

Exergy, techno-economic and exergoeconomic analyses for improving energy efficiency of a typical sugar mill and designing integrated biorefineries

By

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March 2020

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Abstract

The sugar industry is energy-intensive, consuming about 350–600 kg of steam and 25–32 kWh electricity per ton of sugarcane processed into raw crystalline sugar. Though mostly energy-self-sufficient, improving its energy efficiency is necessary to produce sugar more cost-effectively. Besides, the decreasing trend and fluctuations in the world sugar prices necessitate product diversity to ensure the economic sustainability of the industry. Therefore, improving the current energy status of the sugar industry will make sugarcane resources available for further valorisation. This study aimed to improve the energy efficiency of a typical South African sugar mill towards its economic sustainability and competitiveness.

The first objective (objective 1) towards achieving this aim was to identify the locations, magnitudes, and causes of inefficiencies in a typical South African sugar mill through exergy analysis. This analysis was based on rigorous mass and energy balances calculated from an Aspen Plus® simulation of a typical 250 ton per hour sugar mill. The cogeneration system had the highest exergy destruction (90 582 kW) representing 86% of the total sugar mill irreversibility. However, with the lowest exergetic efficiency of 9.6%, the crystallization unit recorded the most inefficient use of energy due to the process complexity.

Following the exergy results, objective 2 was to select energy-efficient technologies to improve the sugar mill efficiency while objective 3 involved assessing the economic feasibility of integrating them into the mill. Organic Rankine cycle (ORC) and absorption heat pump (AHP) technologies were selected, which improved the cogeneration exergetic efficiency by 1.7% and minimized overall system irreversibility by 0.14%, saving 0.83% on total bagasse for valorisation, respectively. Though only marginal improvements were achieved, both ORC and AHP integrations were economically feasible and could be optimized to achieve better energy improvements.

Furthermore, promising biorefinery products; succinic acid (SA) and short-chain fructooligosaccharides (scFOS) were integrated for the economic competitiveness of the industry in objective 4. Based on the exergy results, the biorefineries were developed to utilise A-molasses for the production of SA and scFOS in seven different scenarios, which were all highly profitable with internal rates of return (IRRs) between 24 and 62% compared to the minimum required IRR of 9.7%, due to the integration benefits. Moreover, co-utilization of C-molasses and lignocellulose residue as first- and second-generation (1G-2G) feedstocks for the

production of SA was also considered to fully valorise the sugarcane plant considering the current crystallization scheme.

Objective 5 applied an aggregated system exergoeconomic methodology to assess the holistic performance of the biorefineries and to identify the most cost-effective one. With the lowest cost rate of 1 029 US\$/h and exergoeconomic factor of 0.56, the scFOS powder scenario (S-F_P) showed a good balance between the irreversibility- and investment-related costs and was considered the most cost-effective biorefinery for integration into the sugar mill. Overall, this study presented a broad spectrum of solutions to the energy and economic challenges of the sugar industry to be explored further for implementation, using exergy/exergoeconomic methodology as better design tools than conventional energy and economic analysis.

Opsomming

Die suikerindustrie is energie-intensief, met die verbruik van omtrent 350–600 kg stoom en 25–32 kWh elektrisiteit per ton suikerriet geprosesseer in rou kristalvormige suiker. Al is dit meestal energieselponderhoudend, is dit nodig om die energiedoeltreffendheid te verbeter om suiker meer koste-effektief te produseer. Buitendien, die afnemende tendens en fluktuasies in die wêreld se suikerpryse maak produkdiversiteit noodsaaklik om die ekonomiese volhoubaarheid van die industrie te verseker. Daarom sal die verbetering van die huidige energiestatus van die suikerindustrie suikerrietbronne beskikbaar maak vir verdere valorisasie. Hierdie studie beoog om die energiedoeltreffendheid van 'n tipiese Suid-Afrikaanse suikermeule te verbeter na ekonomiese volhoubaarheid en mededingendheid.

Die eerste doelwit (doelwit 1) om hierdie mikpunt te bereik, was om die ligging, groottes, en oorsake van ondoeltreffendhede in 'n tipiese Suid-Afrikaanse suikermeule te identifiseer deur eksergie-analise. Hierdie analise is gebaseer op streng massa- en energiebalanse bereken uit 'n Aspen Plus®-simulasie van 'n tipiese 250 ton per uur suikermeule. Die kogenerasiesstelsel het die hoogste eksergie verwoesting (90 582 kW) gehad, wat 86% van die totale suikermeule onomkeerbaarheid verteenwoordig. Met die laagste eksergieke doeltreffendheid van 9.6%, het die kristallisatie-eenheid egter die mees ondoeltreffende gebruik van energie aangeteken as gevolg van die proseskompleksiteit.

Na afleiding van die eksergie resultate, was doelwit 2 om energiedoeltreffende tegnologieë te kies om die suikermeuldoeltreffendheid te verbeter, terwyl doelwit 3 die assessering van die ekonomiese uitvoerbaarheid van die integrasie daarvan in die meule ingehou het. Organiese Rankine siklus (ORC) en absorpsie verhittingspomp (AHP) tegnologieë is gekies, wat die kogenerasie eksergie doeltreffendheid met 1.7% verbeter het en die algehele stelsel onomkeerbaarheid met 0.14% geminimeer het, wat 0.83% totale bagasse vir valorisasie spaar, onderskeidelik. Al is slegs marginale verbetering bereik, is beide ORC- en AHP-integrasie ekonomies uitvoerbaar en kan geoptimeer word om na energieverbeteringe te lei.

Verder is belowende bioraffineerdersprodukte suksiensuur (SA) en kortketting-fruktooliggosakkarieses (scFOS) geïntegreer vir die ekonomiese mededingendheid van die industrie in doelwit 4. Gebaseer op die eksergie resultate, is die bioraffineerderie ontwikkel om A-molasse te gebruik vir die produksie van SA en scFOS in sewe verskillende scenario's, waarvan almal hoogs winsgewend was met interne opbrengskoerse (iok) van tussen 24 en 62%

in vergelyking met die minimum vereiste iok van 9.7% as gevolg van die integrasie voordele. Verder, kogebruik van C-molasse en lignoselluloserestu as eerste- en tweede-generasie (1G-2G) voermateriaal vir die produksie van SA, is ook oorweeg om die suikerrietplant ten volle te valoriseer met die inagneming van die huidige kristallasieskema.

Doelwit 5 het 'n versamelde stelsel eksergie-ekonomiese metodologie toegepas om die holistiese doeltreffendheid van die bioraffineerderye te assessee en om die mees koste-effektiewe een te identifiseer. Met die laagste koers van 1 029 US\$/h en eksergie-ekonomiese faktor van 0.56, het die scFOS-poeier scenario (S-FP) 'n goeie balans tussen die onomkeerbaarheid- en belegging-verwante kostes en is oorweeg as die mees koste-effektiewe bioraffineerderye vir integrasie in die suikermeule. Oor die algemeen het hierdie studie 'n wye spektrum oplossings vir die energie en ekonomiese uitdagings van die suikerindustrie getoon om verder ondersoek te word vir implementasie, deur eksergie/eksergie-ekonomiese metodologie as beter ontwerphulpmiddels te gebruik eerder as konvensionele energie en ekonomiese analise.

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Dedication

I dedicate this dissertation to the praise and glory of my God and Saviour Jesus Christ by whose grace I live and do exploits.

“All the glory must be to the Lord

For He is worthy of our praise

No man on earth should give glory to himself

All the glory must be to the Lord”

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Nomenclature

List of units

GJ	gigajoule
h	hour
J	joule
K	kelvin
kg	kilogram
kJ	kilojoule
kt	kiloton
kW	kilowatt
mol	mole
MW	megawatt
s	second
t	ton
w%	weight percent
y	year
%	percent
MUS\$	Million United States of America dollar (MUSD)
US\$	United States of America dollar (USD)

List of symbols

a	stoichiometric coefficient of oxygen
A	ash content, w%
b	stoichiometric coefficient of bagasse component
B	brix, %
c	unit exergy cost, US\$/GJ
\dot{C}	cost rate, US\$/h
CFR	capital recovery factor
ex	specific exergy, J/kg or J/mol
$\dot{E}x$	exergy rate, kW
f	exergoeconomic factor
g_f	free energy of formation, J/mol
h	specific enthalpy flow, kJ/kg
H	annual operational hours, h
i	interest rate
\dot{I}	irreversibility or exergy destruction rate, kW
$\dot{I}P$	exergetic improvement potential, kW
M	moisture content, w%
N	plant life, y
\dot{m}	mass flow rate, kg/s
\dot{n}	mole flow rate, mol/s
P	pressure, bar
\dot{Q}	heat flow rate, kW
r	relative cost difference
s	specific entropy, J/kg.K
T	temperature, K
\dot{W}	electrical energy/exergy flow, kW
Z	capital cost, US\$
\dot{Z}	capital cost rate, US\$/h
x	mole fraction
ε	heat exchanger effectiveness
η	boiler efficiency, %
ψ	exergy efficiency, %
\emptyset	maintenance factor
<i>Waste-heat sources</i>	
A	flue gas
B	boiler blowdown
C	flash drum vapour
<i>Heat sinks for absorption heat pump</i>	
1	boiler make-up water
2	sugar drying air

List of subscripts and superscripts

<i>ch</i>	chemical
<i>D</i>	destruction
<i>F</i>	fuel
<i>FW</i>	feed water
<i>i</i>	stream component
<i>in</i>	inlet / input
<i>j</i>	element
<i>k</i>	exergy stream/
<i>L</i>	losses
<i>out</i>	outlet/ output
<i>P</i>	product
<i>ph</i>	physical
<i>st</i>	steam
<i>used</i>	resource utilised
<i>w</i>	waste
0	environmental state
00	dead state
1, 2, 3, etc	state points of streams
I, II, III, IV etc	process units

List of abbreviations

1G; 2G	first generation; second generation
AHP	absorption heat pump
BDO	1,4-butanediol (bio-butanediol)
BIGCC	biomass integrated gasification combined cycle
BPST	back pressure steam turbine
CEST	condensing-extraction steam turbine
COP	coefficient of performance
GCV	gross calorific value
HMF	5-hydroxymethyl furfural
HP	high pressure
IBL	inside battery limit
IRR	internal rate of return
MSP	minimum selling price
OBL	outside battery limit
OPEX	operating costs
ORC	organic Rankine cycle
PEC	purchased equipment cost
SA	succinic acid
scFOS	short-chain fructooligosaccharides
SPECO	specific exergy costing
SHF	separate hydrolysis and fermentation
TCI	total capital investment

Chapter 1

Introduction

1.1. Background

The sugar industry is one of the oldest agro-industries in the world, producing 80 % of the world's sugar from sugarcane grown in tropical regions, and the remaining from sugar beet grown in temperate regions. Sugarcane is the world's largest cash crop and is grown in over 90 countries for sugar production (Bezerra & Ragauskas, 2016; International Sugar Organization, 2016; Pippo & Luengo, 2013). The cane sugar production process is energy-intensive. However, the energy required for the process is generated by burning bagasse (the sugarcane residue after juice extraction) in a cogeneration system consisting mainly of boilers and steam turbines, making the process energy self-sufficient (Moya et al., 2013).

Currently, conditions such as facility ageing, excess steam requirements of various auxiliary processes, as well as the demand on bagasse for value-added products have necessitated the use of supplementary energy sources such as coal in the South African sugar industry (M. J. Reid, 2006). This is because traditional sugar mill cogeneration systems were designed inefficiently to get rid of excess bagasse, due to challenges with its disposal and unforeseen prospects in utilizing it for value-added products (Kamate & Gangavati, 2009a; Meyer, Rein, Turner, & Mathias, 2011). The use of coal in the sugar industry is undesirable due to its dwindling reserves and associated environmental pollution (Ali Mandegari, Farzad, & Görgens, 2017; M. J. Reid, 2006). Therefore, it is desirable to improve the energy efficiency of existing sugar mills to make bagasse available for further valorisation and to avoid the need for supplementary fuel. The recent development of sugarcane biorefineries is an added motivation to improve the energy efficiency of sugar mills, to save sugarcane resources as raw materials for the biorefineries.

Sugar industries all over the world are undergoing a change from sugar-only production to diversifying their product base in a biorefinery concept (Anouar, Abderafi, & Bounahmidi, 2016; Martínez-Guido, Betzabe González-Campos, Ponce-Ortega, Nápoles-Rivera, & El-Halwagi, 2016; Moncada, El-Halwagi, & Cardona, 2013; Renó, Olmo, Palacio, Lora, & Venturini, 2014). The rising trend of sugarcane-biorefinery stems from a global shift towards bio-based products to replace fossil-based ones due to increasing environmental issues, (Krajnc & Glavič, 2009; Vaswani, 2010). Moreover, increasing consumer demand for healthier food products is causing a drop in global sugar

demand and prices, which threatens the economic sustainability of traditional sugar mills producing sugar only, forcing the industry to diversify (Eggleston & Lima, 2015; OECD/FAO, 2016). The biorefinery concept presents the possibility of multiple products including biochemicals, bioenergy, biofuels and food products from a common feedstock (Moncada et al., 2013). Two of the products that have recently received much attention due to their importance and promising market include succinic acid (SA) and short-chain fructooligosaccharides (scFOS).

