Recommendations

With the knowledge which has been gained over the last 3 years there are many recommendations on improving the model.

Further work has been discovered regarding simulations in Aspen Plus®. Equation oriented modelling would be a worthwhile method to investigate (Mansouri, 2015).

The enthalpy values were calculated from individual component heat capacities in Aspen Plus® whereas the MATLAB™ model makes use of empirical correlations for sugar stream enthalpies. It is a recommendation to incorporate a FORTRAN subroutine for the enthalpy calculations in Aspen Plus® which has been accomplished by Palacios-Bereche et. al. (2013).

Two major technical upgrades to the Aspen Plus® software have recently been accomplished: Custom models integrate directly into Aspen Plus® now. This means that creating a custom model of difficult units like batch pans and centrifuges would be easier now. The second upgrade is 64-bit support for Aspen Simulation Workbook®. It is hard to keep track of all the process parameters in Aspen Plus® but using the Aspen Simulation Workbook® in Microsoft Excel® this would be much simpler. Custom icons were attempted to be made in Aspen Plus® V8.8 but the software crashed repeatedly.

Various small changes to the Aspen Plus® model are recommended:

Extraction module:

- Incorporate heat losses, draft juice and megasse temperatures as functions in the diffuser separator block/model.
- Create custom icons for the diffuser and dewatering mills.

Evaporation module:

- Change the way evaporators are modelled to a combination of heat exchangers and flash vessels.
- Relax the condition of controlling mixed juice temperature from the primary mixed juice heater (this would mean that the 3rd effect vapour bleed splitter would not be over-specified). The flow rate of exhaust steam to the 1st effect evaporator could then be solved using a PI controller based on brix control of the syrup.
- The barometric condenser should be modelled as a mixer, with a calculator block governing the flow of cooling water in order to achieve a desired temperature of the outlet mixture.

Crystallisation module:

- Create custom models for the pan boiling processes.
- Calculate initial values for the magma and remelt recycle streams based on a regression of MATLAB™ results and input it to Aspen Plus® before simulations are run.
- Convert design specifications for brix control of exit massecuites and magma into calculator blocks in order to improve convergence.

Cooling tower module:

- Proper cooling water control, make-up and/ effluent calculator blocks.
- The MATLAB™ model specifies that the water from the cooling tower is at a temperature of 25 ^{o}C. It is recommended that this temperature be increased to a more realistic value of 35 ^{o}C.

Drying module:

- Ensure drying in heating and cooling sections of the dryer.
- Create a custom model of the dryer in order to reduce the number of units represented in the drying module.

6.3 Future work

6.3.1 High pressure boiler modelling for cogeneration

Sugar cane fibre naturally retains its same mass in moisture. Thus, no matter how much you press it, it will still retain about 45 % moisture (Davis, 2018). Moisture in the boiler fuel has to be heated to the combustion temperature and evaporated using up both sensible and latent heat. However, in the past, bagasse was seen as a 'waste' product hence burning it to produce steam was the best option. Boilers were not designed to be efficient since there was generally an excess of bagasse.

High pressure boilers (70 or 105 bara) would give much greater efficiencies, and hence more steam or electricity could be produced for the same amount of bagasse. Boilers represent a huge capital outlay for sugar mills though and thus with the 'relatively' cheap price of coal it is cheaper for sugar mills with biorefineries (e.g. Sezela mill) to burn coal to supplement the bagasse (Kruger, 2016). Sezela mill sends all their bagasse to reactors (steam explosion) where many high value products are formed with the main being furfural. The residue from the process is then sent to their boilers to produce steam. Future work on the Aspen Plus® model would be to consider high pressure boilers and an isolated boiler feed water loop. An isolated loop protects the boiler from impurities which can damage the boiler at high pressures.

6.3.2 Ethanol plant model

An ethanol biorefinery has been successfully modelled in MATLAB™ and it would be interesting to model this in Aspen Plus® as well.

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APPENDIX A: ASPEN PLUS® MODEL UNITS

This appendix explains the units which were used to build the Aspen Plus® model of a sugarcane biorefinery.

The following information is provided:

- The unit operation/process which has been modelled.
- The name of the unit/block as it appears in Aspen Plus®.
- A brief description of the unit operation/process.
- How the unit operation/process was modelled in Aspen Plus® in order to adequately describe the mass and energy balance equations.
- Any assumptions used in the formulation of the unit/block.

A.1 Extraction module

A.1.1 Cane knives

Aspen Plus® block: C-KNIVES (Dupl)

Description: Cane knives cut the whole-stick cane into pieces. The composition does not change.

Modelled by: Stream duplicator block (Dupl can be found in the model palette of Aspen Plus®, in the Manipulators tab).

Assumption: Steam used by the motor drives of the cane knives is proportional to the feed of cane. The knives use 0.0207 kg steam/kg cane.

A.1.2 Cane shredder

Aspen Plus® block: SHREDDER (Dupl)

Description: The cane shredder exposes the juice-bearing cells by smashing the chopped cane. The composition does not change.

Modelled by: stream duplicator.

Assumptions: The motor drives of the shredders use 0.0621 kg steam/kg cane.

A.1.3 Diffuser

Aspen Plus® block: DIFFUSER (Sep)

Description: Juice extraction occurs in a diffuser or milling tandem. In this work a diffuser is modelled.

Modelled by: Component separator. Separates components based on specified split fractions. Split fractions to the draft juice are: 0.4690; 0.8089; 0.8714; 0.02945 for water, sucrose, non-sucrose and fibre, respectively. The split fraction of water is different to the MATLAB™ model due to the scalding juice recycle stream being taken into account in Aspen Plus®. The scalding juice recycle stream is absent in the MATLAB™ model.

Assumptions: The diffuser can be sufficiently modelled as a perfectly mixed tank. In order to account for the difference in temperature (in reality) between the megasse and the draft juice stream, heat is transferred from the draft juice to the megasse by means of two fictitious units, namely DJCOOL and MEGHEAT.

A.1.4 Diffuser heat exchanger (scalding juice heater)

Aspen Plus® block: DIFFHX (HeatX)

Description: Heats the scalding juice and recycles it into the diffuser.

Modelled by: Heat exchanger. Vapour bleed steam (V2) was used as the heating medium.

Assumptions: Hot stream outlet temperature was specified as 106 °C. This ensures that the steam condenses.

A.1.5 Heat losses in the diffuser

Aspen Plus® block: HLOSS (Heater)

Description: This block models the heat lost to the environment by the diffuser.

Modelled by: Cooler with a specified temperature drop.

Assumptions: The heat losses in the diffuser are accounted for by dropping the draft juice temperature by 3.2 °C.

A.1.6 Temperature correction of draft juice and megasse

Aspen Plus® block: DJCOOL (Heater) and MEGHEAT (Heater)

Description: Draft juice and megasse leave at different temperatures from a real diffuser. Since the diffuser is modelled by a perfectly mixed tank the exit stream temperatures are identical. The temperatures are corrected by these two fictitious units.

Modelled by: Heater and cooler. Heat is transferred from the draft juice stream to the megasse stream. The draft juice exit temperature from DJCOOL is specified. The calculated heat duty is then transferred to the MEGHEAT block. A calculator block (MEG-DJ-Q) performs the heat transfer.
Assumptions: The draft juice temperature is 59.96 °C.

A.1.7 Dewatering mills

Aspen Plus® block: MILLS (Sep)
Description: Megasse is dried in a series of dewatering mills to increase the calorific value.
Modelled by: Component separator. Separates components based on specified split fractions. Split fractions to the bagasse stream are: 0.1421; 0.1512; 1 for sucrose, non-sucrose and fibre, respectively. A calculator block determines the split fraction of water due to a specification that there is 50.1 % water in the bagasse.
Assumptions: The motor drives of the mills use 0.0639 kg steam/kg fibre in megasse.

A.1.8 Heat losses in the dewatering mills

Aspen Plus® block: HLOSS1 (Heater)
Description: This block models heat losses in the dewatering mills.
Modelled by: Cooler with a specified temperature drop.
Assumptions: The press water loses 1 °C to the environment during the dewatering process.

A.1.9 Press water pump

Aspen Plus® block: PW-PUMP (Pump)
Description: The press water is pumped from the dewatering mills to a tank.
Modelled by: Pump.
Assumptions: The discharge pressure was assumed to be 2.068 bara.

A.1.10 Press water tank

Aspen Plus® block: PWTANK (Mixer)
Description: Temporary hold up tank for press water before recycling to diffuser.
Modelled by: Stream mixer. Direct steam injection maintains a specified temperature.
Assumptions: No heat loss. Maintained at a pressure of 2.068 bara.

A.1.11 High pressure steam splitter

Aspen Plus® block: S-SPLIT1 (FSplit)

Description: Splits high pressure steam (31 bara) to the cane knives, shredders and dewatering mills turbines.

Modelled by: Stream splitter. Stream split flow rates are set by calculator blocks.

Assumptions: No heat loss.

A.1.12 Motor drives

Aspen Plus® block: ED1, ED2 and ED3 (Compr)

Description: These motor drives turn the cane knives, cane shredder and dewatering mills.

Modelled by: Isentropic turbine.

Assumptions: The discharge pressure is set to 2 bara. The isentropic efficiency is set at 0.856 in order to match the temperature of the MATLAB™ exhaust streams. The valid phases were set at Vapor-Only or else an error occurred (The error was dew point temperature reached at some intermediate condition).

A.1.13 Low pressure steam collector (exhaust from motor drives)

Aspen Plus® block: C-MIX-DR (Mixer)

Description: Collects the exhaust steam from the extraction module turbines.

Modelled by: Stream mixer. The three low pressure (2 bara) steam streams (from the cane knives, cane shredder and dewatering mills) are mixed together and then sent to the evaporation module.

Assumptions: No heat loss. No pressure change.

A.2 Clarification module

A.2.1 Mixed juice tank

Aspen Plus® block: MJTANK (Mixer)

Description: Collects three streams: Draft juice from the diffuser, recovered juice from the vacuum filter and a small amount of sludge from the syrup clarifier in the evaporation module.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure change.

A.2.2 Primary mixed juice heater

Aspen Plus® block: MJHX1 (HeatX)

Description: Raises the temperature of mixed juice to 77.2 °C. Steam bled from the third effect evaporator (V3) provides the heat.

Modelled by: Heat exchanger. Hot stream outlet temperature was specified as 85.9 °C in order to condense all the steam. Model fidelity was set as shortcut. Flow direction was set to countercurrent.

Assumptions: No heat loss. No pressure change.

A.2.3 Limer

Aspen Plus® block: LIMER (Mixer)

Description: Adds milk of lime (stream LIM) to the mixed juice.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure change.

A.2.4 Mixed juice pump

Aspen Plus® block: MJPUMP (Pump)

Description: Increases the pressure of the mixed juice to 3.5 bara.

Modelled by: Pump with discharge pressure set to 3.5 bara.

Assumptions: No heat loss. Pump efficiency is 0.65.

A.2.5 Secondary mixed juice heater

Aspen Plus® block: MJHX2 (HeatX)

Description: Raises the temperature of mixed juice to 93.9 °C. Steam bled from the second effect evaporator (V2) provides the heat.

Modelled by: Heat exchanger. Hot stream outlet temperature was specified as 106 °C in order to condense all the steam. Model fidelity was set as shortcut. Flow direction was set to countercurrent.

Assumptions: No heat loss. No pressure change.

A.2.6 Tertiary mixed juice heater

Aspen Plus® block: MJHX3 (HeatX)

Description: Raises the temperature of mixed juice to 103.9 °C. Steam bled from the first effect evaporator provides the heat.

Modelled by: Heat exchanger. Hot stream outlet temperature was specified as 113.3 °C in order to condense all the steam. Model fidelity was set as shortcut. Flow direction was set to countercurrent.

Assumptions: No heat loss. No pressure change.

A.2.7 Mixed juice flash

Aspen Plus® block: FLASH1 (Flash2)

Description: Flashes the mixed juice in order to remove suspended air particles. The vapour stream is vented to atmosphere.

Modelled by: Flash vessel. Pressure set at atmospheric (1.013 bara).

Assumptions: Temperature set at 100.309°C in order to match the MATLAB™ model vent stream flow rate.

A.2.8 Clarifier

Aspen Plus® block: CLARIFY (Sep)

Description: Removes impurities from the mixed juice. Impurities flocculate together and sink to the bottom of the vessel.

Modelled by: Component separator. Split coefficients of sucrose, non-sucrose, fibre and lime are 0.1154, 0.1573, 1 and 1, respectively. The split coefficient of water is set by calculator block W-MUD in order to maintain a specified moisture content in the mud stream.

Assumptions: Mass fraction of water in mud is 0.804. Heat loss is taken into account.

A.2.9 Heat losses in clarifier

Aspen Plus® block: CLA-HL1 and CLA-HL2 (Heater)

Description: Heat is lost to the environment due to the residence time and elevated temperature in the clarifier.

Modelled by: Cooler with pressure and temperature change specified.

Assumptions: A temperature drop of 0.5 °C is assumed. No pressure drop.

A.2.10 Mud blender

Aspen Plus® block: MUD-BLEN (Mixer)

Description: Adds bagacillo (modelled as bagasse) to the clarifier mud stream.

Modelled by: Stream mixer.

Assumptions: The flow rate of bagasse to mud-bagasse blender is set at 0.02743 kg/kg mud.

A.2.11 Vacuum filter

Aspen Plus® block: VACFIL (Sep)

Description: Recovers juice from the mud and forms a filter cake under vacuum.

Modelled by: Component separator. Split coefficients of sucrose, non-sucrose, fibre and lime are 0.05147, 0.1789, 1 and 0.8267, respectively. Split coefficient of water is set by calculator block W-CAKE in order to maintain a specified moisture content in the filter cake.

Assumptions: Mass fraction of water in filter cake is 0.7000.

A.2.12 Filtrate juice splitter

Aspen Plus® block: FJ-SPLIT (FSplit)

Description: Provides the capacity to purge some filtrate juice from the system.

Modelled by: Stream splitter.

Assumptions: No purge currently. Redundant block which illustrates a potential feedstock stream to a bio-reactor.

A.2.13 Clear juice pump

Aspen Plus® block: CJPUMP (Pump)

Description: Pumps the clear juice to the evaporators.

Modelled by: Pump.

Assumptions: Discharge pressure is 2.392 bara. Pump efficiency is 0.65.

A.3 Evaporation module

A.3.1 Clarified juice preheater

Aspen Plus® block: JUICE-HX (HeatX)

Description: Raises the temperature of clarified juice to 112.5 °C. Exhaust steam from the turbo-alternator (boiler module) and motor drive turbines (extraction module) provide the heat.

Modelled by: Heat exchanger. Hot stream outlet temperature was specified as 120.21 °C in order to condense all the steam. Model fidelity was set as shortcut. Flow direction was set as countercurrent.

Assumptions: No heat loss. No pressure change.

A.3.2 Exhaust steam mixer

Aspen Plus® block: SMIX1 (Mixer)

Description: Joins exhaust steam from the turbo-alternator (boiler module stream SB2) and motor drive turbines (extraction module stream SD7).

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.3 Exhaust steam splitter

Aspen Plus® block: SSPLIT1 (FSplit)

Description: Distributes the exhaust steam to the clarified juice preheater and the 1st effect evaporator.

Modelled by: Stream splitter. Steam flow rate to the clarified juice preheater is manipulated by calculator block EVAPS.

Assumptions: No heat loss. No pressure change.

A.3.4 Steam condensation in 1st effect

Aspen Plus® block: 1ST-CON (Heater)

Description: Exhaust steam condenses in the calandria of the 1st effect evaporator. In Aspen Plus® this is modelled by a separate condenser. Latent heat of vapourisation is sent to the evaporation process (modelled by flash tank). Calculator block HL1 accomplishes the heat transfer after applying heat losses.

Modelled by: Cooler with temperature specified as 120.212 °C (just below the boiling point) to ensure all the steam condenses. The pressure was also specified as 2 bara.

Assumptions: No pressure drop. Heat loss of 0.58 %.

A.3.5 Evaporation process in 1st effect

Aspen Plus® block: 1ST-EFF (Flash2)

Description: Models the evaporation of water from the clarified juice.

Modelled by: Flash vessel. Pressure was specified as 1.6 bara.

Assumptions: Liquid entrainment in vapour outlet is 0.7745 %.

A.3.6 Vapour bleed (1st effect)

Aspen Plus® block: V-DIST1 (FSplit)

Description: Distributes the evaporated vapour in the 1st effect to the following: diffuser, 'A' pan, 'B' pan, press water tank, tertiary mixed juice heater and 2nd effect evaporator.

Modelled by: Stream splitter. All flow rates are specified besides the vapour to the 2nd effect evaporator. The specified flow rates are manipulated by calculator blocks.

Assumptions: No heat loss. No pressure change.

A.3.7 Hydraulic pressure drop - 1st effect

Aspen Plus® block: PDROP1 (Valve)

Description: In order to account for hydraulic temperature losses in the steam lines a pressure drop of 0.02 bar was assumed.

Modelled by: Valve with specified pressure drop.

Assumptions: No heat loss. Pressure drop of 0.02 bar.

A.3.8 Condensate mixer (preheater and 1st effect condensates)

Aspen Plus® block: C-MIX1 (Mixer)

Description: Joins the condensates from the clarified juice preheater and 1st effect condenser. The outlet condensate is then sent to the boiler module for use as boiler feed water.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.9 Steam condensation in 2nd effect

Aspen Plus® block: 2ND-CON (Heater)

Description: Vapour formed in the 1st effect evaporator is used to provide the heating duty of the 2nd effect. Latent heat of vapourisation is sent to the evaporation process (modelled by a flash tank). Calculator block HL2 accomplishes the heat transfer after applying heat losses.

Modelled by: Cooler with outlet temperature specified as 112.9 °C (just below the boiling point) to ensure all the steam condenses. The pressure was also specified as 1.58 bara.

Assumptions: No pressure drop. Heat loss of 0.6 %.

A.3.10 Condensate mixer (2nd effect)

Aspen Plus® block: CON-MIX2 (Mixer)

Description: Joins the condensates from: the 2nd effect evaporator, tertiary mixed juice heater, 'A' pans and 'B' pans. The outlet condensate is then sent to a flash vessel in order to recover some more process steam by lowering the pressure.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.11 Condensate flash (2nd effect)

Aspen Plus® block: FLASH2 (Flash2)

Description: Flashes the condensate from the 2nd effect condensate mixer.

Modelled by: Flash vessel. Pressure was specified as 1.25 bara.

Assumptions: Duty set to -200 MJ/h in order to match the MATLAB™ results.