Succinic acid is reported among the top twelve biochemicals with near-term deployment potential based on its large projected market because of its potential as a platform chemical (feedstock for higher-value products) for more than thirty commercially important products, and its wide usage in the food, pharmaceuticals and chemicals industries (Barros, Freitas, Padilha, & Alegre, 2013; Vaswani, 2010). Likewise, short-chain fructooligosaccharides (scFOS), commercially produced from sucrose, has gained increasing importance in the food and nutraceutical industries as low-calorie sweeteners and prebiotics, which are now in great demand because of common health challenges. As prebiotics, scFOS stimulates the growth of colon probiotic bacteria, i.e. bifidobacteria, and inhibit the growth of harmful microorganisms, preventing colon cancer among other health benefits (Bali, Panesar, Bera, & Panesar, 2015). Consequently, succinic acid and scFOS have great economic potential for the biorefinery move of the sugar industry due to their increasing market size resulting from high demands of them (Bedzo, Mandegari, & Görgens, 2019; Nieder-Heitmann, Haigh, & Görgens, 2019; Ur Rehman, Kovacs, Quitmann, Ebrahimi, & Czermak, 2016). However, adding extra products also places extra energy demand on the current sugar mill. Thus, improved energy efficiency is required to make available both feedstock and process energy for the new products to be added.

With these motivations in view, it is necessary to assess the current energy status of sugar industries and identify components and locations of inefficiencies for improvement and suitable points for biorefinery integration for an energy-efficient and cost-effective industry. The traditional method for assessing the energy efficiency of manufacturing processes has been conservation of energy analysis, based on the first law of thermodynamics. However, since this method does not account for entropy generation in real processes, which leads to degradation of energy quality, a supplementary method called the exergy analysis was developed (Hevert & Hevert, 1980). Exergy analysis is instrumental in identifying the location, magnitude, and causes of energy quality (exergy) degradation, using both the first and second laws of thermodynamics. Thus, exergy analysis provides a more comprehensive insight into efficiency improvement measures to be taken. Its main limitation, however, is that it does not account for economic trade-offs, necessary for decision making. A more useful and advanced methodology is the exergoeconomic analysis, which combines exergy analysis with economic reasoning to holistically assess energy systems for improvement (G Tsatsaronis & Czesla, 1999).

In this study, a comprehensive exergy analysis methodology was presented to assess the causes, locations, and magnitude of inefficiencies in a typical sugarcane mill, to allow better operation of the processes for better energy quality preservation. Two possible ways of improvement were analysed based on the exergy assessment results. Firstly, organic Rankine cycle (ORC) and absorption heat pump (AHP) technologies were integrated with cost considerations to improve the exergy performance of the mill through waste heat recovery. Secondly, two biorefinery products with promising market demands, succinic acid and short-chain fructooligosaccharides (scFOS), were introduced into the existing sugar mills for more efficient use of energy resources in the mill. The concept, design and integration of the biorefinery were based on the knowledge of inefficiencies identified from the exergy analysis of the mill. Moreover, by applying exergoeconomic analysis, which is an exergy-based costing analysis to assess the holistic performance of the processes, the preferred biorefinery scenario was selected for a cost-effective and energy-efficient integration into the sugar mill. Overall, the study seeks to present an energy-efficient sugar industry with an economically viable integrated biorefinery. The entire work was simulation-based accomplished in Aspen Plus® software as robust chemical process simulator.

1.2. Aim and objectives

This overall aim of this work is to improve the energy efficiency and ensure the economic sustainability and competitiveness of the sugar industry by assessing the industry's energy inefficiency, implementing energy-efficient technologies and integrating biorefinery products into existing sugar mills in a cost-effective manner.

The specific objectives set out to achieve this aim are

- i. To evaluate the inefficiency of a typical South African sugar mill through exergy analysis. This involved the evaluation and modification of an existing Aspen Plus simulation of the sugar mill, in addition to a new detailed simulation of a cogeneration system corresponding to the sugar mill.
- ii. To select suitable energy-efficient technologies for integration into the base sugar mill towards energy/exergy efficiency improvement.
- iii. To assess the economic feasibility of integrating the selected technologies into the sugar mill.
- iv. To design economically feasible biorefineries integrated into the sugar mill based on exergy principles.
- v. To apply exergoeconomic analysis for the selection of the preferred biorefinery for a cost-effective integration into the sugar mill.

1.3. Dissertation layout

This dissertation consists of seven (7) chapters. Following the background and objectives in this chapter (Chapter 1), Chapter 2 presents the sugar industry processes and energy use, the exergy methodology as well as the opportunities for improving the energy efficiency of the industry. In line with economic competitiveness of the industry, and exergoeconomic analysis as well as the integration of biorefinery products. Chapters 3, 4, 5 and 6 contain journal publications (Chapters 3 and 4) and draft manuscripts (Chapters 5 and 6) for publication addressing the methodology, results and discussions of the objectives of this dissertation. Chapter 3 is about the assessment of the energy performance of the sugar mill through exergy analysis methodology. The assessment of thermodynamic improvement achieved through the integration of waste-heat recovery technologies is detailed in Chapter 4. In Chapter 5, the economic viability of integrated biorefineries annexed a sugar mill, designed to use A-molasses based on the exergy outcome of Chapter 3, is assessed. Chapter 6 focuses on selecting the preferred biorefinery by assessing the holistic performance of the integrated biorefineries using exergoeconomic methodology. The conclusions and recommendations of the dissertation are presented in Chapter 7.

References

- Ali Mandegari, M., Farzad, S., & Görgens, J. F. (2017). Economic and environmental assessment of cellulosic ethanol production scenarios annexed to a typical sugar mill. *Bioresource Technology*, 224, 314–326. <http://doi.org/10.1016/j.biortech.2016.10.074>
- Anouar, M., Abderafi, S., & Bounahmidi, T. (2016). Promising bioethanol processes for developing a biorefinery in the Moroccan sugar industry. *International Journal of Hydrogen Energy*, 41(45), 20880–20896. <http://doi.org/10.1016/j.ijhydene.2016.07.035>
- Bali, V., Panesar, P. S., Bera, M. B., & Panesar, R. (2015). Fructo-oligosaccharides: Production, Purification and Potential Applications. *Critical Reviews in Food Science and Nutrition*, 55(11), 1475–1490. <http://doi.org/10.1080/10408398.2012.694084>
- Barros, M. De, Freitas, S., Padilha, G. S., & Alegre, R. M. (2013). Biotechnological Production of Succinic Acid by *Actinobacillus Succinogenes* Using Different Substrate . *Chemical Engineering Transactions*, 32, 985–990.
- Bedzo, O. K. K., Mandegari, M., & Görgens, J. F. (2019). Comparison of immobilized and free enzyme systems in industrial production of short-chain fructooligosaccharides from sucrose using a techno-economic approach. *Biofuels, Bioproducts and Biorefining*, bbb.2025.

<http://doi.org/10.1002/bbb.2025>

- Bezerra, T. L., & Ragauskas, A. J. (2016). A review of sugarcane bagasse for second-generation bioethanol and biopower production. *Biofuels, Bioproducts and Biorefining*, *10*, 634–647. <http://doi.org/10.1002/bbb>
- Eggleston, G., & Lima, I. (2015). Sustainability Issues and Opportunities in the Sugar and Sugar-Bioproduct Industries. *Sustainability*, *7*(9), 12209–12235. <http://doi.org/10.3390/su70912209>
- Hevert, H. W., & Hevert, S. C. (1980). Second law analysis: An alternative indicator of system efficiency. *Energy*, *5*(8–9), 865–873. [http://doi.org/10.1016/0360-5442\(80\)90102-4](http://doi.org/10.1016/0360-5442(80)90102-4)
- International Sugar Organization. (2016). About sugar. Retrieved August 28, 2017, from <http://www.isosugar.org>
- Kamate, S. C., & Gangavati, P. B. (2009). Cogeneration in Sugar Industries: Technology Options and Performance Parameters—A Review. *Cogeneration & Distributed Generation Journal*, *24*(4), 6–33. <http://doi.org/10.1080/15453660909595148>
- Krajnc, D., & Glavič, P. (2009). Assessment of different strategies for the co-production of bioethanol and beet sugar. *Chemical Engineering Research and Design*, *87*(9), 1217–1231. <http://doi.org/10.1016/j.cherd.2009.06.014>
- Martínez-Guido, S. I., Betzabe González-Campos, J., Ponce-Ortega, J. M., Nápoles-Rivera, F., & El-Halwagi, M. M. (2016). Optimal reconfiguration of a sugar cane industry to yield an integrated biorefinery. *Clean Technologies and Environmental Policy*, *18*(2), 553–562. <http://doi.org/10.1007/s10098-015-1039-1>
- Meyer, J., Rein, P., Turner, P., & Mathias, K. (2011). *Good management practices manual for the cane sugar industry (Final)*. Johannesburg, South Africa.
- Moncada, J., El-Halwagi, M. M., & Cardona, C. A. (2013). Techno-economic analysis for a sugarcane biorefinery: Colombian case. *Bioresource Technology*, *135*, 533–543. <http://doi.org/10.1016/J.BIORTECH.2012.08.137>
- Moya, C., Domínguez, R., Van Langenhove, H., Herrero, S., Gil, P., Ledón, C., & Dewulf, J. (2013). Exergetic analysis in cane sugar production in combination with Life Cycle Assessment. *Journal of Cleaner Production*, *59*, 43–50. <http://doi.org/10.1016/j.jclepro.2013.06.028>
- Nieder-Heitmann, M., Haigh, K., & Görgens, J. F. (2019). Process design and economic evaluation of integrated, multi-product biorefineries for the co-production of bio-energy, succinic acid, and

polyhydroxybutyrate (PHB) from sugarcane bagasse and trash lignocelluloses. *Biofuels, Bioproducts and Biorefining*, 13(3), 599–617. <http://doi.org/10.1002/bbb.1972>

OECD/FAO. (2016). Sugar. In *OECD-FAO Agricultural Outlook 2016-2025* (pp. 1–13). Paris: OECD Publishing, Paris. http://doi.org/10.1787/agr_outlook-2016-9-en

Pippo, W. A., & Luengo, C. A. (2013). Sugarcane energy use : Accounting of feedstock energy considering current agro-industrial trends and their feasibility. *International Journal of Energy and Environmental Engineering*, 4(10), 1–13.

Reid, M. J. (2006). Why do we continue to burn so much coal? In *Proceedings of The South African Sugar Technologists' Association* (Vol. 80, pp. 353–363).

Renó, M. L. G., Olmo, O. A. del, Palacio, J. C. E., Lora, E. E. S., & Venturini, O. J. (2014). Sugarcane biorefineries: Case studies applied to the Brazilian sugar–alcohol industry. *Energy Conversion and Management*, 86, 981–991. <http://doi.org/10.1016/j.enconman.2014.06.031>

Tsatsaronis, G., & Czielsa, F. (1999). Strengths and limitations of exergy analysis. In *Thermodynamic Optimization of Complex Energy Systems* (Vol. I, pp. 93–100). Dordrecht: Springer Netherlands. http://doi.org/10.1007/978-94-011-4685-2_6

Ur Rehman, A., Kovacs, Z., Quitmann, H., Ebrahimi, M., & Czermak, P. (2016). Enzymatic production of fructo-oligosaccharides from inexpensive and abundant substrates using a membrane reactor system. *Separation Science and Technology*, 01496395.2016.1167740. <http://doi.org/10.1080/01496395.2016.1167740>

Vaswani, S. (2010). *Bio-based succinic acid*. Menlo Park, California 94025. Retrieved from https://www.ihs.com/pdf/RW2010-14_220240110917062932.pdf

Chapter 2

Literature review

This chapter first of all reviews the production processes of a typical South African raw sugar mill and assesses its energy use. Secondly, the methodology for revealing the true thermodynamic efficiency of energy systems, i.e. by exergy analysis, is presented together with its advanced form, the exergoeconomic analysis. Finally, two aspects of sustainability for the sugar industry are presented namely, improving energy and exergy efficiency using energy-efficient technologies and improving economics through integrated biorefineries, with a focus on succinic acid (SA) and short-chain fructooligosaccharides (scFOS) as examples for application of this methodology.

2.1. Sugar production from sugarcane

More than 75 million metric tons of sugar was produced from sugarcane worldwide in 2015 (International Sugar Organization, 2016). A typical sugar production plant (sugar mill) involves a series of liquid-solid and liquid-vapour separation methods, such as juice extraction, juice clarification, evaporation and crystallization to isolate the sugar in the sugarcane (Hugot, 1972). The required moisture content of the final sugar product is obtained by means of drying the sugar in a drier. In most sugar cane mills, the energy demand of the process in the form of steam (heat) and electricity, is provided by burning the cane fibre separated from the juice at the extraction stage as fuel in a cogeneration system ((Narasimha Rao & Nagarajan, 2012). Cooling is provided by means of cooling towers or spray ponds. As a result, cogeneration system and the cooling tower form part of the sugar mill systems as utility plants. The main processes that make up a raw sugarcane mill as illustrated in Figure 2-1 are discussed in this section. Table 2-1 shows the principal equipment of each systems. All stream tags used in Figure 2-2 to Figure 2-8 are described in Table 2-2.