A.3.12 Evaporation process in 2nd effect

Aspen Plus® block: 2ND-EFF (Flash2)

Description: Models the evaporation of water from the liquid outlet of the 1st effect. (i.e. models the evaporation of water in the 2nd effect evaporator)

Modelled by: Flash vessel. Pressure was specified as 1.25 bara.

Assumptions: Liquid entrainment in vapour outlet is 0.3927 %.

A.3.13 Vapour mixer (2nd effect)

Aspen Plus® block: V-MIX2 (Mixer)

Description: Mixes the vapour formed in the 2nd effect evaporator and vapour produced in the 2nd effect condensate flash.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.14 Vapour bleed (2nd effect)

Aspen Plus® block: V-DIST2 (FSplit)

Description: Distributes the vapour from the 2nd effect vapour mixer to the following: secondary mixed juice heater, scalding juice heater, remelter, 'C' pan and 3rd effect evaporator.

Modelled by: Stream splitter. All flow rates set besides the vapour to the 3rd effect evaporator.

Assumptions: No heat loss. No pressure change.

A.3.15 Hydraulic pressure drop - 2nd effect

Aspen Plus® block: PDROP2 (Valve)

Description: In order to account for hydraulic temperature losses in the steam lines a pressure drop of 0.02 bar was assumed. Also, the steam is throttled to the 3rd effect evaporator in order to control the final effect syrup brix. This valve is assumed to have a 0.15 bar pressure drop.

Modelled by: Valve with specified pressure drop.

Assumptions: No heat loss. Pressure drop of 0.17 bar.

A.3.16 Steam condensation in 3rd effect

Aspen Plus® block: 3RD-CON (Heater)

Description: The condensation of vapour in the calandria of the 3rd effect evaporator is modelled by a separate condenser. Latent heat of vapourisation is sent to the evaporation process (modelled by flash tank). Calculator block HL3 accomplishes the heat transfer after applying heat losses.

Modelled by: Cooler with temperature specified as 101.77 °C (just below the boiling point) to ensure all the steam condenses. The pressure was also specified as 1.08 bara.

Assumptions: No pressure drop. Heat loss of 0.1 %.

A.3.17 Condensate mixer (3rd effect)

Aspen Plus® block: CON-MIX3 (Mixer)

Description: Joins the condensates from: the 3rd effect evaporator, scalding juice heater, 'C' pan, secondary mixed juice heater and the liquid stream from the 2nd effect condensate flash. The condensate outlet is then sent to a flash vessel in order to form some more process steam.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.18 Condensate flash (3rd effect)

Aspen Plus® block: FLASH3 (Flash2)

Description: Flashes the condensate from the 3rd effect condensate mixer.

Modelled by: Flash vessel. Pressure was specified as 0.6 bara.

Assumptions: Duty set to -719 MJ/h in order to match the MATLAB™ results.

A.3.19 Evaporation process in 3rd effect

Aspen Plus® block: 3RD-EFF (Flash2)

Description: Models the evaporation of water from the liquid outlet of the 2nd effect.

Modelled by: Flash vessel. Pressure was specified as 0.6 bara.

Assumptions: Liquid entrainment in vapour outlet is 0.2353 %.

A.3.20 Vapour mixer (3rd effect)

Aspen Plus® block: V-MIX3 (Mixer)

Description: Mixes the vapour formed in the 3rd effect evaporator and vapour produced in the 3rd effect condensate flash.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.21 Vapour bleed (3rd effect)

Aspen Plus® block: V-DIST3 (FSplit)

Description: Distributes the vapour outlet from the 3rd effect vapour mixer to the following: Primary mixed juice heater and 4th effect evaporator.

Modelled by: Stream splitter. The flow rate to the 4th effect evaporator is specified and manipulated by a design specification (described in appendix C.1).

Assumptions: No heat loss. No pressure change.

A.3.22 Hydraulic pressure drop (3rd effect)

Aspen Plus® block: PDROP3 (Valve)

Description: In order to account for hydraulic temperature losses in the steam lines a pressure drop of 0.02 bar was assumed.

Modelled by: Valve with specified pressure drop.

Assumptions: No heat loss. Pressure drop of 0.02 bar.

A.3.23 Inversion

Aspen Plus® block: INV-1, INV-2, INV-3, INV-4 and INV-5 (RStoic)

Description: Sucrose inversion has been modelled in all five effects.

Modelled by: Stoichiometric reactor. Reaction is: Sucrose + Water → 1.76625 Non-sucrose.

Assumptions: No heat of reaction is taken into account.

A.3.24 Entrainment separation

Aspen Plus® block: ENTRAIN1, ENTRAIN2, ENTRAIN3, ENTRAIN4 and ENTRAIN5 (Flash2)

Description: The droplet entrainment in the vapour stream leaving the evaporators is purged from the system in order to match the MATLAB™ model.

Modelled by: Flash vessel operated at the identical pressure as the vapour streams. (I.e. functions as a vapour-liquid separator with no pressure drop)

Assumptions: Theoretical unit which can be removed at a later stage in order to make a more realistic model.

A.3.25 Steam condensation in 4th effect

Aspen Plus® block: 4TH-CON (Heater)

Description: The condensation of vapour in the calandria of the 4th effect evaporator is modelled by a separate condenser. Latent heat of vapourisation is sent to the evaporation process (modelled by a flash tank). Calculator block HL4 accomplishes the heat transfer after applying heat losses.

Modelled by: Cooler with temperature specified as 85.06 °C (just below the boiling point) to ensure all the steam condenses. The pressure was also specified as 0.58 bara.

Assumptions: No pressure drop. Heat loss of 0.385 %.

A.3.26 Condensate mixer (4th effect)

Aspen Plus® block: CON-MIX4 (Mixer)

Description: Joins the condensates from: the 4th effect evaporator, primary mixed juice heater and the liquid stream from the 3rd effect condensate flash. The condensate is then sent to a flash vessel in order to form some more process steam.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.27 Condensate flash (4th effect)

Aspen Plus® block: FLASH4 (Flash2)

Description: Flashes the condensate from the 4th effect condensate mixer.

Modelled by: Flash vessel. Pressure was specified as 0.4 bara.

Assumptions: Duty set to -317 MJ/h in order to match the MATLAB™ results.

A.3.28 Evaporation process in 4th effect

Aspen Plus® block: 4TH-EFF (Flash2)

Description: Models the evaporation of water from the liquid outlet of the 3rd effect.

Modelled by: Flash vessel. Pressure was specified as 0.4 bara.

Assumptions: Liquid entrainment in vapour outlet is 0.2213 %.

A.3.29 Vapour mixer (4th effect)

Aspen Plus® block: V-MIX4 (Mixer)

Description: Mixes the vapour formed in the 4th effect evaporator and vapour produced in the 4th effect condensate flash.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure change.

A.3.30 Hydraulic pressure drop (4th effect)

Aspen Plus® block: PDROP4 (Valve)

Description: In order to account for hydraulic temperature losses in the steam lines a pressure drop of 0.02 bar was assumed.

Modelled by: Valve with specified pressure drop.

Assumptions: No heat loss. Pressure drop of 0.02 bar.

A.3.31 Steam condensation in 5th effect

Aspen Plus® block: 5TH-CON (Heater)

Description: The condensation of vapour in the calandria of the 5th effect evaporator is modelled by a separate condenser. Latent heat of vapourisation is sent to the evaporation process (modelled by a flash vessel). Calculator block HL5 accomplishes the heat transfer after applying heat losses.

Modelled by: Cooler with temperature specified as 74.63 °C (just below the boiling point) to ensure all the steam condenses. The pressure was also specified as 0.38 bara.

Assumptions: No pressure drop. Heat loss of 0.2254 %.

A.3.32 Condensate splitter (5th effect)

Aspen Plus® block: CSPLIT (FSplit)

Description: Condensate from the 5th effect evaporator is split to the following: 5th effect condensate mixer, 'remelter' and centrifuges.

Modelled by: Stream splitter. All flow rates specified besides the flow rate to the 5th effect condensate mixer.

Assumptions: No heat loss. No pressure change.

A.3.33 Condensate mixer (5th effect)

Aspen Plus® block: CON-MIX5 (Mixer)

Description: Joins the condensates from: the 5th effect evaporator and the liquid stream from the 4th effect condensate flash. The condensate is then sent to a flash vessel in order to form some more process steam.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.34 Condensate flash (5th effect)

Aspen Plus® block: FLASH5 (Flash2)

Description: Flashes the condensate from the 5th effect condensate mixer.

Modelled by: Flash vessel. Pressure set at 0.16 bara.

Assumptions: Duty set to -447 MJ/h in order to match the MATLAB™ results.

A.3.35 Evaporation process in 5th effect

Aspen Plus® block: 5TH-EFF (Flash2)

Description: Models the evaporation of water from the liquid outlet of the 4th effect.

Modelled by: Flash vessel. Pressure set at 0.16 bara.

Assumptions: Liquid entrainment in vapour outlet is 0.2627 %.

A.3.36 Vapour mixer (5th effect)

Aspen Plus® block: V-MIX5 (Mixer)

Description: Mixes the vapour formed in the 5th effect evaporator and vapour produced in the 5th effect condensate flash.

Modelled by: Stream mixer. All settings left as defaults.

Assumptions: No heat loss. No pressure change.

A.3.37 Barometric condenser

Aspen Plus® block: BARCON (Heater)

Description: The vapour from the 5th (final) effect evaporator is condensed in order to produce a vacuum.

Modelled by: Cooler with temperature specified as 40 °C. Atmospheric pressure was specified (1.013 bara) since the condensate leaving the barometric leg is open to atmosphere.

Assumptions: No pressure drop. No heat loss.

A.3.38 Condensate pump

Aspen Plus® block: C-PUMP (Pump)

Description: The liquid condensate from the 5th effect condensate flash vessel is pumped at 3.081 bara to various parts of the plant.

Modelled by: Pump.

Assumptions: No heat loss.

A.3.39 Condensate splitter

Aspen Plus® block: C-SPLIT5 (FSplit)

Description: The liquid condensate from the 5th effect condensate flash vessel is distributed to the following: imbibition water, wash water to the vacuum filter and the remainder is an effluent stream from the plant (stream WEC).

Modelled by: Stream splitter. All flow rates were specified besides the effluent stream.

Assumptions: No heat loss. No pressure change.

A.3.40 Syrup (liquid from fifth effect evaporator) pump

Aspen Plus® block: L5PUMP (Pump)

Description: The liquid from the 5th effect is pumped at 1.013 bara to the syrup clarifier.

Modelled by: Pump.

Assumptions: No heat loss.

A.3.41 Syrup clarifier

Aspen Plus® block: SYR-FIL (Sep)

Description: The final effect liquid stream is called syrup. This syrup is clarified in order to remove impurities. A small sludge stream is recycled to the mixed juice tank while the majority of the syrup is sent to the crystallisation module.

Modelled by: Component separator. Split coefficients to the syrup stream are: 0.997; 0.998 and 0.996 for water, sucrose and non-sucrose, respectively.

Assumptions: No heat loss. No pressure drop.

A.4 Crystallisation module

A.4.1 Syrup distributor

Aspen Plus® block: SYRSPLIT (FSplit)

Description: Distributes syrup to the magma mingler and to the 'A' pans.

Modelled by: Stream splitter. The flow rate to the magma mingler is manipulated by a design specification (described in appendix C.5).

Assumptions: No heat loss. No pressure change.

A.4.2 Feed mixer to 'A' pans

Aspen Plus® block: A-MIX (Mixer)

Description: Joins three streams: Syrup from the syrup splitter, magma from the mingler and a 'remelt' stream. The outlet from the mixer is then sent to the 'A' pans.

Modelled by: Stream mixer. Pressure was specified as 1.013 bara.

Assumptions: No heat loss. Operates at atmospheric pressure.

A.4.3 Sucrose inversion in pans

Aspen Plus® block: INV-A, INV-B and INV-C (RStoich)

Description: These reactors model the inversion of sucrose in the three boilings.

Modelled by: Stoichiometric reactor. Reaction is: Sucrose + Water → 1.76625 Non-sucrose. The stoichiometric coefficient of non-sucrose comes from the mass balance since molecular weight of non-sucrose has been specified as being 204 g/mol.

Assumptions: No heat of reaction is taken into account.

A.4.4 Entrainment

Aspen Plus® block: ENT-A, ENT-B and ENT-C (Sep)

Description: These separators model droplet entrainment in the pans.

Modelled by: Component separator. Total input sucrose was calculated (sucrose + crystals fed to pans).

Assumptions: No heat loss.

A.4.5 Crystallisation in 'A' pans

Aspen Plus® block: A-PAN-CX (RStoich)

Description: This stoichiometric reactor models the crystallisation of sucrose to crystal sugar. The conversion of sucrose is governed by the SLE model for the ternary system of water, sucrose and non-sucrose. The conversion is dependent on a specified temperature (final temperature of 'A' pans), the supersaturation of the exit massecuite and a specified dry solids content.

Modelled by: Stoichiometric reactor. Temperature was specified as 60.4 °C and pressure was specified as 1.013 bara.

Assumptions: No heat of dissolution was taken into account.

A.4.6 Evaporation in 'A' pans

Aspen Plus® block: A-PANS (Flash2)

Description: Models the evaporation of water in the 'A' pans.

Modelled by: Flash vessel. Temperature was specified as 67.12 °C. The heat duty required for evaporation comes from the steam condensation process (A-CON block).

Assumptions: Heat loss was modelled by a calculator block (HL-A block) which applies a reduction factor to the energy liberated by the steam condensation process.

A.4.7 Steam condensation in 'A' pans

Aspen Plus® block: A-CON (Heater)

Description: Vapour bled from the 1st effect evaporator (V1) condenses in the calandria of the 'A' pans. In Aspen Plus® this is modelled by a separate condenser. Latent heat of vapourisation is sent to the evaporation process (modelled by a flash tank).

Modelled by: Condenser with temperature specified as 113.3 °C (just below the boiling point) to ensure all the steam condenses. The pressure was also specified as 1.6 bara.

Assumptions: No pressure drop. Heat loss of 4.48 %.

A.4.8 Cooling in 'A' crystallisers

Aspen Plus® block: COOLA (Heater)

Description: The crystallisers are cooled in order to increase the supersaturation and hence enable more sucrose to crystallise.

Modelled by: Cooler with hot side outlet temperature specified as 56.23 °C.

Assumptions: Atmospheric pressure. No heat loss.

A.4.9 Crystallisation in 'A' crystallisers

Aspen Plus® block: A-CX (RStoich)

Description: This stoichiometric reactor models the crystallisation of sucrose to crystal sugar in the 'A' station crystallisers. The conversion of sucrose is governed by the SLE equations.

Modelled by: Stoichiometric reactor. Temperature specified at 56.23 °C and pressure of 1.013 bara.

Assumptions: No heat of dissolution is taken into account.

A.4.10 Crystal losses in 'A' centrifuges

Aspen Plus® block: A-CX-LOS (RStoich)

Description: Due to the addition of wash water to the centrifuges, there is a change in supersaturation and some of the crystals dissolve.

Modelled by: Stoichiometric reactor.

Assumptions: The extent of crystal loss was calculated based on the MATLAB™ results. Temperature was specified as 56.23 °C (Same temperature as the massecuite leaving the 'A' crystallisers). Operated at atmospheric pressure.

A.4.11 Separation in 'A' centrifuges

Aspen Plus® block: A-FUGALS (Sep)

Description: Sugar crystals are separated from the surrounding liquid in screen basket centrifuges.

Modelled by: Component separator.

Assumptions: The split coefficients were calculated based on the MATLAB™ results. The exit temperature of molasses was specified as 60.2 °C. Temperature of the 'A' sugar outlet was specified as 56.2 °C. Pressure of outlets was specified as 1.013 bara.

A.4.12 Wash water distributor to centrifuges

Aspen Plus® block: WW-SPLIT (FSplit)

Description: Distributes wash water to the 'A', 'B' and 'C' centrifuges.

Modelled by: Stream splitter. The flow rate to the 'A' and 'B' centrifuges were specified.

Assumptions: No heat loss. No pressure change.

A.4.13 Crystallisation in 'B' pans

Aspen Plus® block: B-PAN-CX (RStoich)

Description: This stoichiometric reactor models the crystallisation of sucrose to crystal sugar. The conversion of sucrose is governed by the SLE model. The conversion is dependent on a specified temperature (final temperature of 'B' pans), the supersaturation of the exit massecuite and a specified dry solids content.

Modelled by: Stoichiometric reactor. Temperature was specified as 67 °C and pressure specified as 1.013 bara.

Assumptions: No heat of dissolution was taken into account.

A.4.14 Evaporation in 'B' pans

Aspen Plus® block: B-PANS (Flash2)

Description: Models the evaporation of water in the 'B' pans.

Modelled by: Flash vessel. Temperature was specified as 66.88 °C. The heat duty required for evaporation comes from the steam condensation process (B-CON block).

Assumptions: Heat loss was modelled by a calculator block (HL-B block) which applies a reduction factor to the energy liberated by the steam condensation process.

A.4.15 Steam condensation in 'B' pans

Aspen Plus® block: B-CON (Heater)

Description: Vapour bled from the 1st effect evaporator (V1) condenses in the calandria of the 'B' pans. In Aspen Plus® this is modelled by a separate condenser. Latent heat of vapourisation is sent to the evaporation process (modelled by a flash tank) in order to evaporate water from the 'A' molasses.

Modelled by: Condenser with temperature specified as 113.3 °C (just below the boiling point) to ensure all the steam condenses. The pressure was also specified as 1.6 bara.

Assumptions: No pressure drop. Heat loss of 5.1%.

A.4.16 Cooling in 'B' crystallisers

Aspen Plus® block: COOLB (Heater)

Description: The 'B' crystallisers are cooled with air in order to increase the supersaturation and hence allow more sucrose to crystallise.

Modelled by: Cooler with outlet temperature specified as 49.9 °C. The air stream was not shown.

Assumptions: Operated at atmospheric pressure. No heat loss.

A.4.17 Crystallisation in 'B' crystallisers

Aspen Plus® block: B-CX (RStoich)

Description: This stoichiometric reactor models the crystallisation of sucrose to crystal sugar. The conversion of sucrose is governed by the SLE model.

Modelled by: Stoichiometric reactor. Temperature and pressure were specified as 49.9 °C and 1.013 bara respectively.

Assumptions: No heat of dissolution is taken into account.

A.4.18 Crystal losses in 'B' centrifuges

Aspen Plus® block: B-CX-LOS (RStoich)

Description: Models the loss of crystals in the 'B' centrifuges due to wash water addition.

Modelled by: Stoichiometric reactor. Temperature and pressure specified.