Table 2-1: The principal equipment of the sugar mill systems (Guest, Stark, & Starzak, 2019)

Unit	Principal components
Juice Extraction	cane knives, shredders, a counter-current diffuser, bagasse dewatering mills, heat exchangers and mechanical drives.
Juice Clarification	mixed juice tank, mixed juice heaters, juice flash tank, clarifier and filter station
Evaporation	clarified juice preheater, a 5-effect evaporator station, throttle valve for brix control, flash tanks, syrup filter, barometric condenser
Crystallization	3-stages (A, B and C) of vacuum pans, crystallisers and centrifuges, remelter, mingler and barometric condenser
Drying	air heater, sugar dryer
Cooling tower	cooling tower
Cogeneration	boiler and steam turbine

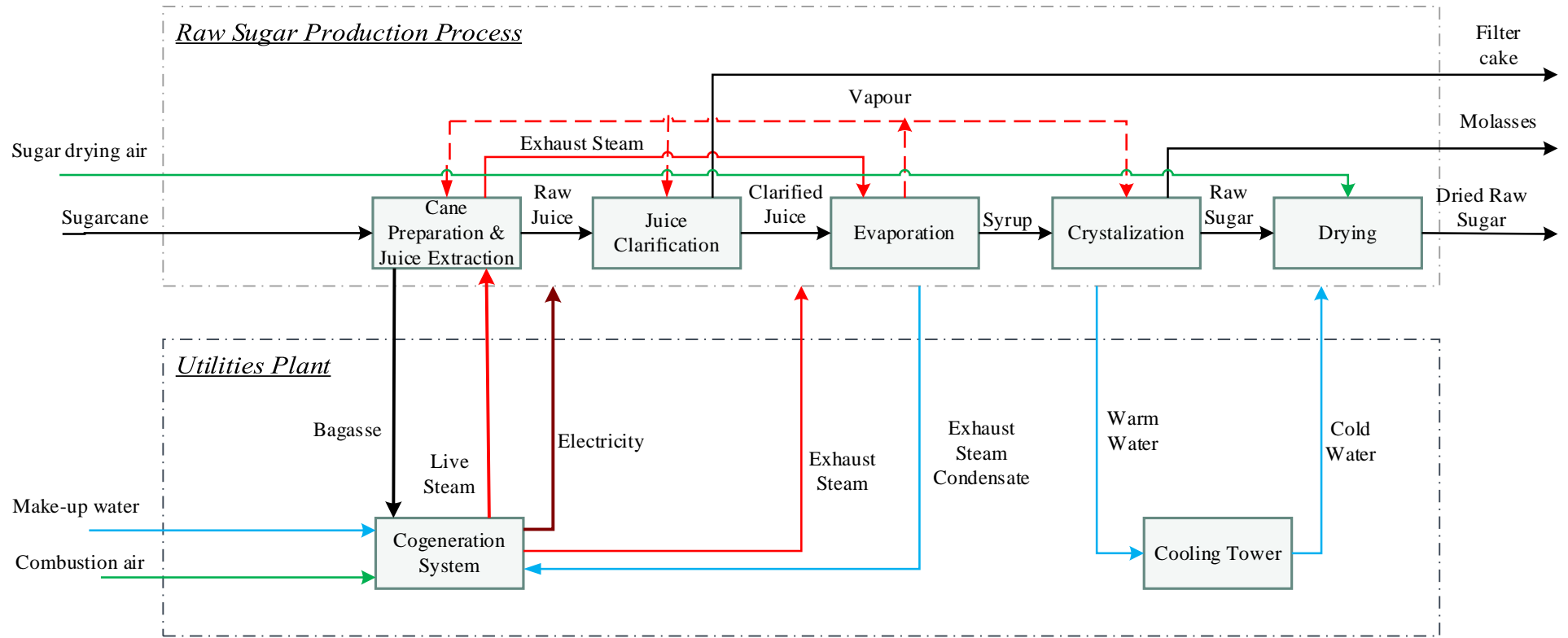


Figure 2-1: Raw Sugar Mill Process Block Diagram (redrawn based on (Starzak & Davis, 2015))

Table 2-2: Description of stream tags in Figure 2-2 to Figure 2-8

Stream tag	Description	Stream tag	Description
AIR	Air	CT2	Cooling water to pan barometric condenser
ASH	Ash	CTV	Cooling tower vapours
BAG	Total bagasse produced	CWA1	Cold water to A-cooler
BAG1	Total bagasse to cogeneration unit	CWA2	Outlet water from A- cooler
BAG2BOIL	Bagasse burnt in boiler unit	CWB1	Cold water to B-cooler
BAGOUT	Bagasse exported	CWB2	Outlet water from B- cooler
BGO	Bagacillo	CWC1	Cold water to C-cooler
CANE	Raw cane	CWC2	Outlet water from C- cooler
CCF	Total centrifuge wash water	DAI	Drying air
CDH	Diffuser heaters condensate	DAO	Air from drying plant
CER	Condensate to remelter	DJ	Draft juice
CHP	Condensate from primary heater	DSTEAM	Steam to deaerator
CHS	Condensate from secondary heater	DVENT	Deaerator vent
CHT	Condensate from tertiary heater	EXCS	Condensate from drying plant
CJ	Clarified juice	EXCS1	Drying plant condensate from condensate tank
CKA	A-pan condensate	EXSS	Exhaust steam to drying plant
CKB	B-pan condensate	FC	Filter cake
CKC	C-pan condensate	FLUEGAS	Flue gas released to atmosphere
CT1	Waste water	IW	Imbibition water to cane diffuser

Table 2-2: Description of stream tags in Figure 2-2 to Figure 2-8 (continued)

Stream tag	Description	Stream tag	Description
LIM	Milk of lime	SL-EXS	Exhaust steam sundry losses
M	Cooling water to evaporator barometric condenser	SL-HPS	HP steam sundry losses
MOLC	C-molasses from C-centrifuge	SLU	Sludge from syrup filter
SB0	Live steam to turbo alternator	SPW	Steam to press water heater
SB00	Exhaust steam from turbo alternator	SUA	Sugar from dryer
SB001	Desuperheated steam	SUGA	A-sugar from A-centrifuge
SB1	Live steam to mill turbines	SYR	Final syrup from syrup filter
SB2	Exhaust steam to evaporation plant	VMJ	Vapour from flash drum
SBF	Live steam from boiler	W	Water from evaporator barometric condenser
SD7	Total exhaust steam from extraction plant	WB	Condensate exported as Boiler feed water from deaerator
SDH	Steam to diffuser heaters	WBB	Boiler blowdown
SDI	Steam injected to diffuser	WBF	Boiler feed water from evaporators
SER	Steam to remelter	WBF1	Boiler feed water from condensate tank
SHP	Steam to primary heater	WBM	Make-up boiler feed water
SHS	Steam to secondary heater	W-DESUP	Desuperheating water
SHT	Steam to tertiary heater	WEC	Condensate exported from sugar mill
SKA	A-pan heating steam	WK	Water from pan barometric condenser
SKB	B-pan heating steam	WTM	Make-up water to cooling tower
SKC	C-pan heating steam	WW1	Wash water to vacuum filter

2.1.1. Cane preparation and juice extraction

The purpose of cane preparation before juice extraction is to reduce the cane to sufficiently small sizes, to be suitable for juice extraction (Rein, 2007). Using knives and shredders (fiberizers), the hard structure of the cane is broken and the juice cells of the cane are ruptured to ease juice extraction (Seebaluck, Mohee, Sobhanbabu, Leal, & Johnson, 2008).

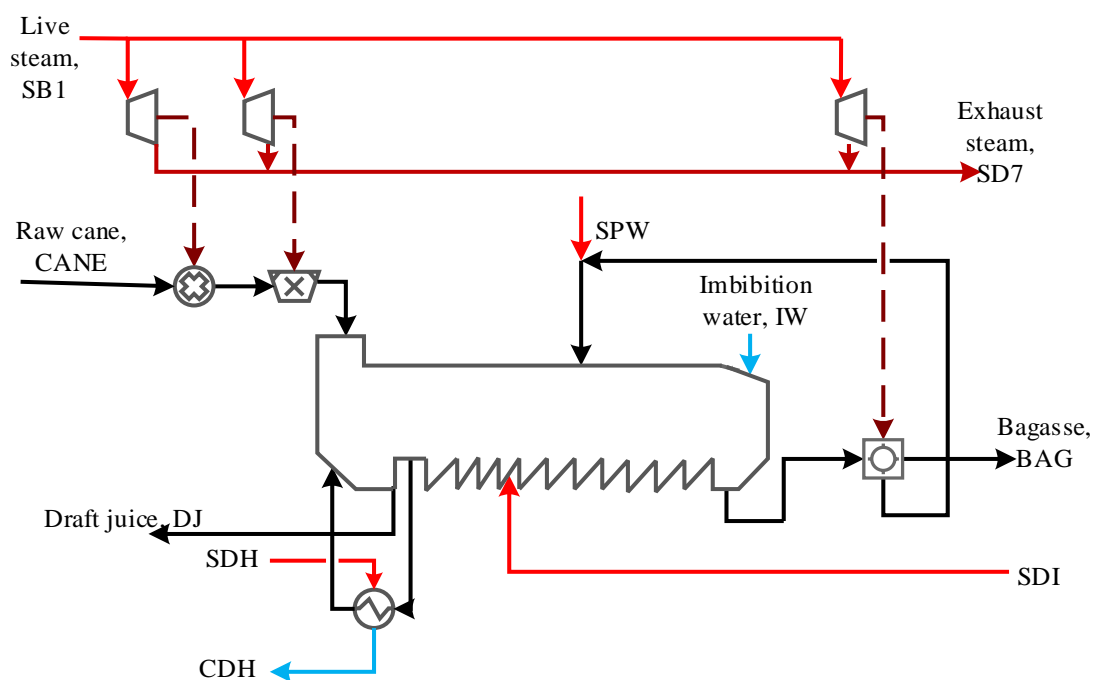


Figure 2-2: Process flow diagram of cane preparation and juice extraction unit (Redrawn based on (Starzak & Zizhou, 2015)). Stream tags are described in the nomenclature.

There are two methods by which juice can be extracted from the prepared cane: milling and diffusion. The milling process uses a series of mills called the milling tandem, made up of rotating rolls. The rotating rolls compress the prepared cane and squeeze out juice from ruptured cells, leaving the fibre, known as bagasse. By adding imbibition water, the amount of juice that leaves with the bagasse is reduced. Hot water imbibition eases breaking of unbroken cells for maximum extraction. The milling process is highly energy intensive due to the high power requirements of mill drives. It accounts for 60 – 65% of energy requirement of sugar mills (Seebaluck et al., 2008). Continuous diffusion is a more preferred method of juice extraction in terms of energy consumption, even though higher imbibition water volumes are required (Hugot, 1972; Mbohwa, 2013). The diffusion method of extraction (illustrated in Figure 2-2) operates by a hot water leaching process, where weak juice and imbibition water flows over the prepared cane in a counter current manner. Extraction by diffusion depends

largely on the proportion of ruptured cells (Hugot, 1972), therefore, heavy duty shredders are required to get between 90-94% of the juice storage cells ruptured. Usually, the diffuser is operated at a temperature between 70 and 75 °C with low-pressure steam from the evaporators, to prevent growth of microorganisms and to minimize sucrose inversion (Seebaluck et al., 2008). Even though higher temperatures reduce fluid viscosity, thus promoting rate of sugar extraction (dissolution), its effect on extraction is not as important as factors such as cane preparation and imbibition rate (Rein, 2007). Moreover, the operating temperatures allow the use of low quality steam from evaporators for heating purposes in the diffuser. The diffuser is accompanied by two sets of bagasse (dewatering) mills that reduce the bagasse moisture content from 85% to about 50%. The dewatering mills also contribute to the extraction process (Modesto, Zemp, & Nebra, 2009; Seebaluck et al., 2008). Diffusers are common in South African sugar mills (Gurumurthy, 2011; Seebaluck et al., 2008).

2.1.2. Juice clarification

Apart from sucrose and water, the raw juice contains some non-sucrose impurities and fine bagasse, which are removed by a series of processes in the juice clarification system, including heating, addition of lime, clarification and mud filtration, as illustrated by Figure 2-3. The raw juice is heated and lime is added to cause precipitation of impurities, which are subsequently separated from the clear juice by means of settling in clarifiers. Addition of lime to the hot juice provides a suitable pH (nearly neutral) for clarification, to prevent sucrose inversion in subsequent evaporation. Before the clarifier, the treated juice is flashed to remove dissolved air from the mixture and to enhance flocculent settling (Rein, 2007; Starzak & Davis, 2016). A suspension of precipitated impurities in clear juice (mud), settles at the bottom of clarifiers and is filtered in a rotary vacuum filter. In factories where diffusers are used for juice extraction instead of the milling tandem, there is an attractive option to recycle mud back to the diffuser (the cane bed in the diffuser serving as a filter) and eliminate the filtration step (Gurumurthy, 2011; Mbohwa, 2013; Meadows, Schumann, & Soji, 1998).