Assumptions: The degree of conversion was calculated based on the MATLAB™ results. Temperature was specified as 49.9 °C (Same temperature as the massecuite leaving the 'B' crystallisers). No pressure change.

A.4.19 Separation in 'B' centrifuges

Aspen Plus® block: B-FUGALS (Sep)

Description: Sugar crystals are separated from the surrounding liquid in screen basket centrifuges.

Modelled by: Component separator.

Assumptions: The split coefficients were calculated based on the MATLAB™ results. The exit temperature of 'B' molasses was specified as 50.66 °C. The 'B' sugar exit temperature was specified as 49.9 °C. Pressure of outlet streams were specified as 1.013 bara.

A.4.20 Sugar splitter (distributes ‘B’ sugar)

Aspen Plus® block: BSUG-DIS (FSplit)

Description: Distributes ‘B’ sugar to the magma mingler and remelter.

Modelled by: Stream splitter. The split fraction to the mingler was specified.

Assumptions: No heat loss. No pressure change. The split fraction was calculated based on the MATLAB™ results. The split fraction to the mingler is 0.7677.

A.4.21 Magma mingler

Aspen Plus® block: MINGLER (Mixer)

Description: Mixes a portion of the ‘B’ sugar with syrup. This magma is then sent to the feed mixer for the ‘A’ pans.

Modelled by: Stream mixer.

Assumptions: No heat loss. Operated at atmospheric pressure.

A.4.22 Crystal losses in magma mingler

Aspen Plus® block: CX-MING (RStoich)

Description: This stoichiometric reactor models the sugar crystal loss in the magma mingler. Syrup (undersaturated) is added to the ‘B’ sugar in the mingler and this causes crystals to dissolve.

Modelled by: Stoichiometric reactor. Temperature (51.9 °C) and pressure (1.013 bara) specified. The conversion is manipulated by calculator block CX-MINGL which performs a SLE balance.

Assumptions: No heat of dissolution is taken into account.

A.4.23 Feed mixer to ‘C’ pans

Aspen Plus® block: C-FD-MIX (Mixer)

Description: Mixes a portion of the ‘A’ molasses with all of the ‘B’ molasses. This mixture is then sent to the ‘C’ pans.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure drop.

A.4.24 Crystallisation in ‘C’ pans

Aspen Plus® block: C-PAN-CX (RStoich)

Description: This stoichiometric reactor models the crystallisation of sucrose in the final boiling. The conversion of sucrose is governed by the SLE model. The conversion is dependent on a specified temperature (final temperature of ‘C’ pans), the supersaturation of the exit massecuite and a specified dry solids content.

Modelled by: Stoichiometric reactor. Temperature was specified as 60 °C and pressure specified as 1.013 bara.

Assumptions: No heat of dissolution is taken into account.

A.4.25 Condensation in ‘C’ pans

Aspen Plus® block: C-CON (Heater)

Description: Vapour bled from the 2nd effect evaporator (V2) condenses in the calandria of the ‘C’ pans. In Aspen Plus® this is modelled by a separate condenser. Latent heat of vapourisation is sent to the evaporation process (modelled by a flash tank) in order to evaporate water from the ‘B’ molasses (with some ‘A’ molasses added to increase the purity).

Modelled by: Condenser with temperature specified as 105.97 °C (just below the boiling point) to ensure all the steam condenses. The pressure is also specified as 1.25 bara.

Assumptions: No pressure drop. Heat loss of 8%.

A.4.26 Evaporation in ‘C’ pans

Aspen Plus® block: C-PANS (Flash2)

Description: Models the evaporation of water in the ‘C’ pans.

Modelled by: Flash vessel. Temperature was specified as 66.91 °C. The energy liberated by the steam condensation process (C-CON block) provides the heat duty for evaporation.

Assumptions: Heat losses were modelled by a calculator block (HL-C block).

A.4.27 Cooling in ‘C’ crystallisers

Aspen Plus® block: COOLC (Heater)

Description: The ‘C’ crystalliser is cooled with water in order to increase the supersaturation and hence allow more sucrose to crystallise.

Modelled by: Cooler with temperature approach of outlet specified as 2 °C. (Due to temperature crossover in MATLAB™ model).

Assumptions: No pressure drop. No heat loss.

A.4.28 Crystallisation in ‘C’ crystallisers

Aspen Plus® block: C-CX (RStoich)

Description: This stoichiometric reactor models the crystallisation of sucrose to crystal sugar in the ‘C’ crystallisers. The conversion of sucrose is governed by the SLE model.

Modelled by: Stoichiometric reactor. Temperature and pressure was specified as 45.22 °C and 1.013 bara respectively.

Assumptions: No heat of dissolution is taken into account.

A.4.29 Crystal losses in ‘C’ centrifuges

Aspen Plus® block: C-CX-LOS (RStoich)

Description: Due to the addition of wash water to the centrifuges, there is a change in supersaturation and some of the crystals dissolve.

Modelled by: Stoichiometric reactor. Temperature and pressure were specified.

Assumptions: The degree of conversion was calculated based on the MATLAB™ results. Temperature was specified as 45.22 °C (Same temperature as the massecuite leaving the ‘C’ crystallisers). Atmospheric pressure was assumed in all the centrifuges.

A.4.30 Separation in ‘C’ centrifuges

Aspen Plus® block: C-FUGALS (Sep)

Description: Sugar crystals are separated from the surrounding liquid in screen basket centrifuges.

Modelled by: Component separator.

Assumptions: The split coefficients were calculated based on the MATLAB™ results. The exit temperature of final molasses was specified as 56.1 °C. The ‘C’ sugar exit temperature was specified as 45.3 °C. Pressure of outlet streams were specified as 1.013 bara.

A.4.31 Barometric condenser for pans

Aspen Plus® block: V-COND (Mixer)

Description: Models the barometric condensers of all the batch pans.

Modelled by: Stream mixer. Pressure was specified as 3 bara to match MATLAB™ results.

Assumptions: No heat loss.

A.4.32 Remelter – crystal melting

Aspen Plus® block: CX-SUC (RStoich)

Description: All the ‘C’ sugar and some of the ‘B’ sugar crystals are dissolved in the remelter.

Modelled by: Stoichiometric reactor. Temperature (45 °C) and pressure (1.013 bara) was specified.

Assumptions: Total conversion of crystal sugar to sucrose.

A.4.33 Remelter – stream mixing

Aspen Plus® block: REM-MIX (Mixer)

Description: All the ‘C’ sugar and a portion of the ‘B’ sugar is mixed with some steam (V2) and condensate (from CSPLIT block in evaporation module).

Modelled by: Mixer.

Assumptions: No heat loss. No pressure drop.

A.4.34 Remelter – fictitious cooler

Aspen Plus® block: T-REMC (Heater)

Description: The ‘remelt’ comes out at a lower temperature than is predicted (83 °C) by Aspen Plus® due to a difference in the enthalpy calculation.

Modelled by: Heater.

Assumptions: The temperature of the ‘remelt’ was specified as 78.58 °C. The pressure was specified as 1.013 bara.

A.5 Drying module

A.5.1 Dry air distributor

Aspen Plus® block: AIR-SPLIT (FSplit)

Description: Distributes air to the dry air heater and sugar cooler.

Modelled by: Stream splitter. The split coefficient to the sugar cooler was specified.

Assumptions: No heat loss. No pressure drop.

A.5.2 Dry air heater

Aspen Plus® block: AIR-HEAT (HeatX)

Description: Air is heated before entering the sugar dryer.

Modelled by: Heat exchanger. Hot stream outlet temperature was specified as 120.21 °C in order to condense all the steam. Model fidelity was set as shortcut. Flow direction was set to countercurrent.

Assumptions: No heat loss. No pressure drop.

A.5.3 Dryer – heat exchanger

Aspen Plus® block: DRYER-HX (HeatX)

Description: Heat is transferred between the hot air and sugar in the cooler.

Modelled by: Heat exchanger. Hot stream outlet temperature was specified as 60 °C. Model fidelity was set as shortcut. Flow direction was set to countercurrent.

Assumptions: No heat loss. No pressure drop.

A.5.4 Dryer – moisture separation

Aspen Plus® block: DRYER (Sep)

Description: Water (moisture) is evaporated from the sugar in the dryer.

Modelled by: Component separator. The split fraction of water was determined from MATLAB™ results.

Assumptions: No heat loss. No pressure drop.

A.5.5 Dryer – moisture mixer

Aspen Plus® block: MOIST-MX (Sep)

Description: The moisture from the sugar in the dryer joins the air stream.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure drop.

A.5.6 Sugar cooler

Aspen Plus® block: AIR-HX (HeatX)

Description: The sugar is cooled by passing a stream of cold air over it.

Modelled by: Heat exchanger. The hot / cold outlet temperature approach was set to 1 °C.

Assumptions: No heat loss. No pressure drop.

A.5.7 Moisture splitter

Aspen Plus® block: MOIS2-SP (Sep)

Description: Moisture from the air is absorbed by the sugar in the cooling section of the dryer due to an error in the MATLAB™ model.

Modelled by: Component separator. The split fraction of water is manipulated by calculator block SUAD-MOI.

Assumptions: No heat loss. No pressure drop.

A.5.8 Moisture mixer 2

Aspen Plus® block: MOIS-MX2 (Sep)

Description: Moisture from the air is absorbed by the sugar in the cooling section of the dryer.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure drop.

A.5.9 Crystallisation in the dryer

Aspen Plus® block: CX-DRYER (RStoich)

Description: Most of the remaining sucrose crystallises in the dryer due to moisture being removed from the sugar.

Modelled by: Stoichiometric reactor. Conversion of sucrose was calculated from MATLAB™ results.

Assumptions: No heat of reaction was considered.

A.5.10 Dry air collector

Aspen Plus® block: AIR-MIX (Mixer)

Description: Collects the air streams from the dryer and sugar cooler.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure change.

A.6 Boiler module

A.6.1 Boiler feed water tank

Aspen Plus® block: W-MIX (Mixer)

Description: Collects three streams: Boiler feed water from the evaporation module, Condensate from the air heater in the dryer module and a make-up water stream. The outlet from the tank goes to the boiler.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure change.

A.6.2 Boiler

Aspen Plus® block: BOILERS (Heater)

Description: Bagasse is burnt in the boiler. High pressure steam is produced.

Modelled by: Heater with pressure and temperature specified.

Assumptions: Pressure was specified as 31 bara. Temperature of high pressure steam was set at 390 °C.

A.6.3 Bagasse distributor

Aspen Plus® block: BAG-SPLIT (FSplit)

Description: Not all of the bagasse from the dewatering mills is needed by the boiler. The excess bagasse could be used to make other high value products.

Modelled by: Stream splitter. The bagasse needed by the boiler is specified.

Assumptions: No heat loss. No pressure change. Specification: 0.45 kg bagasse is needed to produce 1 kg of steam.

A.6.4 Boiler water blowdown

Aspen Plus® block: WBB-SPLT (FSplit)

Description: A small amount of boiler feed water is removed in order to purge suspended solids.

Modelled by: Stream splitter. The flow rate of boiler water blowdown is manipulated by calculator block WBB-FLOW.

Assumptions: No heat loss. No pressure change.

A.6.5 High pressure steam distributor

Aspen Plus® block: SB1-SPLIT (FSplit)

Description: High pressure steam is distributed to the turbo-alternator and also to the motor drives in the extraction module. There is also a small amount of steam which is lost.

Modelled by: Stream splitter. The amount of steam which is required by the motor drives is specified as well as the amount of steam which is lost. Calculator block SBL-FLOW determines the amount of steam which is lost.

Assumptions: No heat loss. No pressure change.

A.6.6 Turbo-alternator

Aspen Plus® block: EB1 (FSplit)

Description: High pressure steam is used to generate electricity in a back pressure turbo-alternator.

Modelled by: Isentropic turbine.

Assumptions: The discharge pressure is set to 2 bara. The isentropic efficiency is set at 0.856 in order to match the temperature of the MATLAB™ exhaust streams.

A.6.7 Exhaust steam distributor

Aspen Plus® block: SB2-SPLT (FSplit)

Description: Exhaust steam is distributed to three places: the evaporation module (preheater and 1st effect evaporator), the deaerator and the dry air heater.

Modelled by: Stream splitter. The flow rates to the deaerator and the dry air heater were specified.

Assumptions: No heat loss. No pressure change.

A.7 Cooling tower module

A.7.1 Cooling water collector

Aspen Plus® block: CT-MIXIN (Mixer)

Description: Collects four streams: Cooling water from the ‘A’ and ‘C’ crystallisers and cooling water streams from the barometric condensers in the evaporation and crystallisation modules.

Modelled by: Stream mixer.

Assumptions: No heat loss. No pressure change.

A.7.2 Effluent splitter

Aspen Plus® block: EF-SPLIT (FSplit)

Description: Effluent is purged from the cooling water cycle.

Modelled by: Stream splitter. The cooling water to the cooling tower is specified.

Assumptions: No heat loss. No pressure change.

A.7.3 Cooling tower

Aspen Plus® block: C-TOWER (Flash2)

Description: Cools the cooling water by evaporating a portion of the feed.

Modelled by: Flash vessel. Temperature was specified as 27.3 °C. Duty was set at zero.

Assumptions: Pressure is calculated due to temperature and duty being specified.

A.7.4 Heat losses in cooling tower

Aspen Plus® block: CT-HLOSS (Heater)

Description: Heat is lost to the environment in the cooling tower.

Modelled by: Cooler with pressure and temperature specified.

Assumptions: Pressure is 3 bara. Temperature of cooling water is 25 °C.

A.7.5 Sucrose purge

Aspen Plus® block: SUC-SPLT (Sep)

Description: To ensure sucrose does not build up in the cooling water cycle, all sucrose is purged.

Modelled by: Component separator. Only sucrose is purged.

Assumptions: No heat loss. No pressure change.

A.7.6 Cooling water distributor

Aspen Plus® block: C-SPLIT (FSplit)

Description: Distributes the cooling water to the barometric condensers and crystallisers.

Modelled by: Stream splitter.

Assumptions: There is a slight excess, stream CT1, due to differences in CTV between MATLAB™ and Aspen Plus®.

APPENDIX B: ASPEN PLUS® CALCULATOR BLOCKS

B.1 Extraction module

KNIVESF Steam flow to knives turbine	MILLSF Steam flow to dewatering mills turbine	SB1-FLOW Steam flow to extraction module turbines	IW-F Imbibition flow	DIFF-WAT Diffuser extraction coefficient - water
SDH-FLOW Steam flow to scalding juice heater	SDI-FLOW Steam flow to diffuser	MEG-DJ-Q Heat transfer from draft juice to megasse	MILL-WAT Dewatering mills water separation coefficient	FSPW Steam flow to press water holdup tank

Figure B.1 Extraction module calculator blocks

B.1.1 Steam flow to cane knives motor turbine

Aspen Plus® block: KNIVESF

Description: Sets the flow rate of steam to the cane knives turbine. The flow rate is directly proportional to the feed rate of cane.

Variables:

1. FCANE – Input variable: Flow rate of cane (stream CANE) in kg/h.
2. FSTEAMK – Output variable: Split of block S-SPLIT1 in extraction flowsheet. This sets the flow rate of steam to the cane knives turbine (stream SD1).

Parameters:

1. KNIVES – Parameter 103: Steam needed by cane knives per kg of cane processed. Value is 0.0207.

Calculation: FSTEAMK = KNIVES × FCANE

B.1.2 Steam flow to dewatering mills motor turbine

Aspen Plus® block: MILLSF

Description: Sets the flow rate of steam to the dewatering mills turbine. This flow rate is directly proportional to the flow rate of fibre in the megasse stream.

Variables:

1. FMEGFIB – Input variable: Flow rate of fibre in megasse (stream MEG) in kg/h.
2. FSTEAMM – Output variable: Split of block S-SPLIT1 in extraction flowsheet. This sets the flow rate of steam to the dewatering mills turbine (stream SD5).

Parameters:

1. MILLS – Parameter 109: Steam needed by dewatering mills per kg of fibre in megasse. Value is 0.0639.

Calculation: FSTEAMM = MILLS $\times$ FMEGFIB

B.1.3 Total high pressure steam flow to extraction module motor turbines

Aspen Plus® block: SB1-FLOW

Description: Sets the flow rate of live steam (31 bara) to the extraction module for use in the turbines (knives, shredders and dewatering mills).

Variables:

1. FCANE – Input variable: Flow rate of cane (stream CANE) in kg/h.
2. FSTEAMK – Calculated variable: Flow rate of steam to the cane knives in kg/h.
3. FSTEAMS – Calculated variable: Flow rate of steam to the cane shredders in kg/h.
4. FMEGF – Input variable: Flow rate of fibre in the megasse (stream MEG) in kg/h.
5. FSTEAMM – Calculated variable: Flow rate of steam to the dewatering mills in kg/h.
6. FSTEAMTU – Output variable: Total flow rate of steam (stream SB1) required by extraction module turbines (knives, shredders and dewatering mills) in kg/h. Sets the split of block SB1-SPLT in the boiler module.

Parameters:

1. KNIVES – Parameter 103: Steam needed by cane knives per kg of cane processed. Value is 0.0207.
2. SHRDRS – Parameter 106: Steam needed by cane shredders per kg of cane processed. Value is 0.0621.
3. MILLS – Parameter 109: Steam needed by the dewatering mills per kg of fibre in megasse. Value is 0.0639.

Calculation:

FSTEAMK = KNIVES × FCANE

FSTEAMS = SHRDRS × FCANE

FSTEAMM = MILLS × FMEGF

FSTEAMTU = FSTEAMK + FSTEAMS + FSTEAMM

B.1.4 Imbibition usage by diffuser

Aspen Plus® block: IW-F

Description: Sets the flow rate of imbibition to the diffuser.

Variables:

1. FFIBREM – Input variable: Flow rate of fibre in megasse (stream MEG) in kg/h.
2. FIW – Output variable: Split of block C-SPLIT5 in the evaporation module. This sets the flow rate of imbibition to the diffuser.

Parameters:

1. IWFACTOR – Parameter 114: Amount of imbibition water usage per kg of megasse fibre processed. Value is 2.954.

Calculation: FIW = IWFACTOR × FFIBREM

B.1.5 Diffuser separation coefficient for water

Aspen Plus® block: DIFF-WAT

Description: Sets the extraction coefficient for water in the diffuser.

Variables:

1. IW – Input variable: Mass flow rate of imbibition (condensate) to the diffuser (stream IW) in kg/h.
2. SDI – Input variable: Mass flow rate of direct steam injection to the diffuser (stream SDI) in kg/h.
3. CANEW – Input variable: Flow rate of water in the cane feed (stream CANE) in kg/h.
4. PWH– Input variable: Flow rate of water in the hot press water (stream PWH) in kg/h.
5. SJHOT – Input variable: Mass flow rate of scalding juice to the diffuser (stream SJHOT) in kg/h.
6. MUW – Output variable: Split of block DIFFUSER in extraction flowsheet. This sets the split fraction of water which exits the diffuser in the draft juice stream.

Parameters:

1. SPLITWAT – Parameter 117: Extraction coefficient of water in diffuser excluding scalding juice. Value is 0.6642.

Calculation:

INNOSJ = IW + SDI + CANEW + PWH	… *All water fed to the diffuser besides the scalding juice.*
OUT = SPLITWAT × ***INNOSJ***	… *Water which should leave in the draft juice.*
TOTALIN = IW + SDI + CANEW + PWH + SJHOT	… *All water fed to diffuser including scalding juice.*
MUW = OUT/TOTALIN	… *Split fraction of water in diffuser to get specified water extraction.*

B.1.6 Steam flow (V2) to scalding juice heaters

Aspen Plus® block: SDH-FLOW

Description: Sets the steam flow rate to the scalding juice heaters.

Variables:

1. FCANE – Input variable: Flow rate of cane (stream CANE) in kg/h.
2. FSDH – Output variable: Split of block V-DIST2 in the evaporation module. This sets the flow rate of steam (stream SDH) to the scalding juice heaters.

Parameters:

1. SDH – Parameter 113: Amount of steam used in scalding juice heater per kg of cane. Value is 0.03541.

Calculation: FSDH = SDH × FCANE

B.1.7 Steam flow (V1) to diffuser

Aspen Plus® block: SDI-FLOW

Description: Sets the flow rate of steam injected into the diffuser.

Variables:

1. FCANE – Input variable: Flow rate of cane (stream CANE) in kg/h.
2. FSDI – Output variable: Split of block V-DIST2 in evaporation flowsheet. This sets the flow of steam (stream SDI) which is injected into the diffuser.

Parameters:

1. SDI – Parameter 112: Amount of steam injected into diffuser per kg of cane processed. Value is 0.01715.

Calculation: FSDI = SDI × FCANE

B.1.8 Temperature correction of draft juice and megasse

Aspen Plus® block: MEG-DJ-Q

Description: Transfers heat from the draft juice stream to the megasse stream (due to the streams exiting at different temperatures from a real diffuser).

Variables:

1. DJQ – Input variable: Calculated heat duty in draft juice cooler (block DJCOOL) in MJ/h.
2. MEGQ – Output variable: Sets the duty in the megasse heater (block MEGHEAT) in MJ/h.

Calculation: MEGQ = −DJQ

B.1.9 Dewatering mills separation coefficient for water

Aspen Plus® block: MILL-WAT

Description: Based on a specified moisture content in the bagasse, this block calculates the separation coefficient of water in the bagasse mills.

Variables:

1. FSUC – Input variable: Flow rate of sucrose in the megasse in kg/h.
2. FNSUC – Input variable: Flow rate of non-sucrose in the megasse in kg/h.
3. FFIB – Input variable: Flow rate of fibre in the megasse in kg/h.
4. FWAT – Input variable: Flow rate of water in the megasse in kg/h.
5. MUW – Output variable: Separation coefficient in block MILLS of water to the bagasse stream.

Parameters:

1. BW – Parameter 121: Mass fraction of water in bagasse. Value is 0.5096.
2. MUS – Parameter 122: Dewatering mill separation coefficient for sucrose (the fraction of sucrose inlet which leaves in the bagasse stream). Value is 0.1421.
3. MUNS – Parameter 123: Dewatering mill separation coefficient for non-sucrose (the fraction of non-sucrose inlet which leaves in the bagasse stream). Value is 0.1512.

4. MUFIB – Parameter 124: Dewatering mill separation coefficient for fibre (the fraction of fibre inlet which leaves in the bagasse stream). Value is 1.

Calculation:

$$\mathrm{MUW} = \frac{(\mathrm{MUS} \times \mathrm{FSUC} + \mathrm{MUNS} \times \mathrm{FNSUC} + \mathrm{MUFIB} \times \mathrm{FFIB}) \times \left(\frac{\mathrm{BW}}{1-\mathrm{BW}}\right)}{\mathrm{FWAT}}$$

B.1.10 Steam flow to press water tank

Aspen Plus® block: FSPW

Description: Based on a specified temperature in the press water holdup tank this calculator block determines the flow rate of steam required.

Variables:

1. FPW – Input variable: Flow rate of press water in kg/h.
2. HPW – Enthalpy flow of press water stream in kJ/kg.
3. HSPW – Enthalpy flow of steam to the press water tank (stream SPW) in kJ/kg.
4. FSPW – Export variable: Sets the split of vapour bleed splitter V-DIST1 in the evaporation module to ensure the required flow rate of steam to the press water tank.

Parameters:

1. HPWH – Parameter 116: Press water temperature after heating. Value is 69.36 ^{o}C. The enthalpy of the hot press water was determined at this temperature.

Calculation:

$$FSPW = \frac{HPWH \times FPW}{HSPW - HPW - HPWH}$$

B.2 Clarification module

PRI-HX Steam flow to primary mixed juice heater	SEC-HX Steam flow to secondary mixed juice heater	TERT-HX Steam flow to tertiary mixed juice heater	LIM-FLOW Milk-of-lime flow
W-MUD Water separation coefficient in clarifier	BGO-FLOW Bagacillo (modelled as bagasse) flow	WW-FILT Wash water flow to vacuum filter	W-CAKE Water separation coefficient in vacuum filter

Figure B.2 Clarification module calculator blocks

B.2.1 Steam flow (V3) to primary mixed juice heater

Aspen Plus® block: PRI-HX

Description: Calculates the flow rate of steam required by the primary mixed juice heater using an energy balance in order for the juice to reach 77.23 °C (parameter 205).

Variables:

1. FLOWMJ0 – Input variable: Flow rate of mixed juice (stream MJ) in kg/h.
2. MJ0H – Input variable: Enthalpy of stream MJ in kJ/kg.
3. MJ1H – Input variable: Enthalpy of stream MJ at 77.23 °C in kJ/kg.
4. ENTHSHP – Input variable: Enthalpy of steam (V3) to primary mixed juice heater in kJ/kg.
5. ENTHCHP – Input variable: Enthalpy of condensate (stream CHP) from primary mixed juice heater in kJ/kg.
6. FLOWSHP – Output variable: Split of block FSSPLIT in evaporation flowsheet. This sets the flow rate of steam (stream SHP) to the primary mixed juice heater.

Calculation:

$$\text{FLOWSHP} = \frac{(\text{MJ1H} - \text{MJ0H}) \times \text{FLOWMJ0}}{\text{ENTHSHP} - \text{ENTHCHP}}$$

B.2.2 Steam flow (V2) to secondary mixed juice heater

Aspen Plus® block: SEC-HX

Description: Calculates the flow rate of steam required by the secondary mixed juice heater using an energy balance in order for the juice to reach 93.85 °C (parameter 206).

Variables:

1. MJPFLOW – Input variable: Flow rate of mixed juice (stream MJP) from mixed juice pump in kg/h.
2. MJPH – Input variable: Enthalpy of MJP in kJ/kg.
3. MJPH95 – Input variable: Enthalpy of MJP at 93.85 °C in kJ/kg.
4. ENTHSHS – Input variable: Enthalpy of steam (V2) to secondary mixed juice heater in kJ/kg.
5. ENTHCHS – Input variable: Enthalpy of condensate (stream CHS) from secondary mixed juice heater in kJ/kg.
6. SHSFLOW – Output variable: Split of block V-DIST2 in evaporation flowsheet. This sets the flow rate of steam (stream SHS) to the secondary heat exchanger.

Calculation:

$$\text{SHSFLOW} = \frac{(\text{MJPH95} - \text{MJPH}) \times \text{MJPFLOW}}{\text{ENTHSHS} - \text{ENTHCHS}}$$

B.2.3 Steam flow (V1) to tertiary mixed juice heater

Aspen Plus® block: TERT-HX

Description: Calculates the flow rate of steam required by the tertiary mixed juice heater using an energy balance in order for the juice to reach 103.89 °C (parameter 207).

Variables:

1. MJ2FLOW – Input variable: Flow rate of mixed juice (stream MJ2) from secondary heat exchanger in kg/h.
2. MJ2H – Input variable: Enthalpy of MJ2 in kJ/kg.
3. MJ2H105 – Input variable: Enthalpy of MJ2 at 103.89 °C in kJ/kg.
4. ENTHSHT – Input variable: Enthalpy of steam (V1) to tertiary mixed juice heater in kJ/kg.
5. ENTHCHT – Input variable: Enthalpy of condensate (stream CHT) from tertiary mixed juice heater in kJ/kg.
6. FLOWSHT – Output variable: Split of block V-DIST1 in evaporation flowsheet. This sets the flow rate of steam (stream SHT) to the tertiary heat exchanger.

Calculation:

$$\text{FLOWSHT} = \frac{(\text{MJ2H105} - \text{MJ2H}) \times \text{MJ2FLOW}}{\text{ENTHSHT} - \text{ENTHCHT}}$$

B.2.4 Milk-of-lime flow to limer

Aspen Plus® block: LIM-FLOW

Description: Calculates the flow rate of lime required. Firstly, it takes into account the lime already in the mixed juice (stream MJ1) due to the recycle from the vacuum filter.

Variables:

1. MJ1FLOW – Input variable: Flow rate of mixed juice (stream MJ1) from primary mixed juice heater in kg/h.
2. MJ1LIME – Input variable: Flow rate of lime in mixed juice (stream MJ1) in kg/h.
3. LIMFLOW – Output variable: Flow rate of lime required (stream LIM) in kg/h.

Parameters:

1. LIMING – Parameter 204: Mass fraction of lime required in mixed juice after liming (stream MJL). Value is 0.002884.
2. LIMFRAC – Parameter 203: Mass fraction of lime in milk-of-lime (LIM stream). Value is 0.1.

Calculation:

$$\text{LIMFLOW} = \frac{\text{MJ1FLOW} \times (\text{MJ1LIME} - \text{LIMING})}{\text{LIMING} - \text{LIMFRAC}}$$

B.2.5 Clarifier separation coefficient for water

Aspen Plus® block: W-MUD

Description: Calculates the split fraction of water in the clarifier which is necessary to maintain a desired mass fraction of water in the mud leaving the clarifier.

Variables:

1. SUC – Input variable: Mass flow rate of sucrose in the mixed juice stream from the flash vessel (stream MJF) in kg/h.
2. NSUC – Input variable: Mass flow rate of non-sucrose in stream MJF in kg/h.
3. FIB – Input variable: Mass flow rate of fibre in stream MJF in kg/h.
4. LIM – Input variable: Mass flow rate of lime in stream MJF in kg/h.
5. FW – Input variable: Mass flow rate of water in stream MJF in kg/h.
6. WSPLIT – Output variable: Split fraction of water in clarifier (block CLARIFY). This sets the split fraction of water which goes to the mud stream.

Parameters:

1. MUSUC – Parameter 211: Clarifier separation coefficient for sucrose (the fraction of sucrose in the inlet which leaves in the mud stream). Value is 0.1154.
2. MUNSUC – Parameter 212: Clarifier separation coefficient for non-sucrose. Value is 0.1573.
3. MUFIB – Parameter 213: Clarifier separation coefficient for fibre. Value is 1.
4. MULIM – Parameter 214: Clarifier separation coefficient for lime. Value is 1.
5. WMUD – Parameter 215: Specified mass fraction of water in the final mud stream leaving the clarifier. Value is 0.8041.

Calculation:

TOTDRY = SUC × MUSUC + NSUC × MUNSUC + FIB × MUFIB + LIM × MULIM

$$\mathrm{WSPLIT} = \frac{\frac{\boldsymbol{TOTDRY}}{1-\mathrm{WMUD}} - \boldsymbol{TOTDRY}}{\mathrm{FW}}$$

B.2.6 Bagacillo flow rate to mud-bagacillo blender

Aspen Plus® block: W-MUD

Description: Sets the flow rate of bagacillo (modelled as bagasse) to the mud-bagacillo blender.

Variables:

1. FMUD – Input variable: Flow rate of mud (stream MUD) in kg/h.
2. FBGO – Output variable: Split fraction of block BGO-SPLT in boiler flowsheet. This sets the amount of bagasse which goes to the mud-bagacillo blender.

Parameters:

1. BGO2MUD – Parameter 216: Ratio of bagacillo to mud flow. Value is 0.02743.

Calculation: FBGO = BGO2MUD × FMUD

B.2.7 Wash water required by vacuum filter

Aspen Plus® block: WW-FILT

Description: Calculates the required flow rate of wash water to the vacuum filter. The flow rate of wash water is proportional to the flow rate of insolubles (fibre) in filter cake.

Variables:

1. FFIB – Input variable: Mass flow rate of fibre entering the vacuum filter (stream MB) in kg/h.
2. SEPFIB – Input variable: Separation coefficient of fibre in the vacuum filter (block VACFIL).

3. WW1FLO – Output variable: Split fraction of block C-SPLIT5 in evaporation flowsheet. This sets the flow rate of wash water to the vacuum filter (stream WW1).
4. FLIM – Input variable: Mass flow rate of lime entering the vacuum filter (stream MB) in kg/h.
5. SEPLIM – Input variable: Separation coefficient of lime in the vacuum filter (block VACFIL).

Parameters:

1. WWUSAGE – Parameter 222: The amount of wash water used by the filter per kg of fibre in the filter cake. Value is 1.983.

Calculation: WW1FLO = (FFIB × SEPFIB + FLIM × SEPLIM) × WWUSAGE

B.2.8 Vacuum filter separation coefficient for water

Aspen Plus® block: W-CAKE

Description: Calculates the required separation coefficient of water in the vacuum filter to maintain a specified mass fraction of water in the filter cake.

Variables:

1. SUC – Input variable: Flow rate of sucrose to the vacuum filter (in stream MB) in kg/h.
2. NSUC – Input variable: Flow rate of non-sucrose entering the vacuum filter in kg/h.
3. W – Input variable: Flow rate of water entering the vacuum filter in kg/h.
4. LIM – Input variable: Flow rate of lime entering the vacuum filter in kg/h.
5. FIB – Input variable: Flow rate of fibre entering the vacuum filter in kg/h.
6. FF – Input variable: Separation coefficient of fibre in vacuum filter which leaves in filter cake.
7. FS – Input variable: Separation coefficient of sucrose in vacuum filter.
8. FNS – Input variable: Separation coefficient of non-sucrose in vacuum filter.
9. FL – Input variable: Separation coefficient of lime in vacuum filter.
10. WW – Input variable: Flow rate of wash water to the vacuum filter in kg/h.
11. WF – Output variable: Separation coefficient of water in vacuum filter.

Parameters:

1. WC – Parameter 217: Mass fraction of water in the filter cake. Value is 0.7000.

Calculation:

$\boldsymbol{TOTALDRY} = \mathrm{FS} \times \mathrm{SUC} + \mathrm{FNS} \times \mathrm{NSUC} + \mathrm{FF} \times \mathrm{FIB} + \mathrm{FL} \times \mathrm{LIM}$

$$\mathrm{WF} = \frac{\frac{\boldsymbol{TOTALDRY}}{1-\mathrm{WC}} - \boldsymbol{TOTALDRY}}{\mathrm{W}+\mathrm{WW}}$$

B.3 Evaporation module

B.3.1 Exhaust steam flow rate to the preheater and 1st effect evaporator

Aspen Plus® block: EVAP

Description: Manipulates the exhaust steam to the preheater and 1st effect evaporator.

B.3.1.1 Part 1: Energy balance around the preheater.

Variables:

1. FLOW CJ – Input variable: Flow rate of clear juice to the preheater (stream CJ) in kg/h.
2. CJ enthalpy – Input variable: Enthalpy of stream CJ in kJ/kg.
3. L0 enthalpy – Input variable: Enthalpy of stream CJ at 112.5 °C (Parameter 317) in kJ/kg.
4. SEH enthalpy – Input variable: Enthalpy of exhaust steam to preheater (stream SEH) in kJ/kg.
5. CEH enthalpy – Input variable: Enthalpy of condensate from preheater (stream CEH) in kJ/kg.
6. FLOW SEH – Calculated variable: Required flow rate of exhaust steam to preheater in order to heat clear juice to 112.5 °C (Parameter 317).

Calculation:

$$\text{FLOW SEH} = \frac{(\text{L0 enthalpy} - \text{CJ enthalpy}) \times \text{FLOW CJ}}{\text{SEH enthalpy} - \text{CEH enthalpy}}$$

B.3.1.2 Part 2: Initial guess of flow rate of exhaust steam to the 1st effect evaporator.

Variables:

1. L0 – Input variable: Flow rate of clear juice from the preheater to the 1st effect evaporator (stream L0) in kg/h.
2. L0-W – Input variable: Flow rate of water in L0 (kg/h).
3. L0-S – Input variable: Flow rate of sucrose in L0 (kg/h).
4. L0-NS – Input variable: Flow rate of non-sucrose in L0 (kg/h).
5. Brix – Calculated variable: Brix of L0 expressed as a fraction.
6. Predicted L5 – Calculated variable: Predicted flow rate of final syrup (L5) in kg/h.
7. Vtotal – Calculated variable: Predicted total vapour evaporated in all five effects in kg/h.
8. S1 predicted – Calculated variable: Predicted flow rate of vapour evaporated in the first effect which is sent to the second effect (kg/h).

9. Bleed 1 – Calculated variable: Predicted flow rate of vapour bleed from first effect (kg/h).
10. V1 predicted – Calculated variable: Predicted flow rate of vapour evaporated in the first effect in kg/h.
11. S0 predicted – Calculated variable: Predicted flow rate of exhaust steam to the first effect in kg/h.

Parameters:

1. Entrain – Parameter 346: Fraction of droplet entrainment in the first effect. Value is 0.0077352.
2. Brix L5 Desired – Parameter 318: Final syrup dry substance expressed as a fraction. Value is 0.648.