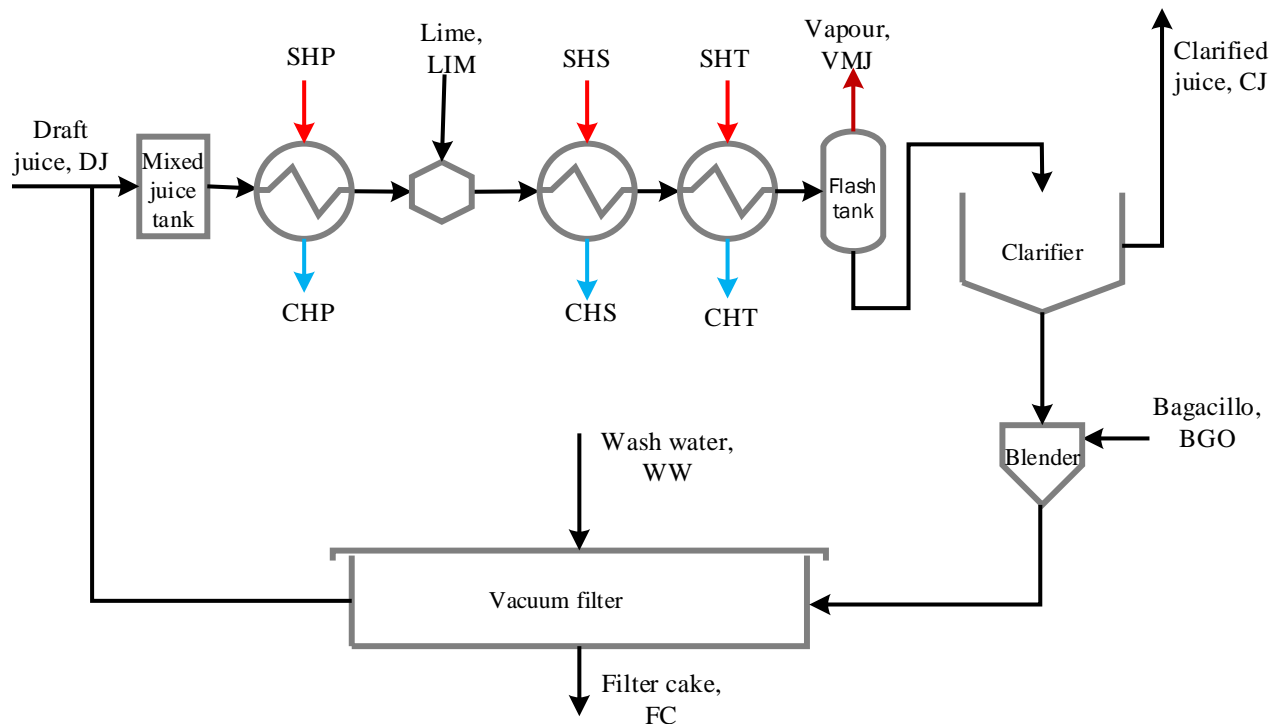


Figure 2-3: Process flow diagram of juice clarification unit (Redrawn based on (Starzak & Zizhou, 2015)). Stream tags are described in the nomenclature.

2.1.3. Juice concentration (evaporation)

Clarified juice is mostly made of sugar dissolved in a large quantity of water (about 15 °Brix¹) (A. V Ensinas, Nebra, Lozano, & Serra, 2006; Higa, Freitas, Bannwart, & Zemp, 2009). A large portion of the water is removed by boiling the juice to concentrate the sugar. Evaporation is the first stage of the concentration process where clarified juice is concentrated to approx. 65–68 °Brix, making use of the most part of exhaust steam (Rein, 2007). The clarified juice is first preheated to about 110 °C in a heat exchanger and then concentrated in a multiple-effect evaporator. A typical South African mill operates a four or five-effect evaporator. The evaporator type commonly used is vertical tube or Robert evaporators. Figure 2-4 illustrates a five-effect evaporation system of a typical sugar mill.

¹ Brix is juice concentration expressed as grams of solids per 100 g of water (Heluane, Colombo, Hernández, Graells, & Puigjaner, 2007).

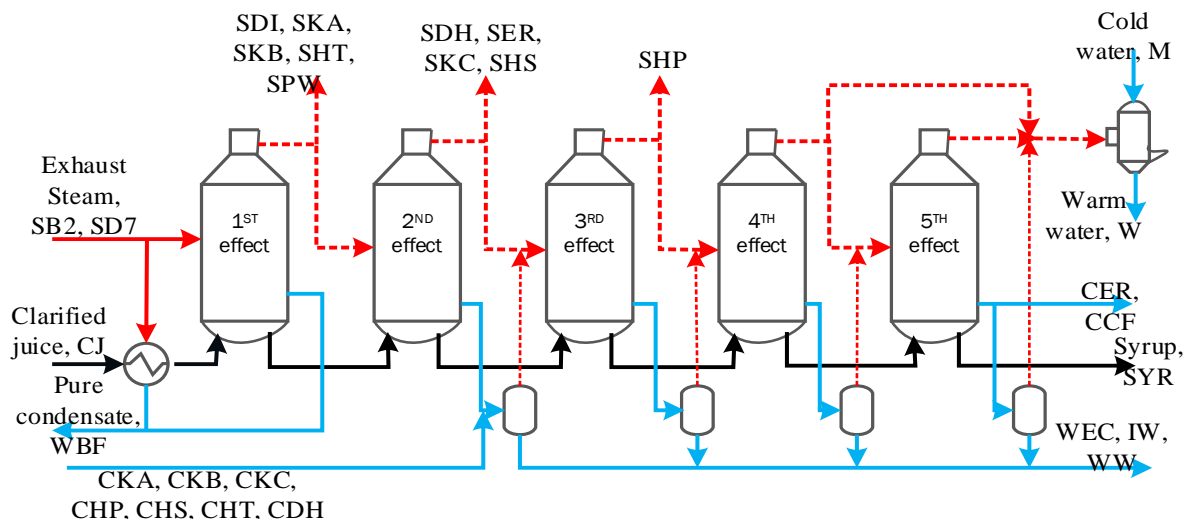


Figure 2-4: Process flow diagram of evaporation unit (Redrawn based on (Starzak & Zizhou, 2015))

. Stream tags are described in the nomenclature.

Unlike in the beet sugar industry, where pressure evaporation is practiced, evaporators in sugar cane mills operate under vacuum pressure – from exhaust steam pressure of about 2 bar in the first-effect to about 0.15 bar (vacuum) in the last-effect. This way, the temperature is kept low to prevent excessive sucrose inversion and exhaust steam usage is reduced (Rein, 2007). As has been noted, the first-effect evaporator uses exhaust steam from the cogeneration and milling turbines. The consequent evaporators, each utilizes vapour from the preceding-effect, and are operated under vacuum, sustained by barometric condenser. The reduced pressure creates the necessary temperature difference between heating vapour and juice for evaporation to take place (A. V. Ensinas, Modesto, Nebra, & Serra, 2009). The exhaust steam condenses after heating the first-effect evaporator and returns to the cogeneration system as boiler feed. Vapour condensates from heating other evaporators are flashed to increase steam economy prior to their use as process water. The concentrated juice from the last-effect evaporator is known as syrup, which is further boiled in the crystallization section.

2.1.4. Crystallization (sugar boiling) and centrifuging

Crystallization, as described by Rein (2007), is carried out under vacuum either in conventional batch or continuous vacuum pans, where the syrup is boiled until crystals begin to form. In the course of boiling, the syrup is seeded with fine sugar crystals in the form of a slurry, to initiate crystallization. Boiling then continues slowly, allowing crystals to grow and to form a mixture of crystals and the mother liquor called massecuite. The hot massecuite is discharged from the

vacuum pan into big tanks known as crystallizers for cooling to allow further crystallization. After sufficient crystallization, the sugar crystals are separated from the massecuite in centrifuges. This is done by filtering the liquor (molasses) through screens with the help of centrifugal force and concurrent washing of the sugar crystals.

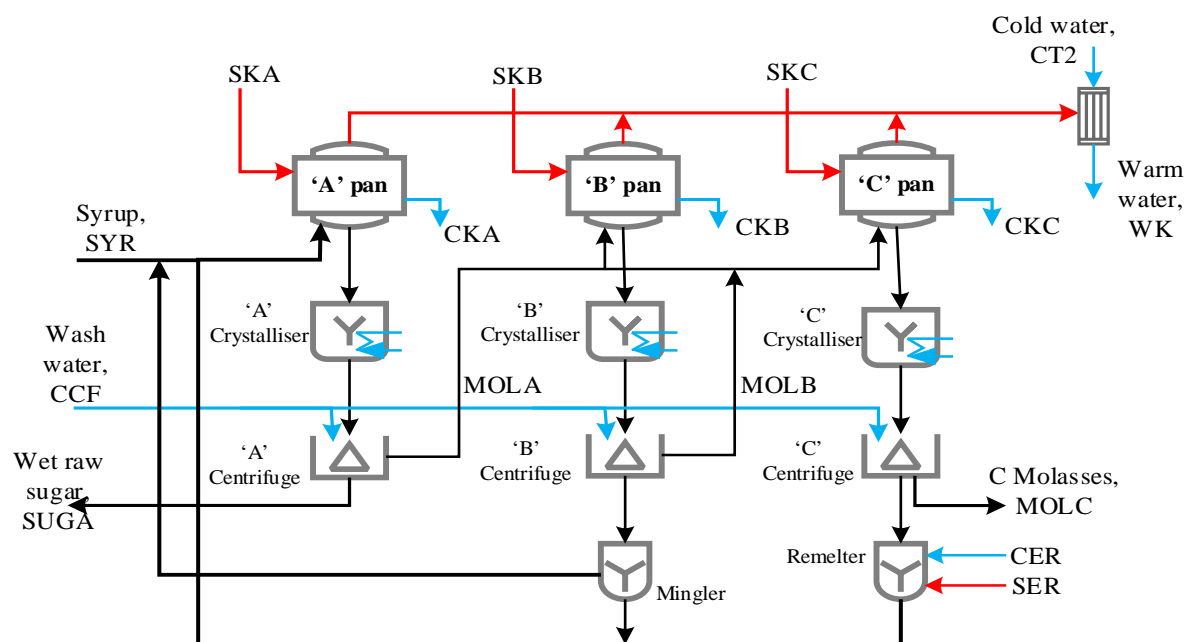


Figure 2-5: Process flow diagram of sugar crystallization unit (Redrawn from (Starzak & Davis, 2016)). Stream tags are described in the nomenclature.

To ensure minimal loss of sugar in the final molasses and recovery of maximum crystalline sucrose, the whole crystallization process is carried out in three stages, as shown in Figure 2-5. The three (A, B and C) stages in South African mills produce A sugar and C molasses. Lower grade sugar produced from the B and C centrifuges are either melted (dissolved in water or clarified juice) in remelter or mingled to serve as seed in the vacuum pans, while A and B molasses are recycled to the Band C vacuum pans (Starzak & Davis, 2016). Finally, the C molasses comes out as a by-product and the moist and hot A sugar (raw sugar) is passed on to be dried (Seebaluck et al., 2008).

2.1.5. Sugar drying

The last process in raw sugar production is drying (Rein, 2007). Traditionally, raw sugar is dried in a rotary drum dryer tilted at an angle of about 5 ° to the horizontal, to allow continuous flow of the sugar through the drier. In the process, hot air is fed into the dryer counter current

to the sugar flow and by continuously lifting and dropping the sugar crystals, they are cooled and dried. Other types of dryers used for sugar drying are tray dryers and fluidized bed dryers (Seebaluck et al., 2008). A typical sugar mill drying unit is depicted in Figure 2-6.

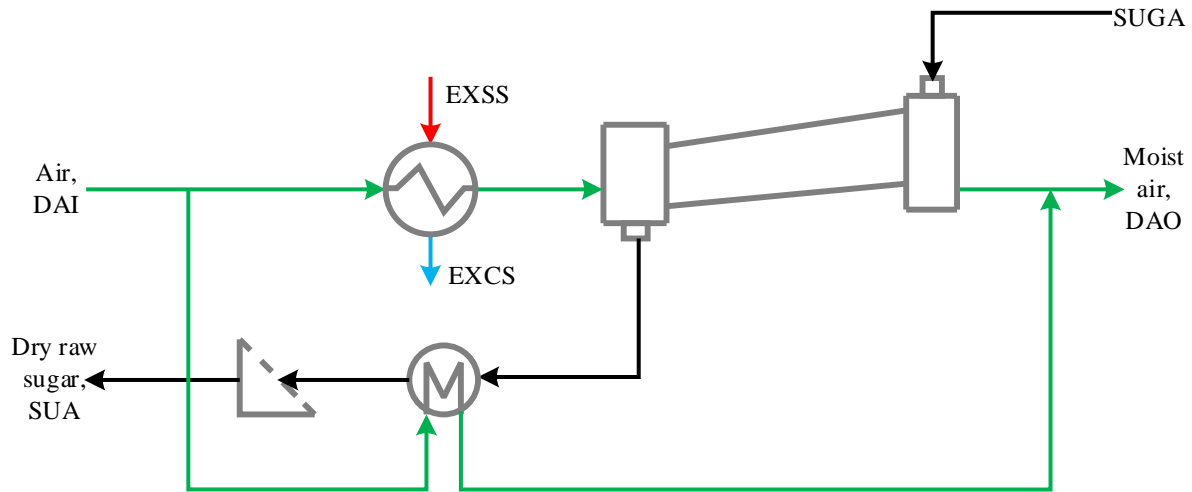


Figure 2-6: Process flow diagram of sugar drying (and cooling) unit (Redrawn based on (Starzak & Zizhou, 2015)). Stream tags are described in the nomenclature.