Calculations:

$$\text{Brix} = \frac{\text{L0-S} + \text{L0-NS}}{\text{L0}}$$

$$\text{Predicted L5} = \frac{\text{L0-S} + \text{L0-NS} - 2 \times \text{Entrain} \times (\text{L0-S} + \text{L0-NS})}{\text{Brix L5 Desired}}$$

$$\text{Vtotal} = \text{L0} - \text{Predicted L5}$$

$$\text{S1 predicted} = \frac{\text{Vtotal}}{5}$$

$$\text{Bleed 1} = 0.01715 \times F_{cane} + 0.103133 \times \text{Predicted L5} + 0.003589 \times F_{cane} + 0.344096 \times \text{Predicted L5} + 0.018975 \times \mathrm{F_{MJ}}$$

Comment on bleed 1 calculation: Vapour bleeds were related to known flow rates in order to predict overall bleed from first effect. F_{cane} is the flow rate of cane and F_{MJ} is the flow rate of mixed juice (all in kg/h). The five terms are the SDI, SKB, SPW, SKA and SHT bleeds.

$$\text{V1 predicted} = \text{S1 predicted} + \text{Bleed 1}$$

$$\text{S0 predicted} = \frac{\text{V1 predicted}}{0.975}$$

This predicted value for S0 can then be used to get an initial convergence of the evaporation module. The next step is to match the STEAMMU and STEAMOUT flows.

This is done in part 3 where a PI controller manipulates the S0 flow in order to get STEAMOUT as close to 7230 kg/h (The value of STEAMMU).

B.3.1.3 Part 3: Manipulation of exhaust steam flow rate to the 1st effect evaporator.

Variables:

1. Counter – Excel variable: Counts the number of iterations which the spreadsheet has gone through.
2. Process Variable – Input variable: Flow rate of fictitious stream STEAMOUT in kg/h.
3. Error (%) – Calculated variable: "The difference between the process variable and the set point expressed as a percentage of the Input Range"
4. Integrator – Calculated variable: "The "accumulator" that sums the accumulating integral action of the controller - this has a "bumpless transfer" action that sets this accumulator to the appropriate value when the control is in manual mode".
5. Output (%) – Calculated variable: "The output from the controller (as a percentage from 0 to 100%)".
6. Output - eng units – Calculated variable: "The current value of the variable being manipulated (in engineering units) - regardless of whether the controller is in manual or automatic mode".
7. Steammu – Input variable: Flow rate of fictitious steam (stream STEAMMU) in kg/h.
8. Steamout – Input variable: Flow rate of steam out (stream STEAMOUT) in order to supply sufficient steam (SHP) to primary mixed juice heater (kg/h).

Parameters:

1. Set Point: Value which the flow rate of stream STEAMOUT needs to be. Value is 7230 kg/h.
2. Input Range: The range which the flow rate of STEAMOUT is expected to vary. Value is 14460 kg/h.
3. Cycles per update: Only after a certain number of spreadsheet iterations, the controller output will be updated. Value is 10.
4. Gain: The proportionality constant for the control action (based on the concept of the gain of a conventional PI controller). Value is -0.01.
5. Integral Factor: A factor to change the extent of "integral action of the controller" - equivalent to the "integral time" of a standard PI controller. Value is 10.
6. Output – min eng. units: The value of the variable being manipulated (in engineering units) that corresponds to a 0% controller output. Value is 40000 kg/h.
7. Output – max eng. units: The value of the variable being manipulated (in engineering units) that corresponds to a 100% controller output. Value is 100000 kg/h.

8. Initial/Manual Output: The value of the variable being manipulated (in engineering units) that must be set if the controller is in manual mode. Value is 83568 kg/h.

Calculations:

$$\text{Error (\%)} = \frac{\text{Process Variable} - \text{Set Point}}{\text{Input Range}} \times 100$$

Integrator = IF(OR(Initialise="I";D18="M");+((Initial/Manual Output - Output – min eng. units)/(Output – max eng. units - Output – min eng. units) ×100) / Gain - Error (%);Integrator +1/ Integral Factor × Error (%))

Comments: The logic test: IF(OR(Initialise="I";D18="M") *means that if the calculation is in initialise mode or the controller is in manual mode then the* Integrator = +((Initial/Manual Output - Output – min eng. units)/(Output – max eng. units - Output – min eng. units) ×100) / Gain - Error (%)

Otherwise if the calculation is in calculate mode and the controller is in automatic then the Integrator = Integrator +1/ Integral Factor × Error (%))

Output (%) = MINA(MAXA(0;(IF(OR(MOD(Counter; Cycles per update)=0;D18="M");+ Gain × (Error (%)+Integrator); Output (%))));100)

Comments:

- *The output is limited between 0 and 100% by the* MINA(MAXA(0;…;100) *statement.*
- *The* MOD(Counter; Cycles per update)=0 *statement means that the* Output (%) *is only changed once every 10 iterations (since* Cycles per update = 10*).*
- *The logic test:* IF(OR(MOD(Counter; Cycles per update)=0;D18="M") *means that if the* Counter *is a multiple of 10 or the controller is in manual mode the* Output (%) = + Gain × (Error (%)+Integrator).

  Otherwise, if the Counter *is not a multiple of 10 and the controller is in automatic mode then the* Output (%) = Output (%) *(It remains unchanged)*

Output - eng units =IF(OR(Initialise="I";D18="M"); Initial/Manual Output;+ Output – min eng. units +(Output – max eng. units - Output – min eng. units) × Output (%) /100)

Comments:

- *The logic test:* IF(OR(Initialise="I";D18="M") *means that if the calculation is in initialise mode or the controller is in manual mode then the* Output - eng units = Initial/Manual Output.
- *Otherwise, if the calculation is in calculate mode and the controller is in automatic then the* Output - eng units = + Output – min eng. units + (Output – max eng. units - Output – min eng. units) × Output (%) /100

B.3.1.4 Part 4: Manipulation of exhaust steam splitter (SSPLIT1) which feeds the preheater and 1st effect evaporator.

Variables:

1. Set S0 – Output variable: Sets the split of SSPLIT1 in evaporation module. This sets the flow rate of exhaust steam to the 1st effect evaporator.
2. Flow SET – Output variable: Sum of required flow rates of exhaust steam to preheater and first effect evaporator (kg/h). Sets the flow rate of stream SET.

Calculations:

Set S0 = Output - eng units *(Calculated in part 3)*

Flow SET = FLOW SEH *(Calculated in part 1)* + Set S0

B.4 Crystallisation module

PANA-ENT Entrainment in 'A' pans	PANA-CX Crystallisation in 'A' pans	A-CX Crystallisation in 'A' cooling crystallisers	PANB-ENT Entrainment in 'B' pans	PANB-CX Crystallisation in 'B' pans
B-CX Crystallisation in 'B' cooling crystallisers	PANC-ENT Entrainment in 'C' pans	PANC-CX Crystallisation in 'C' pans	C-CX Crystallisation in 'C' cooling crystallisers	CX-MINGL Crystal dissolution in magma mingler

Figure B.3 Crystallisation module calculator blocks

B.4.1 Entrainment in 'A' pans

Aspen Plus® block: PANA-ENT

Description: Determines the flow rate of entrainment in the 'A' pans. Before entrainment can be calculated, the entire pan calculation must be done first (crystallisation and evaporation). This is necessary to determine the composition of the mother liquor of the 'A' massecuite when it exits the pans. The entrainment values are then calculated based on the ratios of the components in the mother liquor. The outputs manipulate the initialised values for the separation coefficients in the block which handles the entrainment separation (ENT-A).

Calculator blocks PANB-ENT and PANC-ENT are based on the same principles.

Variables:

1. S – Calculated variable: Pure sucrose solubility. It is a function of t_c.
2. $(SW)^{sat}_{pure}$ – Calculated variable: Sucrose-to-water ratio (another way of expressing solubility). Also just a function of t_c.
3. Non-sucrose in mother liquor – Input variable: Flow rate of non-sucrose in A-INV stream (kg/h).
4. Feed to 'A' pans – Input variable: Mass flow rate of A-TOTAL stream (kg/h).
5. Water in – Input variable: Mass flow rate of water in A-INV stream in kg/h (after inversion).
6. Water in mother liquor – calculated variable: Water in massecuite out from 'A' pans. Function of Dry substance, Water in and Flow of feed to 'A' pans.
7. NSW^{ml} – Calculated variable: (Non-sucrose)-to-water ratio in the 'A' mother liquor. Function of Non-sucrose in mother liquor and Water in mother liquor.
8. Sat. Coeff. (SC) – Calculated variable: The saturation coefficient of mother liquor when the massecuite leaves the 'A' pans. Function of A, B0, B1, C and NSW^{ml}.
9. Sucrose in feed – Input variable: Flow rate of sucrose in A-TOTAL stream (which would leave in the mother liquor if no crystallisation occurred).
10. $(SW^{mol})_{impure}$ – Calculated variable: Sucrose-to-water ratio in an impure solution (mother liquor). Function of Sucrose in feed and Water in mother liquor.
11. $(SW^{mol})^{sat}_{impure}$ – Calculated variable: The sucrose-to-water ratio for a saturated impure solution. Function of Sat. Coeff. (SC) and $(SW)^{sat}_{pure}$.
12. SS actual – Calculated variable: The current supersaturation (before crystallisation has occurred. Function of $(SW^{mol})_{impure}$ and $(SW^{mol})^{sat}_{impure}$.

13. Sucrose required – Calculated variable: The amount of sucrose which would be left in the mother liquor to give the desired supersaturation (SS required). Function of SS actual, SS required and Sucrose in feed.
14. Crystal in – Input variable: Flow rate of crystals in A-INV stream in kg/h (comes from magma).
15. Sucrose in (suc+cryst) – Calculated variable: Sucrose in feed added to Crystal in.
16. Sucrose entrained – Calculated variable: Entrainment parameter multiplied by Sucrose in (suc+cryst).
17. Sucrose into separator – Input variable: Flow rate of sucrose in A-INV stream (kg/h).
18. Sucrose sep. coeff. – Output variable: The fraction of sucrose fed to block ENT-A which ends up in the A-ENT stream (The fraction which is not entrained). Function of Sucrose into separator and Sucrose entrained.
19. Sucrose sep. – Calculated variable: Sucrose entrained divided by Sucrose required (Sucrose entrained as a fraction of the sucrose out of the pan in the mother liquor of the 'A' massecuite).
20. Water sep. – Calculated variable: Sucrose sep. multiplied by Water in mother liquor (The ratios of entrainment are calculated based on the ratios which the components are present in the mother liquor of the 'A' massecuite).
21. Water sep. coeff. – Output variable: The fraction of water fed to block ENT-A which ends up in the A-ENT stream (The fraction which is not entrained). Function of Water in and Water sep.
22. Non-suc sep. – Calculated variable: Sucrose sep. multiplied by Non-sucrose in mother liquor.
23. Non-suc sep. coeff. – Output variable: The fraction of non-sucrose fed to block ENT-A which ends up in the A-ENT stream (The fraction which is not entrained). Function of Non-sucrose in mother liquor and Non-suc sep.

Constants:

- A, B0, B1 and C: Constants in the saturation coefficient (SC) equation.

Parameters:

1. t_c: Exit temperature of massecuite from 'A' pans.
2. Dry substance: The fraction which is not water in the massecuite from 'A' pans.
3. SS required – Exit supersaturation of 'A' massecuite.
4. Entrainment – Specification of entrainment in 'A' pans. Fraction of total sucrose in (including crystal) which is entrained in the vapour.

Calculations:

Pure sucrose solubility:

$$S = 64.447 + 0.08222t_c + 1.6169 \times 10^{-3}t_c^2 - 1.559 \times 10^{-6}t_c^3 - 4.63 \times 10^{-8}t_c^4$$

Sucrose-to-water ratio (pure solution):

$$(\mathrm{SW})_{\mathrm{pure}}^{\mathrm{sat}} = \frac{S}{100 - S}$$

(Non-sucrose)-to-water ratio:

$$\text{Water in mother liquor} = \frac{\text{Feed to 'A' pans} - \text{water in}}{\text{Dry substance}} - \text{Feed to 'A' pans} - \text{Water in}$$

$$\mathrm{NSW}^{\mathrm{ml}} = \frac{\text{Non} - \text{sucrose in mother liquor}}{\text{Water in mother liquor}}$$

Saturation coefficient:

$$\text{Sat. Coeff. (SC)} = A \times \mathrm{NSW}^{\mathrm{ml}} + B0 - B1 \times t_c + (1 - B0 + B1 \times t_c)\exp(-C \times \mathrm{NSW}^{\mathrm{ml}})$$

Sucrose-to-water ratio (impure solution):

$$\left(\mathrm{SW}^{\mathrm{mol}}\right)_{\mathrm{impure}} = \frac{\text{Sucrose in feed}}{\text{Water in mother liquor}}$$

Sucrose-to-water ratio (saturated impure solution):

$$\left(\mathrm{SW}^{\mathrm{mol}}\right)_{\mathrm{impure}}^{\mathrm{sat}} = (\mathrm{SW})_{\mathrm{pure}}^{\mathrm{sat}} \times \text{Sat. Coeff. (SC)}$$

Supersaturation:

$$\text{SS actual} = \frac{\left(\mathrm{SW}^{\mathrm{mol}}\right)_{\mathrm{impure}}}{\left(\mathrm{SW}^{\mathrm{mol}}\right)_{\mathrm{impure}}^{\mathrm{sat}}}$$

Sucrose required to meet specified exit supersaturation condition:

$$\text{Sucrose required} = \frac{\text{SS required}}{\text{SS actual}} \times \text{Sucrose in feed}$$

Total sucrose feed (including crystals):

$$\text{Sucrose in (suc + cryst)} = \text{Sucrose in feed} + \text{Crystal in}$$

Entrainment:

$$\text{Sucrose entrained} = \text{Entrainment} \times \text{Sucrose in (suc + cryst)}$$

$$\text{Sucrose sep. coeff.} = 1 - \frac{\text{Sucrose entrained}}{\text{Sucrose into separator}}$$

$$\text{Sucrose sep.} = \frac{\text{Sucrose entrained}}{\text{Sucrose required}}$$

The ratios of entrainment are calculated based on the ratios which the components are present in the mother liquor of the 'A' massecuite:

$$\text{Water sep.} = \text{Sucrose sep.} \times \text{Water in mother liquor}$$

$$\text{Water sep. coeff.} = 1 - \frac{\text{Water sep.}}{\text{Water in}}$$

$$\text{Non} - \text{suc sep.} = \text{Sucrose sep.} \times \text{Non} - \text{sucrose in mother liquor}$$

$$\text{Non} - \text{suc sep. coeff.} = 1 - \frac{\text{Non} - \text{suc sep.}}{\text{Non} - \text{sucrose in mother liquor}}$$

B.4.2 Crystallisation in 'A' pans

Aspen Plus® block: PANA-CX

Description: Determines the extent of crystallisation in 'A' pans. The crystallisation of sucrose in the 'A' pans is based on Solid-Liquid-Equilibria. The calculation follows the algorithm of Starzak (2015 and 2016a).

Calculator blocks PANB-CX and PANC-CX are based on the same principles.

Variables:

1. S – Calculated variable: Pure sucrose solubility. It is a function of t_c.
2. $(SW)^{sat}_{pure}$ – Calculated variable: Sucrose-to-water ratio (another way of expressing solubility). Also just a function of t_c.
3. Non-sucrose in mother liquor – Input variable: Flow rate of non-sucrose in A-INV stream (kg/h).
4. Feed to 'A' pans – Input variable: Mass flow rate of A-TOTAL stream (kg/h).
5. Water in – Input variable: Mass flow rate of water in A-INV stream in kg/h (after inversion).
6. Water in mother liquor – calculated variable: Water in massecuite out from 'A' pans. Function of Dry substance, Water in and Flow of feed to 'A' pans.

7. NSW^{ml} – Calculated variable: (Non-sucrose)-to-water ratio in the 'A' mother liquor. Function of Non-sucrose in mother liquor and Water in mother liquor.
8. Sat. Coeff. (SC) – Calculated variable: The saturation coefficient of mother liquor when the massecuite leaves the 'A' pans. Function of A, B0, B1, C and NSW^{ml}.
9. Sucrose in feed – Input variable: Flow rate of sucrose in A-TOTAL stream (which would leave in the mother liquor if no crystallisation occurred).
10. $(SW^{mol})_{impure}$ – Calculated variable: Sucrose-to-water ratio in an impure solution (mother liquor). Function of Sucrose in feed and Water in mother liquor.
11. $(SW^{mol})^{sat}_{impure}$ – Calculated variable: The sucrose-to-water ratio for a saturated impure solution. Function of Sat. Coeff. (SC) and $(SW)^{sat}_{pure}$.
12. SS actual – Calculated variable: The current supersaturation (before crystallisation has occurred. Function of $(SW^{mol})_{impure}$ and $(SW^{mol})^{sat}_{impure}$.
13. Sucrose required – Calculated variable: The amount of sucrose which would be left in the mother liquor to give the desired supersaturation (SS required). Function of SS actual, SS required and Sucrose in feed.
14. Sucrose into A-Pan-CX – Input variable: Mass flow rate of sucrose in A-ENT stream in kg/h (flow rate of sucrose into reactor block where crystallisation is modelled)
15. Sucrose conversion – Output variable: The fraction of sucrose in the feed which is crystallised. Sets the conversion of block A-PAN-CX.

Constants:

- A, B0, B1 and C: Constants in the saturation coefficient (SC) equation.

Parameters:

1. t_c: Exit temperature of massecuite from 'A' pans.
2. Dry substance: The fraction which is not water in the massecuite from 'A' pans.
3. SS required – Exit supersaturation of 'A' massecuite.

Calculations:

Pure sucrose solubility:

$$S = 64.447 + 0.08222t_c + 1.6169 \times 10^{-3}t_c^2 - 1.559 \times 10^{-6}t_c^3 - 4.63 \times 10^{-8}t_c^4$$

Sucrose-to-water ratio (pure solution):

$$(\mathrm{SW})^{\mathrm{sat}}_{\mathrm{pure}} = \frac{S}{100 - S}$$

(Non-sucrose)-to-water ratio:

$$\text{Water in mother liquor} = \frac{\text{Feed to 'A' pans} - \text{water in}}{\text{Dry substance}} - \text{Feed to 'A' pans} - \text{Water in}$$

$$\text{NSW}^{\text{ml}} = \frac{\text{Non} - \text{sucrose in mother liquor}}{\text{Water in mother liquor}}$$

Saturation coefficient:

$$\text{Sat. Coeff. (SC)} = A \times \text{NSW}^{\text{ml}} + B0 - B1 \times t_c + (1 - B0 + B1 \times t_c)\exp(-C \times \text{NSW}^{\text{ml}})$$

Sucrose-to-water ratio (impure solution):

$$\left(\text{SW}^{\text{mol}}\right)_{\text{impure}} = \frac{\text{Sucrose in feed}}{\text{Water in mother liquor}}$$

Sucrose-to-water ratio (saturated impure solution):

$$\left(\text{SW}^{\text{mol}}\right)_{\text{impure}}^{\text{sat}} = (\text{SW})_{\text{pure}}^{\text{sat}} \times \text{Sat. Coeff. (SC)}$$

Supersaturation:

$$\text{SS actual} = \frac{\left(\text{SW}^{\text{mol}}\right)_{\text{impure}}}{\left(\text{SW}^{\text{mol}}\right)_{\text{impure}}^{\text{sat}}}$$

Sucrose required to meet specified exit supersaturation condition:

$$\text{Sucrose required} = \frac{\text{SS required}}{\text{SS actual}} \times \text{Sucrose in feed}$$

Sucrose conversion:

$$\text{Sucrose conversion} = \frac{\text{Sucrose into A} - \text{Pan} - \text{CX} \quad - \quad \text{Sucrose required}}{\text{Sucrose into A} - \text{Pan} - \text{CX}}$$

B.4.3 Crystallisation in 'A' cooling crystallisers

Aspen Plus® block: A-CX

Description: Determines the extent of crystallisation in the 'A' cooling crystallisers. The crystallisation of sucrose in the crystallisers is based on Solid-Liquid-Equilibria. The calculation follows the algorithm of Starzak (2015 and 2016a).

Calculator blocks B-CX and C-CX are based on the same principles.

Variables:

1. S – Calculated variable: Pure sucrose solubility. It is a function of t_c.

2. $(SW)^{sat}_{pure}$ – Calculated variable: Sucrose-to-water ratio (another way of expressing solubility). Also just a function of t_c.
3. Non-sucrose in massecuite – Input variable: Flow rate of non-sucrose in the 'A' massecuite (stream MA) in kg/h.
4. Water in massecuite – Input variable: Mass flow rate of water in MA stream (kg/h).
5. NSW^{mas} – Calculated variable: (Non-sucrose)-to-water ratio in the 'A' mother liquor. Function of Non-sucrose in massecuite and Water in massecuite.
6. Sat. Coeff. (SC) – Calculated variable: The saturation coefficient of massecuite. Function of A, B0, B1, C and NSW^{mas}.
7. Sucrose in massecuite – Input variable: Flow rate of sucrose in MA stream (kg/h).
8. $(SW^{mas})_{impure}$ – Calculated variable: Sucrose-to-water ratio in an impure solution (mother liquor). Function of Sucrose in massecuite and Water in massecuite.
9. $(SW^{mas})^{sat}_{impure}$ – Calculated variable: The sucrose-to-water ratio for a saturated impure solution. Function of Sat. Coeff. (SC) and $(SW)^{sat}_{pure}$.
10. SS actual – Calculated variable: The current supersaturation (before crystallisation has occurred. Function of $(SW^{mas})_{impure}$ and $(SW^{mas})^{sat}_{impure}$.
11. Sucrose required – Calculated variable: The amount of sucrose which would be left in the massecuite to give the desired supersaturation (SS required). Function of SS actual, SS required and Sucrose in massecuite.
12. Sucrose conversion – Output variable: The fraction of sucrose in the feed massecuite which is crystallised. Sets the conversion of block A-CX.

Constants:

- A, B0, B1 and C: Constants in the saturation coefficient (SC) equation.

Parameters:

1. t_c: Exit temperature of massecuite from 'A' crystallisers.
2. SS required – Exit supersaturation of 'A' massecuite from 'A' cooling crystallisers.

Calculation:

Pure sucrose solubility:

$$S = 64.447 + 0.08222t_c + 1.6169 \times 10^{-3}t_c^2 - 1.559 \times 10^{-6}t_c^3 - 4.63 \times 10^{-8}t_c^4$$

Sucrose-to-water ratio (pure solution):

$$(SW)^{sat}_{pure} = \frac{S}{100 - S}$$

(Non-sucrose)-to-water ratio:

$$\mathrm{NSW}^{\mathrm{mas}} = \frac{\text{Non} - \text{sucrose in massecuite}}{\text{Water in massecuite}}$$

Saturation coefficient:

$$\text{Sat. Coeff. (SC)} = A \times \mathrm{NSW}^{\mathrm{mas}} + B0 - B1 \times t_c + (1 - B0 + B1 \times t_c)\exp(-C \times \mathrm{NSW}^{\mathrm{mas}})$$

Sucrose-to-water ratio (impure solution):

$$(\mathrm{SW}^{\mathrm{mas}})_{\mathrm{impure}} = \frac{\text{Sucrose in feed}}{\text{Water in massecuite}}$$

Sucrose-to-water ratio (saturated impure solution):

$$(\mathrm{SW}^{\mathrm{mas}})_{\mathrm{impure}}^{\mathrm{sat}} = (\mathrm{SW})_{\mathrm{pure}}^{\mathrm{sat}} \times \text{Sat. Coeff. (SC)}$$

Supersaturation:

$$\text{SS actual} = \frac{(\mathrm{SW}^{\mathrm{mas}})_{\mathrm{impure}}}{(\mathrm{SW}^{\mathrm{mas}})_{\mathrm{impure}}^{\mathrm{sat}}}$$

Sucrose required to meet specified exit supersaturation condition:

$$\text{Sucrose required} = \frac{\text{SS required}}{\text{SS actual}} \times \text{Sucrose in massecuite}$$

Sucrose conversion:

$$\text{Sucrose conversion} = \frac{\text{Sucrose in massecuite} - \text{Sucrose required}}{\text{Sucrose in massecuite}}$$

B.4.4 Crystal dissolution in magma mingler

Aspen Plus® block: CX-MINGL

Description: Determines the extent of crystal dissolution in the magma mingler. The crystal dissolution is based on Solid-Liquid-Equilibria. The calculation follows the algorithm of Starzak (2015 and 2016a).

Variables:

1. t_c – Input variable: Temperature of magma (stream MAGA) in °C.
2. S – Calculated variable: Pure sucrose solubility. It is a function of t_c.

3. $(SW)^{sat}_{pure}$ – Calculated variable: Sucrose-to-water ratio (another way of expressing solubility). Also just a function of t_c.
4. Non-sucrose in MAGA – Input variable: Flow rate of non-sucrose in MAGA stream (kg/h).
5. Water in MAGA – Input variable: Mass flow rate of water in MAGA stream (kg/h).
6. NSW^{MAGA} – Calculated variable: (Non-sucrose)-to-water ratio in the MAGA stream. Function of Non-sucrose in MAGA and Water in MAGA.
7. Sat. Coeff. (SC) – Calculated variable: The saturation coefficient of the MAGA stream. Function of A, B0, B1, C and NSW^{MAGA}.
8. Sucrose in MAGA – Input variable: Flow rate of sucrose in MAGA stream (kg/h).
9. $(SW^{MAGA})_{impure}$ – Calculated variable: Sucrose-to-water ratio in an impure solution (MAGA). Function of Sucrose in MAGA and Water in MAGA.
10. $(SW^{MAGA})^{sat}_{impure}$ – Calculated variable: The sucrose-to-water ratio for a saturated impure solution. Function of Sat. Coeff. (SC) and $(SW)^{sat}_{pure}$.
11. SS actual – Calculated variable: The current supersaturation (before crystallisation has occurred. Function of $(SW^{MAGA})_{impure}$ and $(SW^{MAGA})^{sat}_{impure}$.
12. Crystal in – Input variable: Flow rate of crystal in MAGA stream (kg/h).
13. Sucrose required – Calculated variable: The amount of sucrose which would be left in the massecuite to give the required supersaturation (SS required). Function of SS actual, SS required and Sucrose in MAGA.
14. Crystal conversion – Output variable: The fraction of crystal in MAGA which is dissolved. Sets the conversion of block CX-MING.

Constants:

- A, B0, B1 and C: Constants in the Saturation coefficient (SC) equation.

Parameter:

1. SS required – Exit supersaturation of MAGA1 stream. Value is 1.

Calculation:

Pure sucrose solubility:

$$S = 64.447 + 0.08222t_c + 1.6169 \times 10^{-3}t_c^2 - 1.559 \times 10^{-6}t_c^3 - 4.63 \times 10^{-8}t_c^4$$

Sucrose-to-water ratio (pure solution):

$$(SW)^{sat}_{pure} = \frac{S}{100 - S}$$

(Non-sucrose)-to-water ratio:

$$\mathrm{NSW}^{\mathrm{MAGA}} = \frac{\text{Non} - \text{sucrose in MAGA}}{\text{Water in MAGA}}$$

Saturation coefficient:

$$\text{Sat. Coeff. (SC)} = A \times \mathrm{NSW}^{\mathrm{MAGA}} + B0 - B1 \times t_c$$
$$+(1 - B0 + B1 \times t_c)\exp(-C \times \mathrm{NSW}^{\mathrm{MAGA}})$$

Sucrose-to-water ratio (impure solution):

$$(\mathrm{SW}^{\mathrm{MAGA}})_{\mathrm{impure}} = \frac{\text{Sucrose in MAGA}}{\text{Water in MAGA}}$$

Sucrose-to-water ratio (saturated impure solution):

$$(\mathrm{SW}^{\mathrm{MAGA}})_{\mathrm{impure}}^{\mathrm{sat}} = (\mathrm{SW})_{\mathrm{pure}}^{\mathrm{sat}} \times \text{Sat. Coeff. (SC)}$$

Supersaturation:

$$\text{SS actual} = \frac{(\mathrm{SW}^{\mathrm{MAGA}})_{\mathrm{impure}}}{(\mathrm{SW}^{\mathrm{MAGA}})_{\mathrm{impure}}^{\mathrm{sat}}}$$

Sucrose required to meet specified exit supersaturation condition:

$$\text{Sucrose required} = \frac{\text{SS required}}{\text{SS actual}} \times \text{Sucrose in MAGA}$$

Crystal conversion:

$$\text{Crystal conversion} = \frac{\text{Sucrose required} - \text{Sucrose in MAGA}}{\text{Crystal in}}$$

B.5 Drying module

<u>EXSS-CON</u>	<u>SUAH-MOI</u>	<u>SUAD-MOI</u>	<u>A-CX</u>
Exhaust steam flow rate to dry air heater	Moisture drying in heating section	Moisture absorption in cooling section	Crystallisation in dryer

Figure B.4 Drying module calculator blocks

B.5.1 Exhaust steam flow rate to dry air heater

Aspen Plus® block: EXSS-CON

Description: Manipulates the flow rate of exhaust steam to the dry air heater (AIR-HEAT) in order for the air to reach a specified temperature (80 °C).

Variables:

1. FLOWDAI1 – Input variable: Flow rate of air stream into the dry air heater (stream DAI1) in kg/h.
2. ENTHDAI1 – Input variable: Enthalpy of air stream DAI1 in kJ/kg.
3. ENTHDAH – Input variable: Enthalpy of air stream DAI1 at 80 °C in kJ/kg.
4. ENTHEXSS – Input variable: Enthalpy of exhaust steam (stream EXSS) to dry air heater in kJ/kg.
5. ENTHEXCS – Input variable: Enthalpy of condensate (stream EXCS) from dry air heater in kJ/kg.
6. FLOWEXSS – Output variable: Split of block SB2-SPLT in boiler flowsheet. This sets the flow rate of steam (stream EXSS) to the dry air heater.

Calculation:

$$\text{FLOWEXSS} = \frac{(\text{ENTHDAH} - \text{ENTHDAI1}) \times \text{FLOWDAI1}}{\text{ENTHEXSS} - \text{ENTHEXCS}}$$

B.5.2 Moisture drying in heating section of dryer

Aspen Plus® block: SUAH-MOI

Description: Based on a specified moisture content in the hot sugar leaving the dryer heating section the split fraction of water (in block DRYER) is calculated.

Variables:

1. FLOWIN – Input variable: Flow rate of hot sugar in the heating section of the dryer (stream SUAH) in kg/h.
2. WATERIN – Input variable: Flow rate of water in stream SUAH in kg/h.
3. WATER – Output variable: Separation coefficient in block DRYER of water to the hot sugar stream.

Parameter:

- MOISTURE – Parameter 508: Moisture content (water fraction) of hot sugar leaving the heating section of the dryer. Value is 6.5×10^{-4} on wet basis.

Calculation:

$$\text{WATER} = \frac{\text{MOISTURE} \times \left(\frac{\text{FLOWIN}-\text{WATERIN}}{1-\text{MOISTURE}}\right)}{\text{WATERIN}}$$

B.5.3 Moisture absorption in cooling section

Aspen Plus® block: SUAD-MOI

Description: Based on a specified moisture content in the cold sugar leaving the dryer cooling section the split fraction of water (in block MOIS2-SP) is calculated.

Variables:

1. FLOW– Input variable: Flow rate of cold sugar in the cooling section of the dryer (stream SUAD-1) in kg/h.
2. WATER1 – Input variable: Flow rate of water in stream SUAD-1 in kg/h.
3. WATERIN – Input variable: Flow rate of water in the cooling air (stream DAO-2) in kg/h.
4. WATSPLIT – Output variable: Separation coefficient in block MOIS2-SP of water to the cold sugar stream.

Parameter:

- MOIST – Parameter 508: Moisture content (water fraction) of cold sugar leaving the cooling section of the dryer. Value is 7.8821×10^{-4} on wet basis.

Calculation:

$$\text{WATSPLIT} = \frac{\text{MOIST} \times \left(\frac{\text{FLOW}-\text{WATER1}}{1-\text{MOIST}}\right) - \text{WATER1}}{\text{WATERIN}}$$

B.5.4 Crystallisation in dryer

Aspen Plus® block: SUAH-MOI

Description: Determines the extent of crystallisation in the dryer. The crystallisation of sucrose is based on Solid-Liquid-Equilibria. The calculation follows the algorithm of Starzak (2015 and 2016a).

Variables:

1. S – Calculated variable: Pure sucrose solubility. It is a function of t_c.
2. $(SW)^{sat}_{pure}$ – Calculated variable: Sucrose-to-water ratio (another way of expressing solubility). Also just a function of t_c.

3. Non-sucrose in SUAD – Input variable: Flow rate of non-sucrose in SUAD stream (kg/h).
4. Water in SUAD – Input variable: Mass flow rate of water in SUAD stream (kg/h).
5. NSW^{SUAD} – Calculated variable: (Non-sucrose)-to-water ratio in the SUAD stream. Function of Non-sucrose in SUAD and Water in SUAD.
6. Sat. Coeff. (SC) – Calculated variable: The saturation coefficient of the SUAD stream. Function of A, B0, B1, C and NSW^{SUAD}.
7. Sucrose in SUAD – Input variable: Flow rate of sucrose in SUAD stream (kg/h).
8. $(SW^{SUAD})_{impure}$ – Calculated variable: Sucrose-to-water ratio in an impure solution (SUAD). Function of Sucrose in SUAD and Water in SUAD.
9. $(SW^{SUAD})^{sat}{}_{impure}$ – Calculated variable: The sucrose-to-water ratio for a saturated impure solution. Function of Sat. Coeff. (SC) and $(SW)^{sat}{}_{pure}$.
10. SS actual – Calculated variable: The current supersaturation (before crystallisation has occurred. Function of $(SW^{SUAD})_{impure}$ and $(SW^{SUAD})^{sat}{}_{impure}$.
11. Sucrose required – Calculated variable: The amount of sucrose which would be left in the massecuite to give the required supersaturation (SS required). Function of SS actual, SS required and Sucrose in SUAD.
12. Sucrose conversion – Output variable: The fraction of sucrose in SUAD which is crystallised. Sets the conversion of block CX-DRYER.

Constants:

- A, B0, B1 and C: Constants in the saturation coefficient (SC) equation.

Parameters:

1. t_c: Exit temperature of raw sugar from the dryer module. Value is 35.92 °C.
2. SS required – Exit supersaturation of SUAD stream. Value is 1.

Calculations:

Pure sucrose solubility:

$$S = 64.447 + 0.08222t_c + 1.6169 \times 10^{-3}t_c^2 - 1.559 \times 10^{-6}t_c^3 - 4.63 \times 10^{-8}t_c^4$$

Sucrose-to-water ratio (pure solution):

$$(SW)_{pure}^{sat} = \frac{S}{100 - S}$$

(Non-sucrose)-to-water ratio:

$$\mathrm{NSW}^{\mathrm{SUAD}} = \frac{\text{Non} - \text{sucrose in SUAD}}{\text{Water in SUAD}}$$

Saturation coefficient:

$$\text{Sat. Coeff. (SC)} = A \times \mathrm{NSW}^{\mathrm{SUAD}} + B0 - B1 \times t_c + (1 - B0 + B1 \times t_c)\exp(-C \times \mathrm{NSW}^{\mathrm{SUAD}})$$

Sucrose-to-water ratio (impure solution):

$$(\mathrm{SW}^{\mathrm{SUAD}})_{\mathrm{impure}} = \frac{\text{Sucrose in SUAD}}{\text{Water in SUAD}}$$

Sucrose-to-water ratio (saturated impure solution):

$$(\mathrm{SW}^{\mathrm{SUAD}})^{\mathrm{sat}}_{\mathrm{impure}} = (\mathrm{SW})^{\mathrm{sat}}_{\mathrm{pure}} \times \text{Sat. Coeff. (SC)}$$

Supersaturation:

$$\text{SS actual} = \frac{(\mathrm{SW}^{\mathrm{SUAD}})_{\mathrm{impure}}}{(\mathrm{SW}^{\mathrm{SUAD}})^{\mathrm{sat}}_{\mathrm{impure}}}$$

Sucrose required to meet specified exit supersaturation condition:

$$\text{Sucrose required} = \frac{\text{SS required}}{\text{SS actual}} \times \text{Sucrose in SUAD}$$

Sucrose conversion:

$$\text{Sucrose conversion} = \frac{\text{Sucrose in SUAD} - \text{Sucrose required}}{\text{Sucrose in SUAD}}$$

B.6 Boiler module

BAG2BOIL	EXSD-FLO	SBL-FLOW	WBB-FLOW	WBM-FLOW
Bagasse flow rate to boiler	Exhaust steam flow rate to deaerator	Live steam losses	Blowdown flow rate	Make up water flow rate to boiler

Figure B.5 Boiler module calculator blocks

B.6.1 Bagasse required by boiler

Aspen Plus® block: BAG2BOIL

Description: Calculates the required flow rate of bagasse to the boilers based on a pre-specified amount of bagasse which is needed to produce 1 kg of steam.

Variables:

1. FLOWWB – Input variable: Mass flow rate of boiler feed water (stream WB) in kg/h.
2. SPLITBAG – Output variable: Split fraction of block BAG-SPLT in boiler flowsheet. This sets the amount of bagasse which goes to the boilers.

Parameter:

- MUBAGW – Parameter 604: Amount of bagasse required to produce 1 kg of steam (kg bagasse/ kg steam). Value is 0.45.