2.1.6. Water cooling system

Warm water from cooling crystallizers and barometric condensers in the evaporation and crystallization unit are sent into the cooling tower, where a portion of water is evaporated into the air passing through the tower and escapes through the top of the tower. This process provides cooling to the rest of the water that is recycled to the process for cooling purposes. Some cold water may be added to the hot water tank to cater for the evaporative loss (U.S. Department of Energy, 2011). Figure 2-7 illustrates the process flow diagram of the cooling tower.

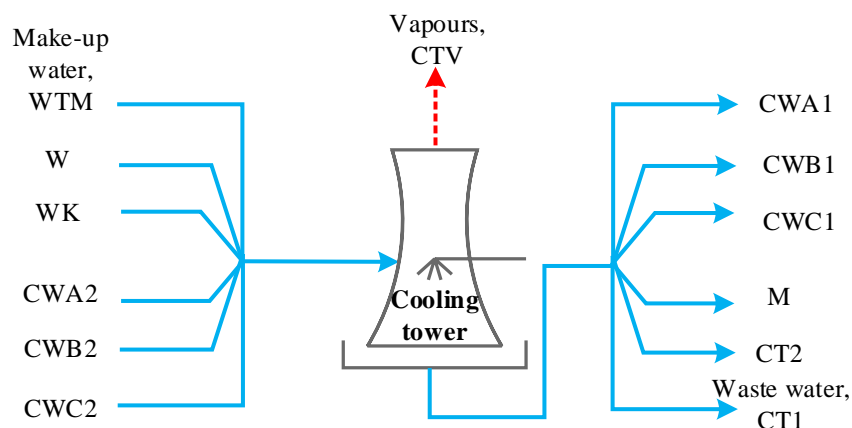


Figure 2-7: Process flow diagram of cooling tower (Redrawn based on (Starzak & Zizhou, 2015)) . Stream tags are described in the nomenclature.

2.1.7. Co-generation

Sugar mill cogeneration system consist mainly of a boiler and turbines as indicated in Table 2-1, and involves burning wet bagasse from extraction unit for simultaneous production of heat and electricity required for the mill (Kamate & Gangavati, 2009a; Seebaluck et al., 2008). Figure 2-8 represents the process flow of the cogeneration system. Conventional sugar mills generate superheated steam (live steam) in low-pressure, inefficient boilers at about 22 - 31 bar and 300 - 390 °C (A. V. Ensinas, Nebra, Lozano, & Serra, 2007; Hofsetz & Silva, 2012; Seebaluck et al., 2008; Starzak & Davis, 2016). The live steam produced in the cogeneration system is exhausted in turbo-alternators at about 2.5 bar to produce electricity for factory use and exhaust steam for the process. The addition of condensates to the exhaust steam de-superheats it to the required steam conditions for the main process. The remaining HP steam drives turbines for knives, shredders and mills in the extraction unit of the mill. A typical South African sugar mill only produces electricity for own use (Conningarth Economists, 2013) and in some cases, a small amount is exported to the national grid. For instance, three Tongaat mills (Maidstone, Amatikulu and Felixton) have cogeneration plants of 72 MW capacity but export only 12% of total production (Mbohwa, 2013). Nowadays, there are various technologies that can utilise high-pressure high-temperature steam of up to 110 bar and 540 °C, respectively from highly efficient boilers, to generate electricity for sale (Kamate & Gangavati, 2009b).

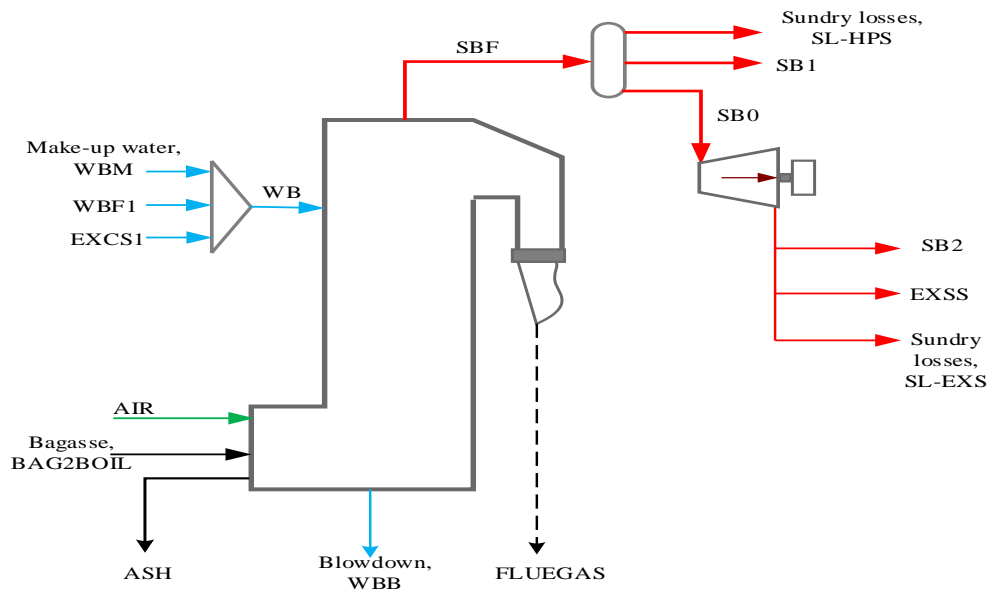


Figure 2-8: Process flow diagram of the cogeneration system (Redrawn from (Starzak, 2016)). Stream tags are described in the nomenclature.

2.2. Energy consumption of sugar mills

Cane sugar processing is energy intensive and mostly, energy inefficient. The cogeneration system supplies the energy requirement of the sugar mill in the form of steam and electricity. Electricity consumption of existing sugar mills ranges between 25 – 32 kWh/ton of cane (Seebaluck et al., 2008). In milling extraction systems driven by electro motors, the electricity demand is about 5 kWh/ton cane less than in diffuser systems (Rein, 2007). Live steam produced in the cogeneration system is exhausted in turbo-alternators (electricity generators) to produce the electricity required. Figure 2-9 shows power requirement and distribution of a typical sugar mill.

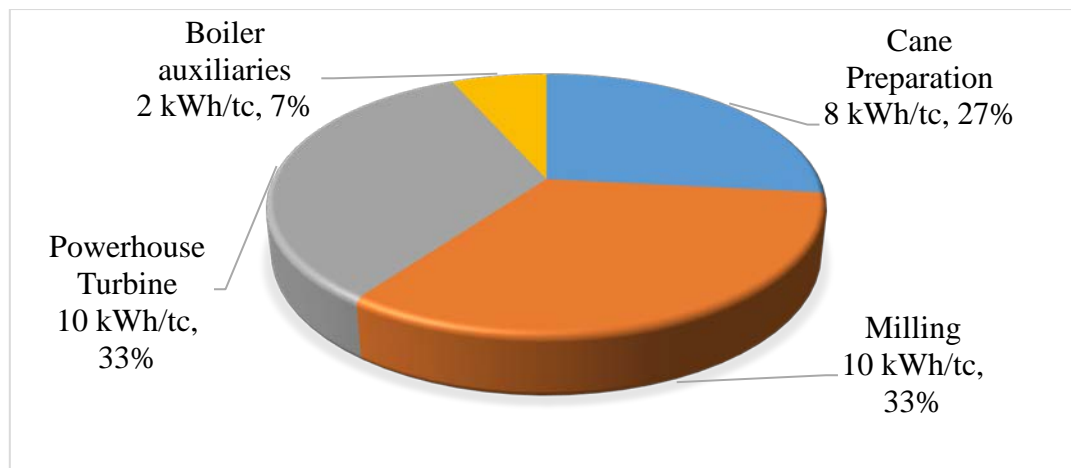


Figure 2-9: A typical Sugar Mill Power Requirement in kWh/tc (Adapted from (Rein, 2007))

South African mills with old inefficient technology use as much as 600 kg steam/ton of cane, while those mills that are considered to be efficient, require about 400 kg steam/ton of cane (S Farzad, Mandegari, & Gorgens, 2015). Figure 2-10 illustrates the steam energy demands of the components of a typical sugar mill. Mill turbines (prime movers) in the cane preparation and extraction systems are driven by live steam from the cogeneration system in the absence of electro motors, which run on electricity (A. V. Ensinas et al., 2009; Rein, 2007). In a well process heat-integrated plant, the exhaust steam from the cogeneration system as well as cane preparation and milling unit's turbines are used mainly to supply heat demands of the first-effect evaporation, as well as the clear juice heater and deaerator. Other heating demands are supplied by vapour bleeds mainly from first and second-effect evaporators (Starzak & Davis, 2016).

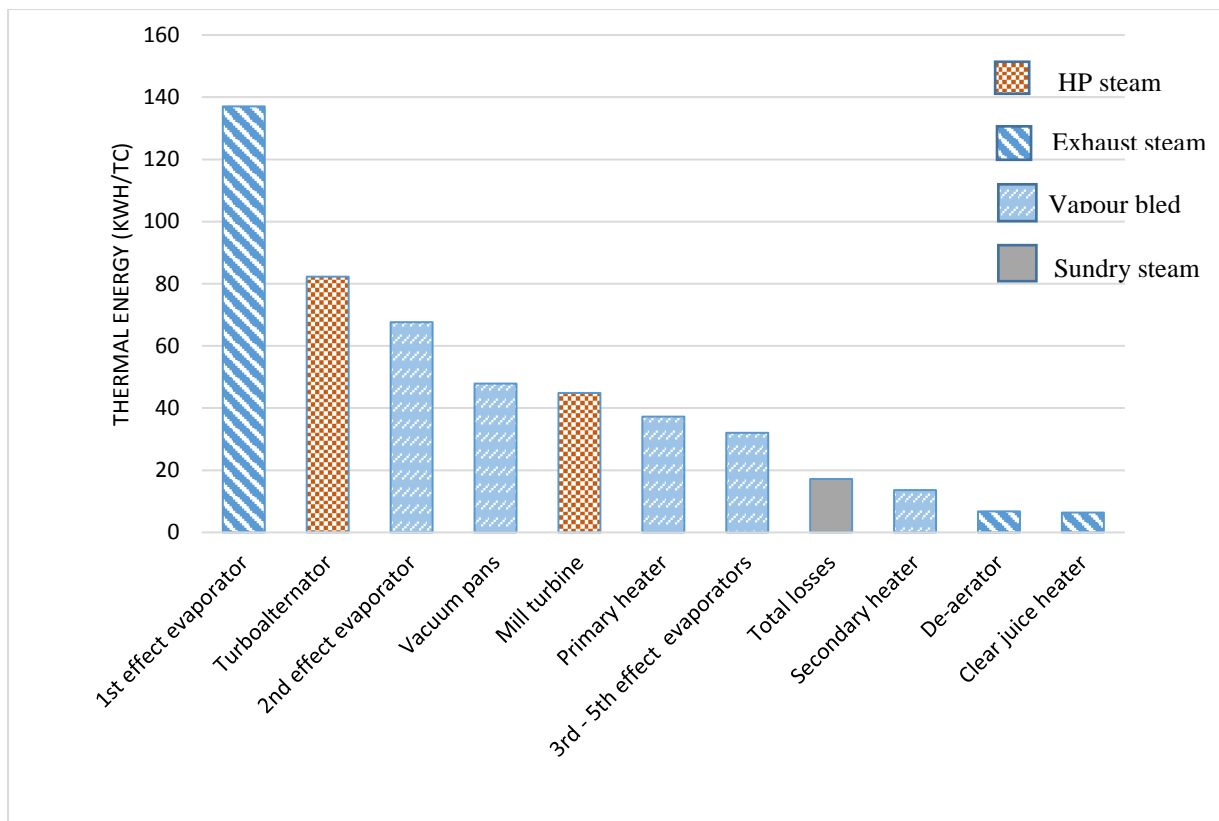


Figure 2-10: Thermal energy demand of a typical sugar mill (Adapted from (Rein, 2007))

The sugar industry is designed to be energy self-sufficient, since under normal, steady-state operational conditions it requires no external energy source. However, some South African factories now burn coal to supplement bagasse due to inefficiencies in process transients and bagasse export for other uses (L Mashoko, Mbohwa, & Thomas, 2013). It is reported that South African sugar mills consume approximately 200 000 tons of coal per season (Hess, Beukes, Smith, & Dinter, 2016). Researchers showed that improving process steam economy could reduce process steam demand to 270 – 300 kg/ton of cane (Ogden, Hochgreb, & Hylton, 1990) to avoid coal burning, while utilizing high-pressure boilers and high-efficiency turbo alternator makes available surplus electricity for sale or surplus bagasse for further valorisation (Kamate & Gangavati, 2009a).