Calculation: SPLITBAG = MUBAGW × FLOWWB

B.6.2 Exhaust steam flow rate to deaerator

Aspen Plus® block: EXSD-FLO

Description: Sets the flow rate of exhaust steam to the deaerator in the boiler module (deaerator is not modelled). This flow rate is proportional to the flow rate of high pressure (live) steam used.

Variables:

1. FSBF – Input variable: Flow rate of live steam (stream SBF) after blowdown separation in kg/h.
2. FEXSD – Output variable: Split fraction of block SB2-SPLT in boiler flowsheet. This sets the amount of exhaust steam which goes to the deaerator.

Parameter:

- DEAERATE – Parameter 605: Deaerator steam demand as a fraction of stream SBF flow. Value is 0.02.

Calculation: FEXSD = DEAERATE × FSBF

B.6.3 High pressure (live) steam losses

Aspen Plus® block: SBL-FLOW

Description: An empirical equation (Rein, 2007) is used to account for steam losses due to leaks, occasional venting, cleaning, etc. The steam loss stream is vented in the model.

Variables:

1. FCANE – Input variable: Flow rate of cane in tonne/h.
2. FSBL – Output variable: Split fraction of block SB1-SPLT in boiler flowsheet. This sets the amount of live steam which is lost (tonne/h).

Calculation: FSBL = 0.1 $\times \mathrm{FCANE}^{0.67}$

B.6.4 Blowdown purge flow rate

Aspen Plus® block: WBB-FLOW

Description: Sets the flow rate of boiler water blowdown. (Comment: In the MATLAB™ model the blowdown is saturated water at 31 bara, however in Aspen Plus® it is separated out as a portion of live steam)

Variables:

1. FWB – Input variable: Flow rate of boiler feed water (stream WB) in kg/h.
2. WBBFLOW – Output variable: Split fraction of block WBB-SPLT in boiler flowsheet. This sets the amount lost as blowdown.

Parameter:

- WBB – Parameter 601: Fraction of boiler feed water which is lost as blowdown. Value is 0.002.

Calculation: WBBFLOW = WBB × FWB

B.6.5 Make-up water flow rate to boiler

Aspen Plus® block: WBM-FLOW

Description: Sets the flow rate of make-up water to the boiler. (WBM = WBB + SBL + EXSD)

Variables:

1. FCANE – Input variable: Flow rate of cane in t/h.
2. FSBF – Input variable: Flow rate of live steam (stream SBF) after blowdown separation in kg/h.
3. FWB – Input variable: Flow rate of boiler feed water (stream WB) in kg/h.
4. FWBM – Output variable: Sets the flow rate of make-up water to the boiler (stream WBM) in kg/h.

Parameters:

1. WBB – Parameter 601: Fraction of boiler feed water which is lost as blowdown. Value is 0.002.
2. DEAERATE – Parameter 605: Deaerator steam demand as a fraction of stream SBF flow. Value is 0.02.

Calculation: $FWBM = 1000 \times 0.1 \times FCANE^{0.67} + DEAERATE \times FSBF + WBB \times FWB$

APPENDIX C: ASPEN PLUS® DESIGN SPECIFICATIONS

Design specifications s are used in the Aspen Plus® model, these are converged using the secant method. A certain variable is specified (e.g. dry solids of exit stream from pan 'A') and then a different variable is manipulated (e.g. steam flow rate to pan 'A') in order to meet the specification.

C.1 Brix control of syrup

Aspen Plus® design spec: DS-CL5

Description: Manipulates the flow rate of steam (V3) to the 4th effect evaporator (stream S3P) in order to maintain a brix of 64.8 % in the syrup from the 5th effect (parameter 318).

Specified variable:

- BRIXL5 – Dependent variable: Mass fraction of water in the syrup from the last effect (stream L5-2). Specified as 0.352 with a tolerance of 0.001. Brix is 1 – (mass fraction of water), thus equal to 0.648.

Manipulated variable:

- ❖ S3P – Independent variable: Split of block V-DIST3 in evaporation flowsheet. This sets the flow rate of steam to the 4th effect evaporator. The manipulated variable may be varied between 20000 and 29000 kg/h.

C.2 Brix control of massecuite from 'A' pans

Aspen Plus® design spec: DS-PANA

Description: Manipulates the flow rate of steam (V1) to the 'A' pans (stream SKA) in order to maintain a brix of 91.69 % in the massecuite from the 'A' pans (parameter 407).

Specified variable:

- BRIXPANA – Dependent variable: Mass fraction of water in the massecuite from the 'A' pans (stream PANA). Specified as 0.083 with a tolerance of 0.001. Brix is 1 – (mass fraction of water), thus equal to 0.917.

Manipulated variable:

- ❖ SKA – Independent variable: Split of block V-DIST1 in the evaporation module. This sets the flow rate of steam to the 'A' pans. The manipulated variable may be varied between 19000 and 21000 kg/h.

C.3 Brix control of massecuite from 'B' pans

Aspen Plus® design spec: DS-PANB

Description: Manipulates the flow rate of steam (V1) to the 'B' pans (stream SKB) in order to maintain a brix of 92.85 % in the massecuite from the 'B' pans (parameter 411).

Specified variable:

- BRIXPANB – Dependent variable: Mass fraction of water in the massecuite from the 'B' pans (stream PANB). Specified as 0.0715 with a tolerance of 0.001. Brix is 1 – (mass fraction of water), thus equal to 0.9285.

Manipulated variable:

- ❖ SKB – Independent variable: Split of block V-DIST1 in evaporation flowsheet. This sets the flow rate of steam to the 'B' pans. The manipulated variable may be varied between 5000 and 7000 kg/h.

C.4 Brix control of massecuite from 'C' pans

Aspen Plus® design spec: DS-PANC

Description: Manipulates the flow rate of steam (V2) to the 'C' pans (stream SKC) in order to maintain a brix of 94.07 % in the massecuite from the 'C' pans (parameter 415).

Specified variable:

- BRIXPANC – Dependent variable: Mass fraction of water in the massecuite from the 'C' pans (stream PANC). Specified as 0.0593 with a tolerance of 0.002. Brix is 1 – (mass fraction of water), thus equal to 0.9407.

Manipulated variable:

- ❖ SKC – Independent variable: Split of block V-DIST2 in evaporation flowsheet. This sets the flow rate of steam to the 'C' pans. The manipulated variable may be varied between 2000 and 3000 kg/h.

C.5 Brix control of magma

Aspen Plus® design spec: DS-MINGL

Description: Manipulates the flow rate of syrup (stream SYRM) to the magma mingler in order to maintain a water percentage of 7.93 % in the magma stream (parameter 456).

Specified variable:

- MINGDS – Dependent variable: Mass fraction of water in the magma (stream MAGA). Specified as 0.0793 with a tolerance of 0.0001.

Manipulated variable:

❖ SYRM – Independent variable: Split of block SYRSPLIT in crystallisation flowsheet. This sets the flow rate of syrup to the mingler. The manipulated variable may be varied between 1000 and 2000 kg/h.

C.6 Temperature control of sugar from the dryer

Aspen Plus® design spec: DS-TSUAD

Description: Manipulates the flow rate of dry air into the drying module (stream DAI) in order to maintain a temperature of 35 °C (parameter 509) in the sugar from the cooling section of the dryer (stream SUAD).

Specified variable:

- TSUAD – Dependent variable: Temperature of sugar from the dryer (stream SUAD). Specified as 35 °C with a tolerance of 0.01 °C.

Manipulated variable:

❖ DAI – Independent variable: Mass flow rate of dry air into the drying module (stream DAI). The manipulated variable may be varied between 140000 and 160000 kg/h.

APPENDIX D: ASPEN PLUS® MODEL PROCESS PARAMETERS

The process parameters of the Aspen Plus® model are listed in table D.1. Some process parameters which were used in the setup of the MATLAB™ model were omitted in the Aspen Plus® model (e.g. crystal loss ratios in centrifuges) due to the model being simplified in places. Other process parameters were modified (values shown in red) due to the different means of application between MATLAB™ and Aspen Plus® (e.g. heat losses), however the end results match as closely as possible.

Table D.1 Process parameters in Aspen Plus® model

Parameter number	Description	Value
001	Cane throughput, t/h	244.2
002	Cane sucrose, %	14.17
003	Cane pol, %	13.86
004	Cane brix, %	16.41
005	Cane fibre, %	15.06
006	Feed pressure of cane, bara	1.013
007	Feed temperature of cane, °C	27
101	Temperature of HP steam, °C	390
102	Pressure of HP steam, bara	31
103	Steam usage by cane knives, t/t cane	0.0207
104	Exhaust steam from cane knives - pressure, bara	2
105	Exhaust steam from cane knives - temperature, °C	121
106	Steam usage by cane shredder, t/t cane	0.0621
107	Exhaust steam from cane shredder - pressure, abs bar	2
108	Exhaust steam from cane shredder - temperature, °C	121
109	Steam usage by drying mills turbine, t/t megasse fibre	0.0639
110	Exhaust steam from drying mills turbine - pressure, bara	2
111	Exhaust steam from drying mills turbine - temperature, °C	121
112	Consumption of steam injected to diffuser, t/t cane	0.01715
113	Consumption of steam used in diffuser heaters, t/t cane	0.03541
114	Imbibition water usage, t/t megasse fibre	2.954
115	Pressure of recycled press water, bara	2.068
116	Press water temperature after heating, °C	69.36
117	Diffuser extraction coefficients - water	0.4690
118	Diffuser extraction coefficients - sucrose	0.8089
119	Diffuser extraction coefficients - non-sucrose	0.8714
120	Diffuser extraction coefficients - fibre	0.02945
121	Water content in bagasse, mass fraction	0.5096
122	Bagasse press separation coefficients - sucrose	0.1421

123	Bagasse press separation coefficients - non-sucrose	0.1512
124	Bagasse press separation coefficients - fibre	1
125	Heat loss in diffuser - draft juice, °C	3.2
126	Heat loss in dewatering mills - press water, °C	1
201	Lime milk pressure, bara	1.013
202	Lime milk temperature, °C	20
203	Mass fraction of lime in lime milk	0.1
204	Mass fraction of lime in mixed juice	0.00288
205	Mixed juice temperature after primary heating, °C	77.23
206	Mixed juice temperature after secondary heating, °C	93.85
207	Mixed juice temperature after tertiary heating, °C	103.9
208	Mixed juice pressure to secondary heater, bara	3.5
209	Mixed juice flash pressure, bara	1.013
210	Clear juice exit pressure, bara	2.392
211	Clarifier separation coefficients - sucrose	0.1154
212	Clarifier separation coefficients - non-sucrose	0.1573
213	Clarifier separation coefficients - fibre	1
214	Clarifier separation coefficients - lime	1
215	Mass fraction of water in mud stream	0.8041
216	Bagasse to mud-bagasse blender, t/t mud	0.02743
217	Mass fraction of water in filter cake	0.7000
218	Vacuum filter separation coefficients - sucrose	0.05147
219	Vacuum filter separation coefficients - non-sucrose	0.1789
220	Vacuum filter separation coefficients - fibre	1
221	Vacuum filter separation coefficients - lime	0.8267
222	Wash water to vacuum filter, t/t insolubles in cake	1.983
226	Heat loss in clarifier - mud and clear juice streams, °C	0.5
301	Number of effects	5
302	Pressure of exhaust steam from turbo alternator, bara	2
304	Evaporator pressure distribution - 1 effect , bara	1.6
305	Evaporator pressure distribution - 2 effect , bara	1.25
306	Evaporator pressure distribution - 3 effect , bara	0.6
307	Evaporator pressure distribution - 4 effect , bara	0.4
308	Evaporator pressure distribution - 5 effect , bara	0.16
314	Hydraulic pressure losses, bara	0.02
315	Throttling valve steam pressure reduction, bara	0.15
316	Temperature of exhaust steam from turbo alternator, °C	121
317	Temperature of juice from preheater, °C	112.5
318	Final syrup DS (Dry Substance, % wt.)	64.81
319	Temperature of water in the barometric condenser well, °C	40
320	Pressure of water in the barometric condenser well, bara	1.013
331	Fractional evaporator heat loss - 1 effect	0.0058
332	Fractional evaporator heat loss - 2 effect	0.006
333	Fractional evaporator heat loss - 3 effect	0.001
334	Fractional evaporator heat loss - 4 effect	0.00385

335	Fractional evaporator heat loss - 5 effect	0.00225
342	Thick juice filter efficiencies - water	0.997
343	Thick juice filter efficiencies - sucrose	0.998
344	Thick juice filter efficiencies - non-sucrose	0.996
345	Exit condensate pump pressure, bara	3.081
346	Degree of droplet entrainment in the 1st effect, %	0.7735
347	Degree of sucrose inversion in the 1st effect, %	0.1245
400	Parameter d in b1 = b-d*T	0.00025
401	Parameter a in SC = a*NS/W+b1+(1-b1)*exp(-c*NS/W)	0.2281
402	Parameter b in SC = a*NS/W+b1+(1-b1)*exp(-c*NS/W)	0.03584
403	Parameter c in SC = a*NS/W+b1+(1-b1)*exp(-c*NS/W)	0.3289
404	Syrup distribution coefficient to B-pan, [-]	0
405	Syrup distribution coefficient to C_pan, [-]	0
406	Supersaturation of A-massecuite from A-pan, [-]	1.335
407	Dry substance of A-massecuite from A-pan, %	91.69
408	Temperature of A-massecuite from A-pan, °C	67.12
409	Fractional heat loss from A-pan	0.0448
410	Supersaturation of B-massecuite from B-pan, [-]	1.212
411	Dry substance of B-massecuite from B-pan, %	92.85
412	Temperature of B-massecuite from B-pan, °C	66.88
413	Fractional heat loss from B-pan	0.051
414	Supersaturation of C-massecuite from C-pan, [-]	1.130
415	Dry substance of C-massecuite from C-pan, %	94.07
416	Temperature of C-massecuite from C-pan, °C	66.91
417	Fractional heat loss from C-pan	0.08
418	Supersaturation of A-massecuite from A-crystallizer, [-]	1.349
419	Pressure of A-massecuite from A-crystallizer, bara	1.013
420	Temperature of A-massecuite from A-crystallizer, °C	56.23
421	Supersaturation of B-massecuite from B-crystallizer, [-]	1.050
422	Pressure of B-massecuite from B-crystallizer, bara	1.013
423	Temperature of B-massecuite from B-crystallizer, °C	49.86
424	Supersaturation of C-massecuite from C-crystallizer, [-]	1.089
426	Pressure of C-massecuite from C-crystallizer, bara	1.013
427	Temperature of C-massecuite from C-crystallizer, °C	45.22
432	Temperature of A-molasses, °C	60.2
433	Temperature of A-sugar, °C	56.2
439	Temperature of B-molasses, °C	50.66
440	Temperature of B-sugar, °C	49.9
446	Temperature of C-molasses, °C	56.1
447	Temperature of C-sugar, °C	45.3
448	Distribution coefficient of C-molasses to fermentation	0
456	Water percentage in magma from mingler, %	7.929
457	Brix of remelt stream, %	78.32
463	Degree of droplet entrainment in A-pan, %	0.4013
464	Degree of sucrose inversion in A-pan, %	0.1899

501	Pressure of exhaust steam to drying air heater, bara	2
502	Temperature of exhaust steam to drying air heater, °C	121
503	Drying air pressure, bara	1.013
504	Drying air inlet temperature, °C	26
505	Drying air inlet relative humidity, %	75
506	Hot drying air temperature, °C	80
507	Hot A-sugar exit temperature, °C	60
508	Hot A-sugar moisture content, % (wet basis)	0.065
509	Cold A-sugar exit temperature, °C	35
510	Cold A-sugar moisture content, % (wet basis)	0.07882
601	Boiler water blowdown fraction	0.002
602	Pressure of boiler make-up water, bara	1.013
603	Temperature of boiler make-up water, °C	25
604	Bagasse-to-boiler feed water ratio	0.45
605	Deaerator steam demand fraction	0.02
608	Cooling tower total pressure, bara	1.013
609	Cooling tower temperature drop, °C	2.3
610	Cooling tower final water temperature, °C	25
611	Cooling tower final water pressure, bara	3

APPENDIX E: A3 FLOWSHEETS OF EVAPORATION AND CRYSTALLISATION MODULES IN ASPEN PLUS®

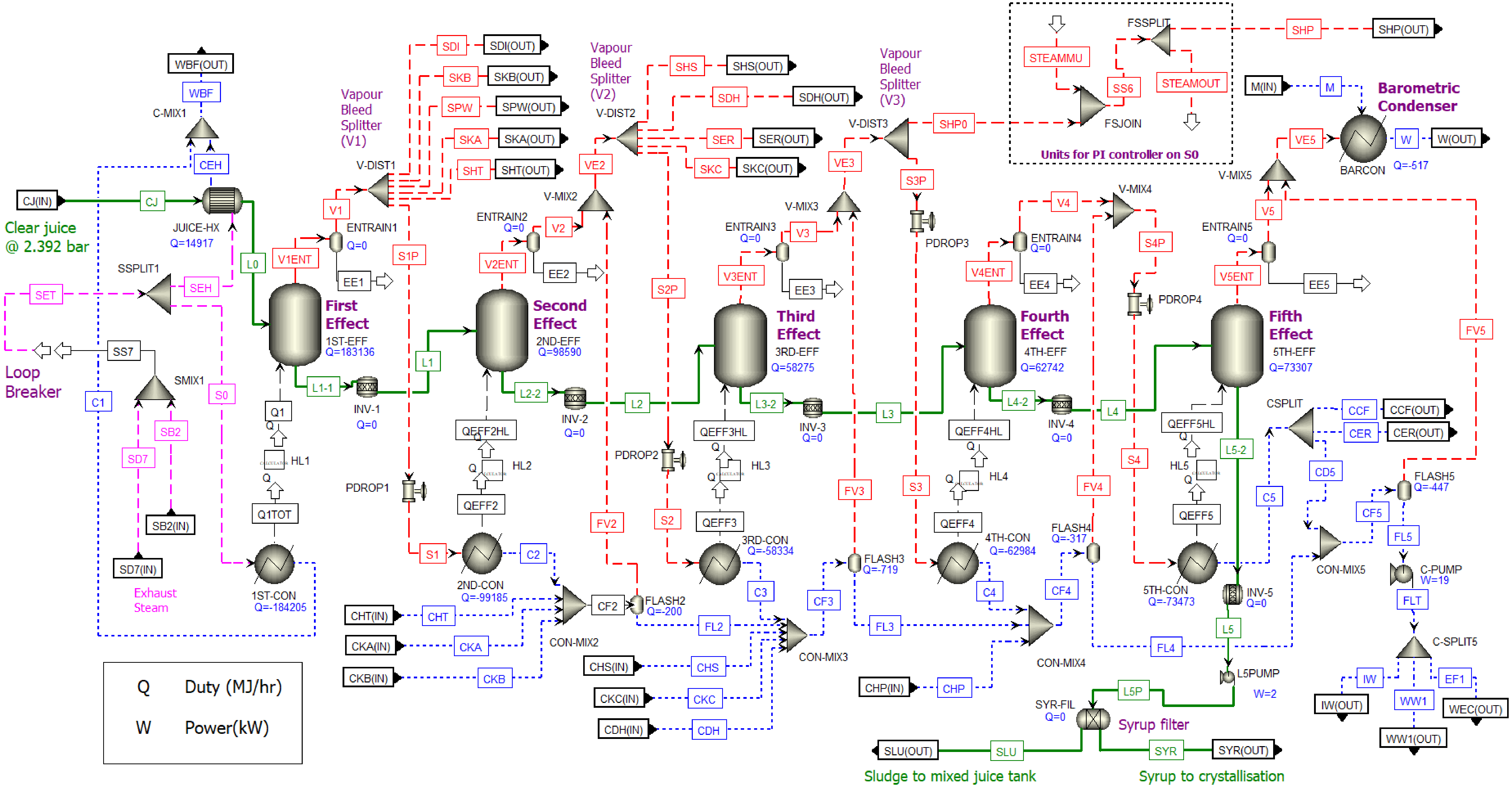

Figure E.1 Enlarged flowsheet of evaporation module in Aspen Plus®

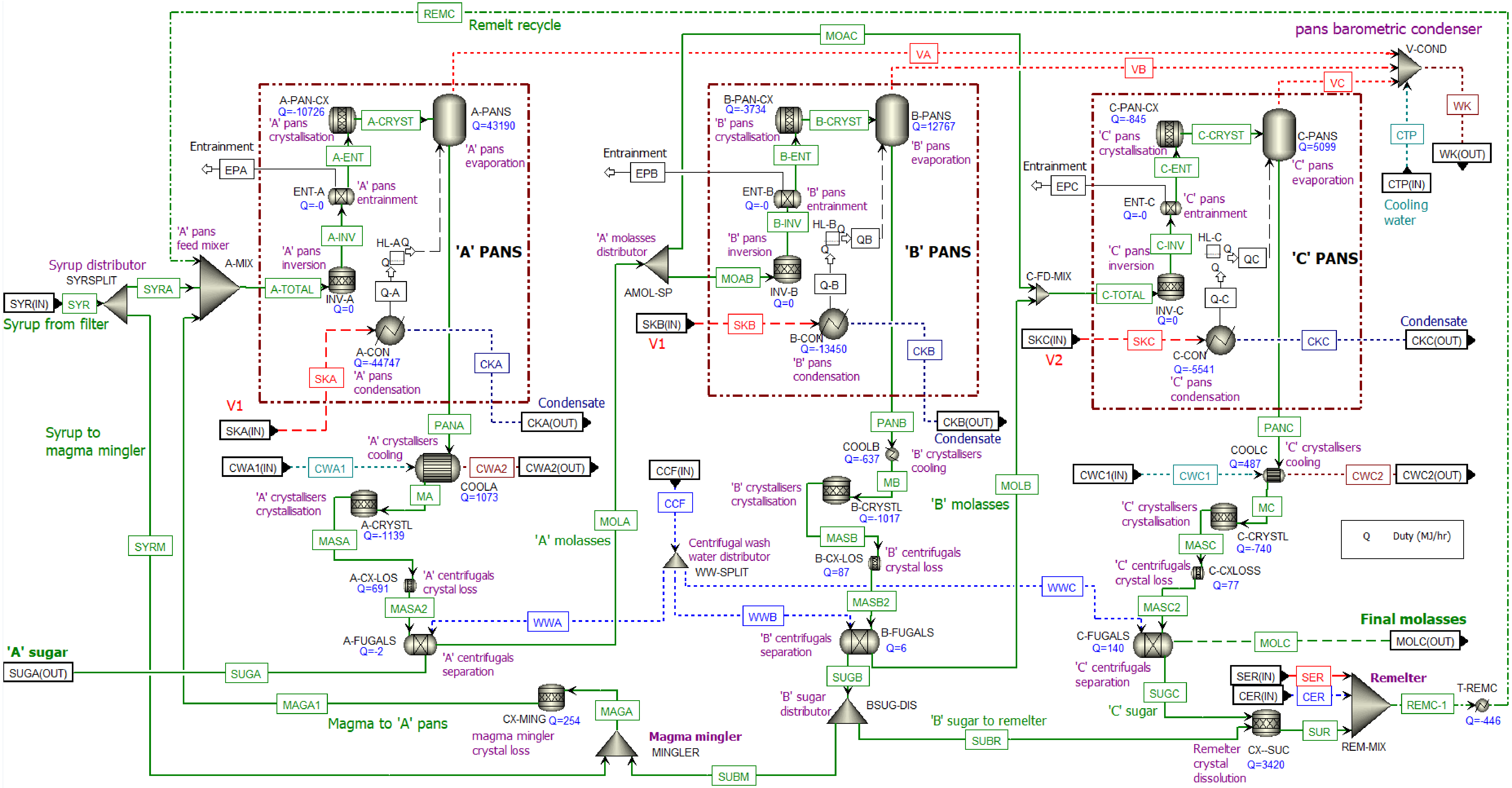

Figure E.2 Enlarged flowsheet of crystallisation module in Aspen Plus®

APPENDIX F: STREAM TABLES FOR THE ASPEN PLUS® MODEL

The stream names as they appear on the Aspen Plus® model flowsheets are listed with a description of the streams.

F.1 Extraction module

Table F.1 Sugar streams in extraction module

Stream	Description
CANE	Sugar cane feed
CANE1	Chopped cane after cane knives
CANE2	Shredded cane after cane shredder
MEGA	Megasse before temperature correction
MEG	Megasse
DJDIFF	Draft juice from diffuser before heat losses
DJHOT	Draft juice before temperature correction
DJ	Draft juice
BAG	Bagasse
PW-2	Press water before heat losses
PW-1	Press water to pump after heat loss
PW	Press water

Table F.2 Steam/condensate streams in extraction module

Stream	Description
SPW	Steam injected to press water tank
SDI	Steam injected to diffuser
SDH	Steam to scalding juice heater
CDH	Condensate from scalding juice heater
IW	Imbibition water
SB1	High pressure (live) steam from boiler
SD1	High pressure steam to cane knives motor turbine
SD2	Low pressure steam from cane knives motor turbine

SD3	High pressure steam to cane shredder turbine
SD4	Low pressure steam from cane shredder turbine
SD5	High pressure steam to dewatering mills turbine
SD6	Low pressure steam from dewatering mills turbine
SD7	Mixture of low pressure steam from cane knives, cane shredders and dewatering mills (turbines)

F.2 Clarification module

Table F.3 Sugar streams in clarification module

Stream	Description
BGO	Bagacillo
CJ	Clear juice after pump
CJ-1	Clear juice after heat loss
CJ-2	Clear juice from clarifier
DJ	Draft juice
FC	Filter cake
FJ	Filtrate juice
FJF	Filtrate juice for further processing
FJR	Filtrate juice recycle
LIM	Milk of lime
MB	Mud-bagacillo mixture
MJ	Mixed juice
MJ1	Mixed juice after primary heater
MJ2	Mixed juice after secondary heater
MJ3	Mixed juice after tertiary heater
MJF	Mixed juice from flash
MJL	Mixed juice after limer
MJP	Mixed juice after pump
MUD	Mud after heat loss
MUD-1	Mud from clarifier
SLU	Sludge from syrup clarifier
VMJ	Vent from flash

Table F.4 Steam/condensate streams in clarification module

Stream	Description
CHP	Condensate from primary heater
CHS	Condensate from secondary heater
CHT	Condensate from tertiary heater
SHP	Steam to primary heater
SHS	Steam to secondary heater
SHT	Steam to tertiary heater
WW1	Wash water to filter

F.3 Evaporation module

Table F.5 Sugar streams in evaporation module

Stream	Description
CJ	Clear juice
L0	Feed to 1st effect evaporator
L1-1	Liquid from 1st effect evaporator before inversion
L1	Liquid from 1st effect evaporator after inversion
L2-2	Liquid from 2nd effect evaporator before inversion
L2	Liquid from 2nd effect evaporator after inversion
L3-2	Liquid from 3rd effect evaporator before inversion
L3	Liquid from 3rd effect evaporator after inversion
L4-2	Liquid from 4th effect evaporator before inversion
L4	Liquid from 4th effect evaporator after inversion
L5-2	Liquid from 5th effect evaporator before inversion
L5	Liquid from 5th effect evaporator after inversion
L5P	Liquid from 5th effect after pump
SLU	Sludge from syrup filter
SYR	Syrup to crystallisation module

Table F.6 Steam/condensate streams in evaporation module

Stream	Description
SD7	Exhaust steam from extraction module motor drives
SB2	Exhaust steam from turbo-alternator
SET	Total exhaust steam
CHT	Condensate from tertiary mixed juice heater
CKA	Condensate from 'A' pans
CKB	Condensate from 'B' pans
CDH	Condensate from scalding juice heater
CKC	Condensate from 'C' pans
CHS	Condensate from secondary mixed juice heater
CHP	Condensate from primary mixed juice heater
IW	Imbibition water
WW1	Wash water to vacuum filter
WEC	Waste water (effluent)
SEH	Exhaust steam to clarified juice preheater
S0	Exhaust steam to 1st effect evaporator
CEH	Condensate from clarified juice preheater
C1	Condensate from 1st effect evaporator
WBF	Mixture of condensates from preheater and 1st effect
V1ENT	Vapour with entrainment from 1st effect evaporator
V1	Vapour only from 1st effect
EE1	Liquid entrainment from 1st effect
SDI	Steam to diffuser
SKB	Steam to 'B' pans
SPW	Steam to press water tank
SKA	Steam to 'A' pans
SHT	Steam to tertiary mixed juice heater
S1P	Steam to 2nd effect evaporator before pressure drop
S1	Steam to 2nd effect evaporator after pressure drop
C2	Condensate from 2nd effect evaporator
CF2	Feed to 2nd effect condensate flash
FV2	Vapour from 2nd effect condensate flash

FL2	Liquid from 2^{nd} effect condensate flash
V2ENT	Vapour with entrainment from 2^{nd} effect evaporator
V2	Vapour only from 2^{nd} effect
EE2	Liquid entrainment from 2^{nd} effect
VE2	Mixture of vapour 2^{nd} effect evaporator and 2^{nd} effect condensate flash
SHS	Steam to secondary mixed juice heater
SDH	Steam to scalding juice heater
SER	Steam to remelter
SKC	Steam to 'C' pans
S2P	Steam to 3^{rd} effect evaporator before pressure drop
S2	Steam to 3^{rd} effect evaporator after pressure drop
C3	Condensate from 3^{rd} effect evaporator
V3ENT	Vapour with entrainment from 3^{rd} effect evaporator
EE3	Liquid entrainment from 3^{rd} effect
V3	Vapour only from 3^{rd} effect
VE3	Mixture of vapour 3^{rd} effect evaporator and 3^{rd} effect condensate flash
SHP0	Steam to primary mixed juice heater before PI controller
SHP	Steam to primary mixed juice heater
S3P	Steam to 4^{th} effect evaporator before pressure drop
S3	Steam to 4^{th} effect evaporator after pressure drop
CF3	Feed to 3^{rd} effect condensate flash
FV3	Vapour from 3^{rd} effect condensate flash
FL3	Liquid from 3^{rd} effect condensate flash
C4	Condensate from 4^{th} effect evaporator
V4ENT	Vapour with entrainment from 4^{th} effect evaporator
EE4	Liquid entrainment from 4^{th} effect
V4	Vapour only from 4^{th} effect
S4P	Steam to 5^{th} effect evaporator before pressure drop
S4	Steam to 5^{th} effect evaporator after pressure drop
CF4	Feed to 4^{th} effect condensate flash
FV4	Vapour from 4^{th} effect condensate flash
FL4	Liquid from 4^{th} effect condensate flash
V5ENT	Vapour with entrainment from 5^{th} effect evaporator
EE5	Liquid entrainment from 5^{th} effect

V5	Vapour only from 5th effect
VE5	Vapour to barometric condenser
M	Cooling water to barometric condenser
W	Cooling water from barometric condenser
C5	Condensate from 5th effect
CCF	Condensate to centrifuges
CER	Condensate to remelter
CD5	Condensate to condensate mixer
CF5	Feed to 5th effect condensate flash
FV5	Vapour from 5th effect condensate flash
FL5	Liquid from 5th effect condensate flash
FLT	Liquid from 5th effect condensate flash after condensate pump
STEAMMU	Fictitious steam in for PI controller
STEAMOUT	Fictitious steam out for PI controller, once solved it matches STEAMMU
SS6	Steam stream in PI controller
SS7	Loop breaker stream, same conditions as SET stream

F.4 Crystallisation module

Table F.7 Sugar streams in crystallisation module

Stream	Description
A-CRYST	Total 'A' pans feed after crystallisation
A-ENT	Total 'A' pans feed after entrainment
A-INV	Total feed to 'A' pans after inversion
A-TOTAL	Total feed to 'A' pans
B-CRYST	Feed to 'B' pans after crystallisation
B-ENT	Feed to 'B' pans after entrainment
B-INV	Feed to 'B' pans after inversion
C-CRYST	Feed to 'B' pans after crystallisation
C-ENT	Feed to 'B' pans after entrainment
C-INV	Feed to 'B' pans after inversion
C-TOTAL	Total feed to 'C' pans
EPA	Entrainment stream from 'A' pans

EPB	Entrainment stream from 'B' pans
EPC	Entrainment stream from 'C' pans
MA	'A' massecuite after cooling in crystalliser
MAGA	Magma to 'A' pans
MAGA-1	Magma before partial crystal dissolution
MASA	'A' massecuite after crystalliser
MASA2	'A' massecuite after crystal loss in centrifuges
MASB	'B' massecuite after crystalliser
MASB2	'B' massecuite after crystal loss in centrifuges
MASC	'C' massecuite after crystalliser
MASC2	'C' massecuite after crystal loss in centrifuges
MB	'B' massecuite after cooling in crystalliser
MC	'C' massecuite after cooling in crystalliser
MOAB	'A' molasses to 'B' pans
MOAC	'A' molasses to 'C' pans
MOLA	'A' molasses
MOLB	'B' molasses
MOLC	'C' molasses
PANA	'A' massecuite from ‘A’ pans
PANB	'B' massecuite from ‘B’ pans
PANC	'C' massecuite from ‘C’ pans
REMC	Remelt to 'A' pans
REMC-1	Remelt before temperature correction
SUBM	'B' sugar to magma mingler
SUBR	'B' sugar to remelter
SUGA	'A' sugar to drying
SUGB	'B' sugar
SUGC	'C' sugar
SUR	Sucrose to remelter (from 'B' and 'C' sugar)
SYR	Syrup from evaporation module
SYRA	Syrup to 'A' pans
SYRM	Syrup to magma mingler

Table F.8 Steam/condensate streams in crystallisation module

Stream	Description
CCF	Condensate to centrifuges
CER	Condensate to remelter
CKA	Condensate from 'A' pans
CKB	Condensate from 'B' pans
CKC	Condensate from 'C' pans
CTP	Cooling water to pans barometric condenser
CWA1	Cooling water to 'A' crystalliser
CWA2	Cooling water from 'A' crystalliser
CWC1	Cooling water to 'C' crystalliser
CWC2	Cooling water from 'C' crystalliser
SER	Steam to remelter
SKA	Steam to 'A' pans
SKB	Steam to 'B' pans
SKC	Steam to 'C' pans
VA	Vapour from 'A' pans
VB	Vapour from 'B' pans
VC	Vapour from 'C' pans
WK	Warm cooling water from pans barometric condenser
WWA	'A' centrifuge wash water
WWB	'B' centrifuge wash water
WWC	'C' centrifuge wash water

F.5 Drying module

Table F.9 All streams in drying module

Stream	Description
DAH	Hot dry air
DAI	Dry air inlet
DAI1	Dry air to heater
DAI2	Dry air to sugar cooler
DAO	Total moist air outlet
DAO1	Moist air from heating section of dryer
DAO-1	Air outlet from heating section of dryer
DAO2	Air outlet from cooling section of dryer
DAO-2	Air outlet from sugar cooler
EXCS	Condensate from dry air heater
EXSS	Steam to dry air heater
MOIST	Water evaporated in drying process
MOIST2	Moisture absorbed by sugar in cooling section of dryer
SUA	Dry raw sugar – final product of raw sugar mill
SUAD	Sugar from cooling section of dryer
SUAD-1	Sugar from sugar cooler
SUAH	Sugar from dryer heat exchanger
SUAH2	Sugar from heating section of dryer
SUGA	'A' sugar from crystallisation module

F.6 Boiler module

Table F.10 All streams in boiler module

Stream	Description
BAG	Bagasse from extraction module
BAG2B	Bagasse needed by boiler
BAGOUT	Excess bagasse
EXCS	Condensate from dry air heater in dryer module
EXSD	Exhaust steam to deaerator

EXSS	Exhaust steam to dry air heater
HPS	High pressure steam
SB0	High pressure steam to turbo-alternator
SB00	Exhaust steam from turbo-alternator
SB1	High pressure steam to motor drives in extraction module
SB2	Exhaust steam to evaporator module
SBF	High pressure steam to splitter
SBL	High pressure steam losses
WB	Total boiler feed water
WBB	Boiler blowdown water (steam is modelled)
WBF	Boiler feed water from evaporation module
WBM	Boiler feed make-up water

F.7 Cooling tower module

Table F.11 All streams in cooling tower module

Stream	Description
CTL	Liquid stream from cooling tower
CTP	Cooling water to pans barometric condenser
CTV	Vapour stream from cooling tower
CT1	Fictitious stream to compensate for insufficient evaporation in cooling tower
CTW	Cold cooling water
CTW-1	Cooling water before sucrose is purged
CWA1	Cooling water to 'A' crystalliser
CWA2	Cooling water from 'A' crystalliser
CWC1	Cooling water to 'C' crystalliser
CWC2	Cooling water from 'C' crystalliser
CWW	Warm cooling water to cooling tower
CWW1	Total warm cooling water
EF2	Effluent
M	Cooling water to evaporators barometric condenser
SUC	Sucrose purged from cooling water
W	Warm cooling water from evaporators barometric condenser
WK	Warm cooling water from pans barometric condenser