2.3. Exergy and exergoeconomic methodology

Most sugar industry improvement measures to reduce production cost, increase revenue through new products production and improve energy efficiency, are based on conventional energy conservation analysis, through the first law of thermodynamics (Martínez-Guido et al., 2016; Mbohwa, 2013; Ogden, Hochgreb, & Hylton, 1989; Ogden et al., 1990). However, the

first law is limited as it only accounts for the quantity of energy and assigns equal work potential to all energy forms, and hence, not accounting for how the quality of energy is degraded through conversion processes (Kotas, 1995). Exergy analysis which make use of both first and second laws of thermodynamics provide better insight into energy utilization and quantifies the actual thermodynamic performance of processes in order to improve them. Therefore, exergy analysis allows better design of industrial processes, for better preservation of energy quality, while exergoeconomics combines exergy analysis and economic analysis to provide a cost-effective way of developing such designs.

2.3.1. Exergy analysis: basic concepts and theory

Exergy of a thermodynamic system is the maximum theoretical useful work that can be obtained in a reversible process, when the system at a specified state is brought to equilibrium with a reference thermodynamic environment (George Tsatsaronis, 2007). Exergy analysis is a thermodynamic tool that uses both the first and second laws of thermodynamics to identify the locations, magnitude, and causes of process inefficiencies. It is supplementary to the first law-based energy analysis, which only accounts for the quantity of energy flow through a system, and does not show how the quality of energy is degraded through a real system (Wall, 2009). Therefore, exergy analysis gives a truer thermodynamic performance indication of the system than the conventional energy analysis (Rosen 2002), providing insight into the nature of irreversibilities associated with the system (Hinderink, Kerkhof, Lie, & Van Der Kooi, 1996).

Exergy analysis primarily involves a balancing the total exergy input and outs of the system, the difference of which indicate the system irreversibility as expressed by equation (2-1).

$$\sum \dot{E}x_{in} - \sum \dot{E}x_{out} = \dot{I} \quad (2-1)$$

Equation (2-2) expresses input and output exergies in terms of exergies transferred by heat input, work output and material streams entering and leaving the system, resulting in a general form of the steady-state exergy rate balance equation (Modesto et al., 2009) as shown in equation (2-2).

$$\dot{Q} \left(1 - \frac{T_0}{T} \right) - \dot{W} + \sum \dot{m}_{in} ex_{in} - \sum \dot{m}_{out} ex_{out} = \dot{I} \quad (2-2)$$

Where

$\dot{Q} \left(1 - \frac{T_0}{T}\right)$ is the exergy transferred by heat,

\dot{W} is exergy transferred by shaft work, and

$\dot{m}ex$ is exergy transferred by material stream

The exergies associated with a stream of matter include physical, chemical, kinetic and potential exergies, which are results of deviations of the stream temperature and pressure, chemical composition, velocity, and height, respectively, from those of the reference environment (George Tsatsaronis, 2007). The specific exergy of a stream (ignoring kinetic and potential exergies) is given by

$$ex = h - h_o - T_o(s - s_o) + ex_{chemical} \quad (2-3)$$

The specific chemical exergy of a mixture, $ex_{chemical}$ is expressed as

$$ex_{chemical} = \sum_i (\mu_{i0} - \mu_{i00})x_i \quad (2-4)$$

The symbols μ_{i0} and μ_{i00} are the chemical potential of component i in the mixture at the environmental state and dead state respectively, whereas x_i is the mole fraction of component i in the mixture (Şahin, Acir, Altunok, Baysal, & Koçyiğit, 2010).

Alternatively, Querol et al (2013) provided a method to calculate the specific chemical exergy of a stream (mixture) from the specific chemical exergy of its components their molar fractions as expressed in equation (2-5).

$$ex_{chemical} = \sum_i x_i ex_{chemical,i} \quad (2-5)$$

In addition to system irreversibility, which shows exergy destroyed in a system due to entropy generation, thermodynamic performance can also be indicated by exergetic efficiency, which is the ratio of desired products expressed in exergy (product exergy) to the exergy of resources expended to provide the product exergy (fuel exergy) (George Tsatsaronis, 2007), as expressed in equation (2-6).

$$\varepsilon = \frac{\dot{E}x_{product}}{\dot{E}x_{fuel}} \quad (2-6)$$

The exergetic efficiency indicates how close a process is to an ideality (Rosen, 2002; Rosen & Bulucea, 2009; Saidur, Ahamed, & Masjuki, 2010) and may be used to compare similar

components of the same or different system, operating under the same conditions (G Tsatsaronis & Czesla, 1999). For dissimilar components, a ratio of component irreversibility to component resource exergy, known as exergy destruction ratio (E. Querol, Gonzalez-Regueral, Ramos, & Perez-Benedito, 2011) may be investigated.

However, since improving exergy efficiency as well as minimizing irreversibility are limited by technological and economic constraints (Cleveland & Ayres, 2004; Kotas, 1995; Sciubba, 2014), another performance indicator, the exergetic improvement potential is evaluated in equation (2-7) to show how much potential for improvement is possible for a component or a system (Sogut, Ilten, & Oktay, 2010).

$$\dot{I}P = (1 - \varepsilon) \times \dot{I} \quad (2-7)$$

Tsatsaronis & Czesla (1999) reviewed the strengths and limitations of exergy analysis and concluded that exergy analysis becomes a more powerful tool and provides a major contribution through exergoeconomic analysis.

2.3.2. Exergoeconomic analysis

Exergoeconomic analysis, also known as exergoeconomics is a combination of exergy and economic analysis methods to provide information for cost-effective plant design and operation. In addition to mass, energy and exergy balances accomplished by the conventional exergy analysis, exergoeconomics involves assigning monetary cost to exergy of material and energy streams of a plant or plant component (classified as fuel, product or loss exergies), which is termed exergy costing, taking into account the investment cost for each plant component (George Tsatsaronis, 2007; Georgios Tsatsaronis & Winhold, 1985). Therefore, exergoeconomics evaluates both process inefficiencies and costs of inefficiencies, which are compared with investment cost required to reduce the inefficiencies (G Tsatsaronis & Czesla, 1999). Moreover, exergoeconomics help to assess the effective use of resources in order to save them (Atmaca & Yumrutaş, 2014a). Evaluation for exergy cost for each stream as well as the formulations cost balance equations are described in the literature (Atmaca & Yumrutaş, 2014a; Lazzaretto & Tsatsaronis, 2006; E. Querol et al., 2011).

The indication of a system's exergoeconomic performance in analysis can be shown by three useful parameters; total cost rate, exergoeconomic factor and relative cost difference (Aghbashlo, Tabatabaei, Jazini, & Ghaziaskar, 2018). Total cost rate, \dot{C}_T of each component is the summation of component capital investment- and exergy-related cost. The component with

the highest \dot{C}_T is prioritised for optimization. Equation (2-8) expresses the evaluation of the total cost rate,

$$\dot{C}_T = \dot{Z} + \dot{C}_D = \dot{Z} + c_{fuel} \cdot \dot{E}x_D \quad (2-8)$$

where

\dot{Z} is the capital cost rate (US\$/s),

\dot{C}_D is the cost rate of exergy destruction (US\$/s)

c_{fuel} is the unit exergetic cost of resources or fuel (US\$/kJ) and

$\dot{E}x_D$ is exergy destruction rate (the same as irreversibility rate, \dot{I}) (kW).

Exergoeconomic factor of a component, f expresses the contribution of the component capital cost to the sum total of the capital cost and cost of exergy destruction (Aghbashlo, Tabatabaei, Soltanian, Ghanavati, & Dadak, 2019; Atmaca & Yumrutaş, 2014a) as shown in equation (2-9).

$$f = \frac{\dot{Z}}{\dot{Z} + \dot{C}_D} \quad (2-9)$$

A high value of f indicates the need to reduce equipment cost even if that reduces exergetic efficiency. On the other hand, a low factor implied a better or more expensive equipment could be employed in order to increase exergetic efficiency (Atmaca & Yumrutaş, 2014a; E. Querol et al., 2011).

The second exergoeconomic performance indication parameter, relative cost difference (r) indicates the increase in unit exergoeconomic cost of products compared to the unit cost of resources used in each component, as expressed in equation (2-10). The relative cost difference is instrumental in ranking component with high r values for optimization (Atmaca & Yumrutaş, 2014a; E. Querol et al., 2011).

$$r = \frac{c_{product} - c_{fuel}}{c_{fuel}} \quad (2-10)$$

2.3.3. Exergy Calculations in Aspen Plus

Aspen Plus version 8.6 has the following prop-set properties, EXERGYFL (exergy flow rate), EXERGYML (exergy on a mole basis) and EXERGYMS (exergy on a mass basis) for evaluating only physical exergy of a stream simulated. Jacobs Consultancy in Leiden, The Netherlands has developed a plug-in software routine for Aspen, ExerCom, to help determine

total material stream exergy, consisting of chemical, physical (and mixing) exergies (Gourmelon et al., 2015; Jacobs Consultancy, 2015). However, ExerCom is only applicable for gases and liquids, and not solids and user-defined components in Aspen. Besides, ExerCom does not include calculation of exergy of work and heat streams, and is, therefore, not generally adequate to support exergy analysis in Aspen.

Alternatively, all exergy components can be calculated in an Excel spreadsheet using physical exergy data generated in Aspen simulation for all streams and the remaining data sourced from thermodynamic tables (Mahamud, Khan, Rasul, & Leinster, 2013). Exergy of heat streams can also be calculated by application of relevant equations. Aspen stream results are updated in Excel using Aspen simulation workbook in Aspen V8.6 (Dogbe, Mandegari, & Görgens, 2018; Finlayson, 2015).

2.3.4. Exergy assessments of sugar mills

In the present study, the actual energy use of the sugar mill, as described before, is assessed through a second law of thermodynamic analysis to reveal process inefficiencies beyond the energy balance obtained by conventional energy analysis. This assessment, called exergy analysis is necessary for the improvement of the sugar mill energy efficiency and cost-effective production.

Relatively few studies on the application of exergy analysis in the sugar industry have been reported in literature for different sugar mill configurations, particularly for beet sugar production and integrated sugar and ethanol plants. Tekin and Bayramoğlu (1998a) studied exergy analysis of a beet sugar production plant in Turkey. Their results showed that the steam-power plant, which produces steam at a pressure of 29 atm and a temperature of 390 °C, is responsible for 74.4 % of exergy losses due to the irreversibility of the combustion process and exergy wastes of the stack gases. Also, Bayrak et al (2003) and Sahin et al (2010) did separate works on the exergy analysis of beet sugar processes in Turkey and found that the raw juice production (sherbet production) and juice crystallization systems had the lowest exergy efficiencies. However, the studies did not include the cogeneration (steam-electricity) system in the analysis. The most recent exergy analysis study on beet sugar plant focused on optimizing the cogeneration unit with 37 bar, 427 °C boiler, which was the most inefficient unit of the plant with an exergy efficiency of 27.7 % (Taner & Sivrioglu, 2015). Their result showed an improvement on the exergy efficiency to 31.7 %, by increasing the power of the unit.

Other sugar mill exergy studies include that of a typical sugar and ethanol plant in Brazil (A. V. Ensinas et al., 2009). The processes for the integrated sugar and ethanol production from cane includes fermentation, distillation, condensate tank and water cooling systems in addition to the five main stages mentioned earlier. It was revealed that 63 % of the total irreversibility generated was from the cogeneration system, with low efficiency boiler that generated steam at 22 bar and 300 °C. The exergy efficiency of the whole plant was 35 %. Similar work was done by Velasquez et al (2013) using a Colombian cane sugar and ethanol plant as a case study. They also obtained similar result for the global plant exergy efficiency (35.1 %) and identified the highest point of irreversibility as the cogeneration system with even much lower exergy efficiency of 13.9 %. Obviously, this result can be attributed to the low-pressure, low-temperature boilers used, though the parameters were not specified.

With the exception of the work of Garg et al. (2016) who evaluated the exergy losses and irreversibilities of only the juice production and clarification units because of their large exergy losses, there is no exergy analysis of raw sugar production from sugarcane (as in South African sugar mills) found in the literature, and therefore, the focus of this study.

2.4. Technologies to improve sugar industry energy/exergy efficiency

Currently, the South African sugar industry is faced with energy inefficiency due to the age of the factories. This study seeks to locate and quantify the inefficiencies of the production system through exergy analysis. Since irreversibility, the principal contributor to inefficiency, is a function of the process, it is insightful to consider alternative processes with less irreversibility, in order to improve energy/exergy efficiency. Research shows that sugar mill (beet sugar) inefficiencies from the exergetic point of view, mostly can be attributed to the high irreversibility of cogeneration unit (especially the boiler), high input exergy requirement in the form of steam and exergy losses through waste streams (especially, the stack gases). In this section, some energy-efficient technologies, novel to the South African sugar industry are reviewed for possible implementation in order to better the exergetic performance of the industry.

2.4.1. Technologies for improving cogeneration

Previous research showed that over 70 % of sugar mill inefficiencies, from the exergetic point of view, are attributed to the high irreversibility of cogeneration unit, especially the boiler

(Héctor et al., 2013; T Tekin & Bayramoğlu, 1998a). Cogeneration units with highly efficient boilers (about 61 bar and 475 °C) was reported to provide optimal exergy efficiency compared to lower-pressure boilers (Kamate & Gangavati, 2009b). Due to the process steam constraints, high-efficiency high-pressure boilers in the sugar industry are usually accompanied with advanced cogeneration systems to exhaust high-pressure steam for electricity production, which is a high exergy product rather than to let down steam irreversibly. These advanced technologies include condensing extraction steam turbine (CEST), biomass integrated gasification combined cycle (BIGCC) and supercritical steam cycle (SuSC).

2.4.1.1. Condensing extraction steam turbine (CEST)

Condensing extraction steam turbine (CEST) is the commonest cogeneration system found in modern sugar mills. This steam power cycle is accompanied by high-pressure boilers and has the advantage of producing more electricity (high exergy product) per unit bagasse than the back-pressure steam turbines (BPST) currently used in South African mills (Alonso-Pippo, Luengo, Felfli, Garzone, & Cornacchia, 2009). Unlike the BPST, where all the steam is expanded to the required process steam pressure of about 2 bar, the CEST system is able to extract only the required amount of process steam at this pressure and the rest is expanded to below atmospheric pressure to generate extra electricity. Its unit operating cost is however high (Deshmukh, Jacobson, Chamberlin, & Kammen, 2013). The process diagram for typical CEST technology for a sugar factory is illustrated in Figure 2-11. Condensing Extraction Steam Turbines are bigger than BPSTs, so they are not recommended for small sugar plants of up to 700 tons cane/day capacity. Moreover, in an exergy studies of a refined sugar plant (Kamate & Gangavati, 2009b), both the first law and second law efficiencies of CEST were lower than those of the BPST, for the same steam conditions mainly due to waste heat rejected to the environment through condensation unlike BPST. Therefore, the CEST technology is only desirable for purposes of surplus electricity generation for sale, which is a less interesting option for the South African sugar industry (Dr K Foxon 2016, SMRI, pers. comm., 21 September).

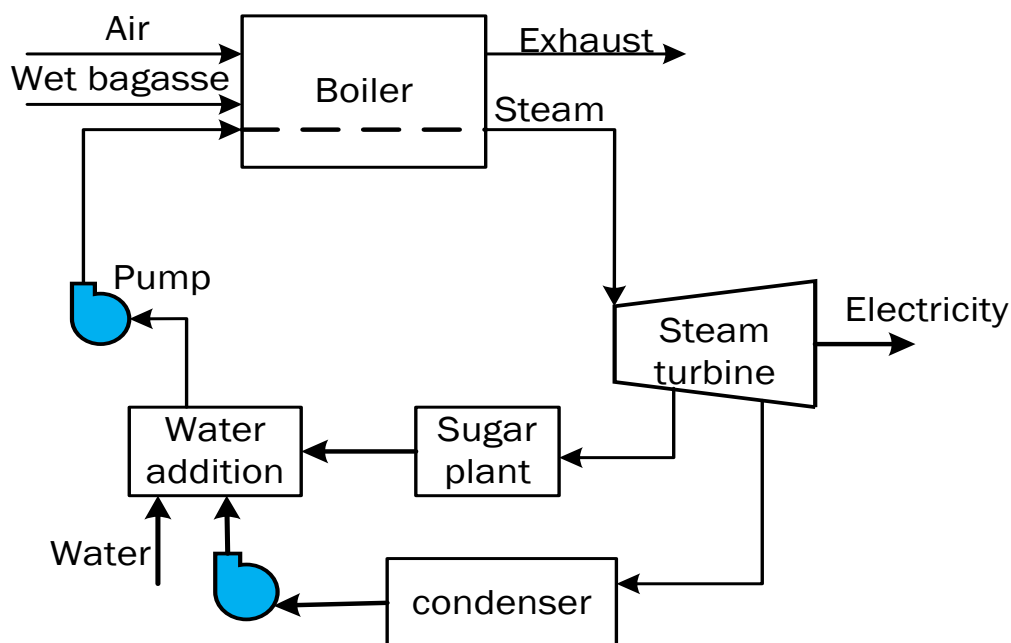


Figure 2-11: Steam Rankine cycle (CEST) cogeneration system for a sugar factory (Redrawn from (Deshmukh et al., 2013))

2.4.1.2. Biomass integrated gasification combined cycle (BIGCC)

The biomass integrated gasification combined cycle (BIGCC) mainly consists of a gasifier, a gas turbine, heat recovery steam generator (HRSG) and a back-pressure steam turbine as shown in Figure 2-12, for a sugar mill. The gasifier converts bagasse to combustible gasses, used to fuel a gas turbine to generate electricity. Then, the HRSG uses the thermal energy of the gas turbine exhaust gases to generate high-pressure steam for the steam turbine. The steam turbine produces electricity and exhaust steam for the sugar mill process (A. V Ensinas et al., 2006; Larson, Williams, & Leal, 2001). BIGCC allow the generation of electricity of about three times that of steam cycles. However, its application is only possible when process steam demand is significantly reduced to about 280 kg/ton of cane (Pellegrini, de Oliveira Jr, & Burbano, 2010). Moreover, the cost of installing BIGCC is very high (Mbohwa, 2013) and even though it is a commercial technology, there is no commercial BIGCC in the sugar industry at present (Deshmukh et al., 2013; Matelli, Perrella Perrella, & Fonseca Filho, 2016).

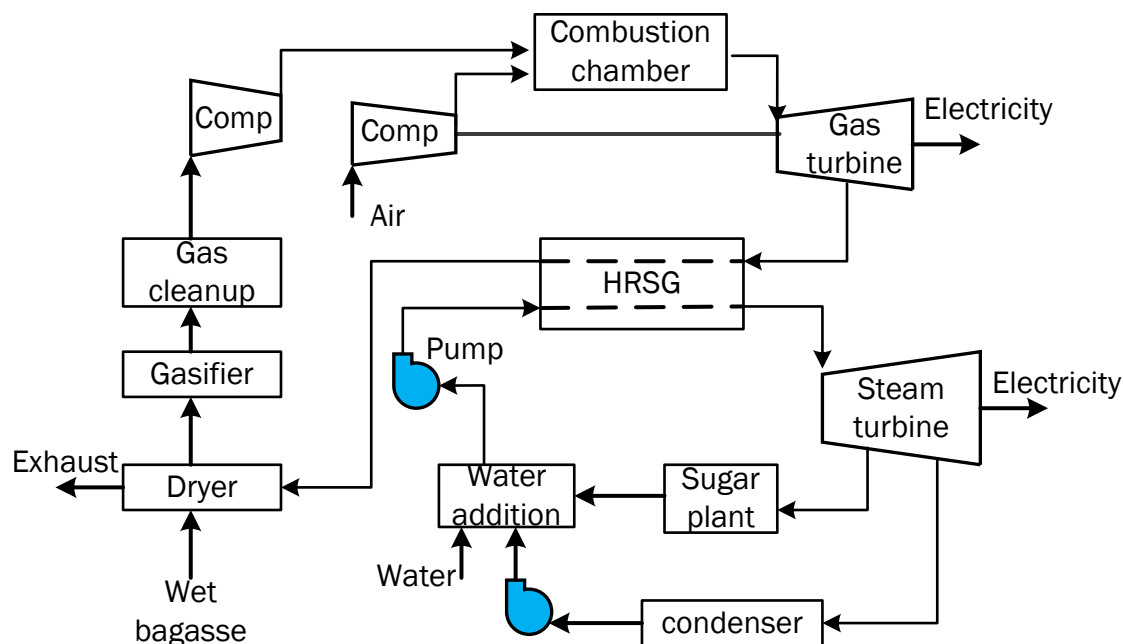


Figure 2-12: BIGCC cogeneration system for a sugar factory (Redrawn from (Deshmukh et al., 2013))

2.4.1.3. Supercritical steam cycle (SuSC)

Supercritical steam cycle (Figure 2-11) is an advanced steam Rankine cycle with higher boiler design pressure, and hence a higher boiling temperature. The boiler used in a supercritical steam cycle is designed without a generating bank (steam drum), because the system operates above 221 bar (the critical pressure of water) and saturation temperature of 374 °C, so that the separation of water and steam is not necessary (CIAB, 2010), thus minimizing the irreversibility of the process. The technology is used in coal power plants and may soon become commercially available for the sugar industry. The SuSC is however, not suitable for small capacities and would be better installed for a pool of mills rather than inside a single mill, when its implementation is realised. The reason is that a smaller SuSC with a generating capacity of 280 MW would require the crushing of at least 6.5 million tons of cane per year, which is far more than the crushing capacity of an average mill (Pellegrini et al., 2010).

2.4.2. Waste heat recovery technologies

Aside from irreversibility of processes, such as combustion in the cogeneration system of the sugar industry, unutilised low grade heat carried away by process waste streams also contributes to process inefficiency (C. W. Chan, Ling-Chin, & Roskilly, 2013). The exergy in these waste streams can be recovered by waste heat recovery technologies for heating, cooling

and power generation purposes, or combination of all these, in the production process. In the raw sugar industry, the exergy in waste streams such as flue gas from stack, boiler blowdown and flash/vent steams are available for recovery. In the following sections, waste heat recovery technologies such as thermodynamic cycles and absorption heat pumps are discussed for possible integration into the South African Sugar mills towards efficiency improvement.

2.4.2.1. Absorption heat pump (AHP)

Heat pump is a device that transfers heat from a low-temperature medium to a higher-temperature one through the circulation of a working fluid between an evaporator and a condenser (Bakhtiari, Fradette, Legros, & Paris, 2010). Conventional heat pumps (vapour compression heat pumps) demand high electric power to provide mechanical work for its compression component. Alternatively, absorption heat pumps use heat instead of mechanical work, use natural refrigerant medium and have quiet operation due to few moving parts (Carrier, n.d.). Thus, absorption heat pumps are more environmentally friendly than the former. Additionally, absorption heat pump is a proven and sustainable technology for low grade heat recovery and utilization. The working fluid of the absorption heat pump is an absorbate - absorbent pair with high heat of absorption. Most commonly and commercially available pairs are water–lithium bromide (water as absorbate) and ammonia–water (ammonia as absorbate) (C. W. Chan et al., 2013); the former is a well-established technology for chilled water production for comfort air conditioning (Keil, Plura, Radspieler, & Schweigler, 2008).

Figure 2-13 illustrates the working principle of the absorption heat pump with its major components; evaporator, absorber, regenerator and condenser. In the absorption heat pump cycle, the absorbate (refrigerant) is evaporated at low pressure in the evaporator, then is absorbed by the absorbent solution in the absorber to form the absorbate–absorbent solution (the working pair), giving off the heat of absorption. The working pair is pumped to the regenerator where the refrigerant is boiled off by a heat source (waste heat). In the condenser, the separated refrigerant is condensed into liquid, which returns to the evaporator to complete the cycle (Keil et al., 2008). The absorption heat pump is used for both heating and cooling purposes - useful heat is released in the absorber and the condenser and the evaporation of the absorbate in the evaporator produces cooling effect. Waste heat can be used as the heat source of the evaporator, or the evaporator can be used for cooling purposes.

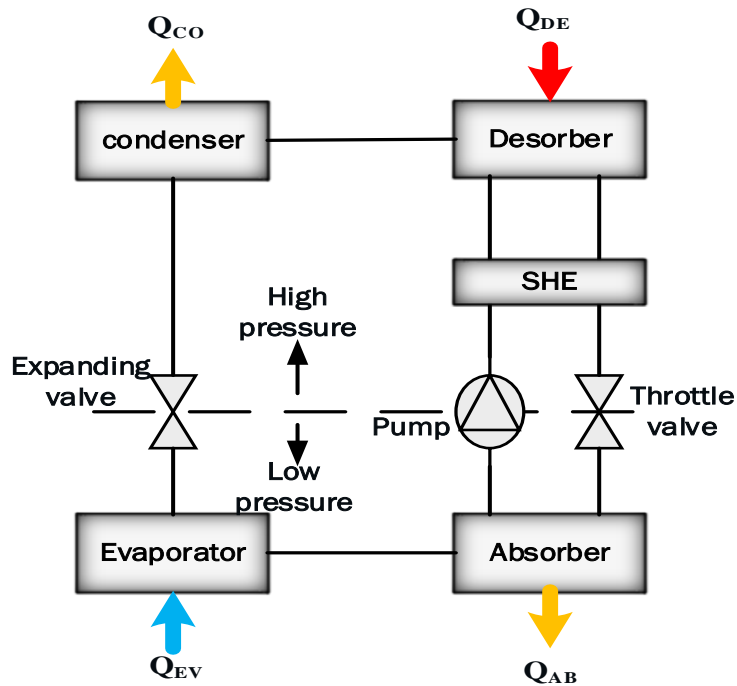


Figure 2-13: Absorption heat pump (Redrawn from (Ibarra-Bahena & Romero, 2014))

The absorption heat pump can be used as a stand-alone equipment, providing heating and cooling, or combined with cogeneration system as a combined cooling, heating and power system (CCHP). Although, electrical-driven refrigeration (or cooling), as in air conditioners, is more energy efficient than heat-driven (absorption) refrigeration, the use of waste heat in the later increases its efficiency (Zhang, Yang, Lu, Shan, & Zhu, 2011) and provides the avenue to recover energy/exergy which otherwise, would have been wasted.

2.4.2.2. Organic Rankine cycle (ORC)

Organic Rankine cycle, like steam cycles is a thermodynamic cycle used to convert heat to electricity. Unlike the steam Rankine cycle, organic Rankine cycle uses organic fluids with high molecular weight and low boiling point as working fluid, instead of water. The use of such organic fluids allow recovery of heat from low temperature (70 – 300 °C) heat sources (mainly, waste heat) which cannot be recovered in conventional Rankine cycle (Chen, Goswami, & Stefanakos, 2010; Eyidogan, Canka, Kaya, & Coban, 2016). Figure 2-14 shows the working principle of the organic Rankine cycle. With only a little energy input, the working fluid is first pumped from low pressure (1) to high pressure (2) into a heat recovery generator where the fluid is vaporized at constant pressure (3). The enthalpy of the vapour is converted to work as it expands in an expander, hence, its pressure and temperature reduces (4). The low-pressure

fluid then enters the condenser to be cooled and forms a saturated liquid (1), and a new cycle begins.

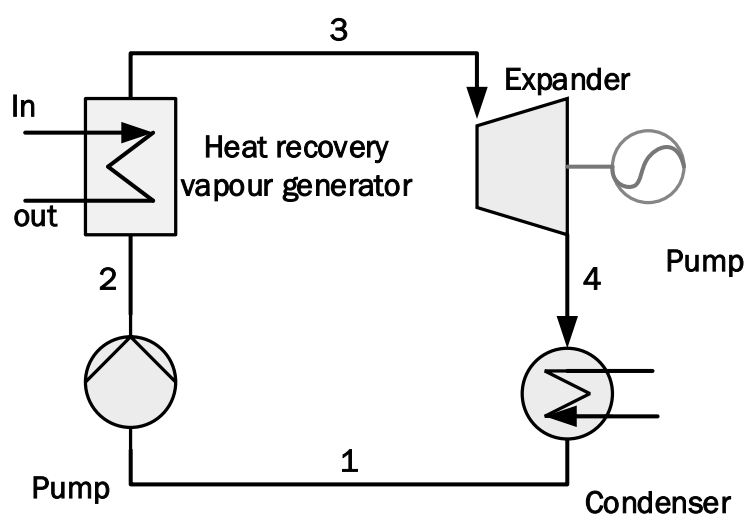


Figure 2-14: Organic Rankine cycle for low heat recovery (Redrawn based on (Lecompte, Huisseune, Van Den Broek, Vanslambrouck, & De Paepe, 2015))

Other thermodynamic cycles for waste heat recovery include supercritical Rankine cycle (SRC), the trilateral cycle (TLC) and the Kalina cycle (KC), which have similar working principles as the ORC. The working fluid of the KC is made up of two fluids with different boiling points (Kumar & Karimi, 2014). For the TLC, expansion starts from a saturated liquid phase instead of a vapour phase, whereas in the SRC, the expanding fluid is a supercritical vapour (a vapour with pressure above its critical pressure). However, among these waste heat recovery thermodynamic cycles, only ORC is commercially available (C. W. Chan et al., 2013). Although the conversion efficiency of thermodynamic cycles is reported to be very low, they are better options compared to direct use of waste heat for heating and cooling purposes (Kumar & Karimi, 2014).

2.4.3. Technologies to reduce process steam demand

In addition to minimizing process irreversibility and exergy loss through waste, another means of improving thermodynamic efficiency of a system is to reduce energy input of the system as much as possible. The process steam consumption of South African sugar mills is about 550 to 600 kg per tonne of cane (Meyer et al., 2011) and is similar to old sugar mills in other parts of the world (Birru, Martin, & Erlich, 2016), which is a contributor to the inefficiency of the industry (Dr K Foxon 2016, SMRI, pers. comm., 8 August). The evaporation system is one of

the most energy intensive processes in raw sugar production, and hence the dominant user of process steam, even with the application of heat integration. Since heat integration is applied in the sugar industry to a large extent (Starzak & Zizhou, 2015), there are limited opportunities for process improvement in the evaporation section in this regard.

However, application of alternative processes for evaporation, such as reverse osmosis and falling film evaporators, have been suggested in literature (Tekin & Bayramoğlu 1998a; Clarke 1999; Bocci et al. 2009; Gurumurthy 2011; Ensinas et al. 2009; Gul & Harasek 2012) to further reduce process steam demand. These technologies are used either in place of the conventional Robert evaporators or to pre-concentrate the sugar juice prior to evaporation. Also the studies regarding the capability of gas hydrate for separating water and concentrating liquid foods (Boch Andersen & Thomsen, 2009; Li et al., 2014; Purwanto, Oshita, Seo, & Kawagoe, 2001) suggest that it is possible to use the technology in concentrating sugar juice, as is further discussed in the following subsections. However, more experimental and economic studies are needed to establish the relevance and sustainability of the gas hydrate and reverse osmosis technologies in the sugar industry, and subsequently, the assessment of their potential to improve exergy efficiency of the industry.

Apart from the evaporators, exergy losses in the juice extraction system are also attributed to high energy consumption of low efficient steam turbines used to drive knives, shredders and bagasse mills. Research shows that replacing these turbines with electric motor results in 2% reduction in irreversibility (A. V. Ensinas et al., 2009), since the electricity needed by the motors is generated in more efficient turbines (Pellegrini & de Oliveira Jr, 2011).

2.4.3.1. Gas hydrate technology (GHT)

Gas hydrates are crystalline solid structures made up of geometric lattices of water molecules with cavities filled by small molecules such as CO₂, N₂, H₂ and light hydrocarbons. Although they are noted for negative impact in the oil and gas processing industry, gas hydrates have positive application for novel separation processes such as gas separation, storage and transportation of natural gas, desalination and carbon dioxide capture (Rajnauth, Barrufet, & Falcone, 2013). In line with recent emerging interest in industrial applications of gas hydrate formation, Eslamimanesh et al (2012) have reported an experimental review on the aforementioned processes and others, including concentration of dilute aqueous solution.

With regards to concentration of liquid foods, various experimental investigations were done using coffee solution as the liquid food and xenon gas (Purwanto et al., 2001; Purwanto, Oshita,

Seo, & Kawagoe, 2014) and the concentration of orange juice by the formation of C_2H_4 hydrate (Li et al., 2014). Boch Andersen and Thomsen (2009) studied sugar juice concentration by formation of CO_2 hydrate. They concluded that gas hydrate technology was not suitable for sugar concentration based on their experimental results. The reasons for this conclusion were the large volumes of water in sugar juice and the high pressure required for the hydrate formation process in order to separate such quantity of water from the juice. It is important to establish the practicality of the gas hydrate technology in the sugar industry (with mass and energy balances in place) as the basis for its assessment through exergy analysis.

2.4.3.2. Reverse osmosis (RO)

Membrane separation processes, with no phase change are emerging technologies desired to replace energy intensive conventional separation processes such as evaporation. Reverse Osmosis (RO) is a commercially available membrane separation technology for seawater desalination. It has the advantage of simplicity and low cost over its counterparts, for example, multi-stage flash distillation (Khawaji, Kutubkhanah, & Wie, 2008).

Some researchers (Gul & Harasek, 2012; Madaeni & Zerehski, 2010) suggested RO as a pre-concentration process prior to evaporation in the beet sugar industry and demonstrated its technical feasibility. It was found out that between 30 – 70% of energy is saved subject to the number of stages and arrangement of membrane. The energy savings is due to reduction in the load to the pre-heaters and evaporators and hence, reduction in steam demand. Figure 2-15 (a) shows a conventional multistage RO pre-concentration process. An improved arrangement as depicted in Figure 2-15 (b), is the novel multistage RO used by Gul and Harasek (2012) to pre-concentrate beet sugar thin juice. This arrangement was found to concentrate thin juice from a Brix degree of 15 to 50, removing more than 80% of water from the juice before evaporation.

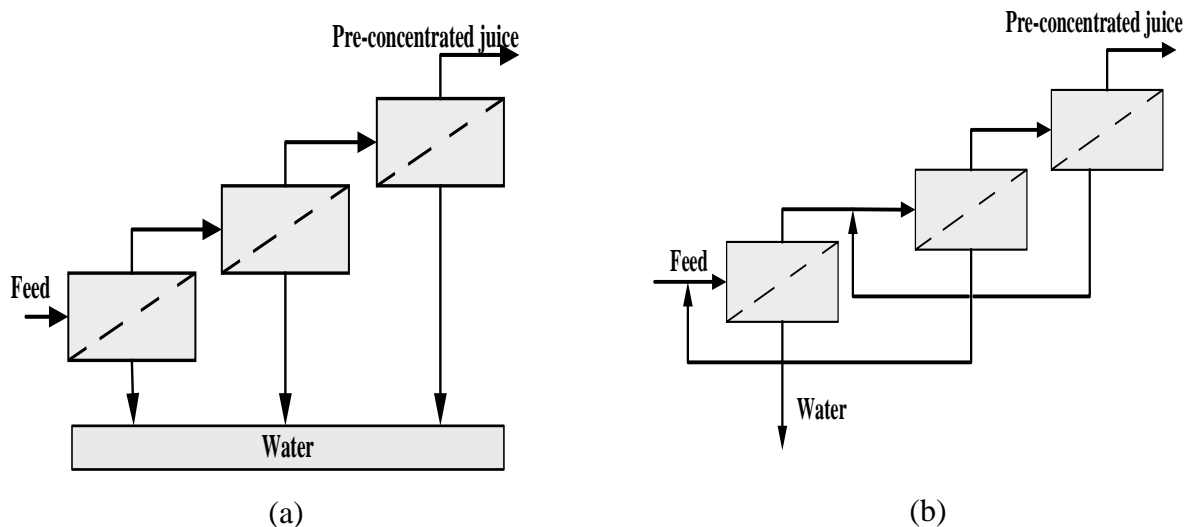


Figure 2-15: RO Process (a) Conventional. (b) Novel (Redrawn from (Gul & Harasek, 2012)).

However, similar to gas hydrate technology, the use of membrane technology in the sugar industry, especially for thin juice concentration, has not been demonstrated commercially due to several setbacks, including (a) thermal creeping within the membrane modules, and (b) high osmotic pressure in the membranes due to large volumes of concentrated thin juice (Hinkova et al. 2002). Consequently, in their study on the potentials of separation membrane in the sugar industry, Hinkova et al (2002) stated that the interest in membrane separation should be focused more on purification of raw juice and direct production of white sugar than concentration of thin sugar juice.

2.4.3.3. Falling film evaporators (FFE)

The application of falling film evaporators (FFE) in the beet sugar industry is widely known and accepted because of its energy saving capability, among several other advantages. However, its implementation in the sugar-cane industry is not well established, although few installations were reported in literature (Coustel & Journet, 2009; Lehnberger, Brahim, & Mallikarjun, 2014) following some progress in works to overcome its setbacks in the industry. All the same, the falling film evaporator has several advantages over the Robert evaporators currently being used in the sugar-cane industry, especially due to very low retention time and reduced process steam requirement (Lehnberger et al., 2014).

Falling film evaporators (FFEs) can be used in combination with conventional Robert evaporators to form the multi-effect evaporator system in which case, FFEs are usually the first or second-effect evaporator. This configuration was used in a cane flexi-factory in Mauritius (Khoodaruth, 2015) and two factories in Reunion Island (Coustel & Journet, 2009). On the

other hand, it is possible to replace the entire evaporator system currently used (Robert evaporators) with FFE. Lehnberger et al (2014) reported the installation and performance of a five-effect FFE in ICPL, a sugar-cane factory in India. They indicated that the process steam consumption of sugar production per ton of cane processed using five-effect falling film evaporators is 300–330 kg against about 400 kg that is currently used under steady state operation (Starzak & Zizhou, 2015). Even though, FFE is desirable for minimum steam consumption, challenges with its use including complex operation and control, higher cost and higher susceptibility to scaling than Robert evaporators remain a barrier to its implementation in the South African sugar industry.

2.5. Sugarcane biorefinery

A biorefinery is a facility that utilises conversion processes and equipment for the sustainable production of marketable bio-based products (food, feed, chemicals, materials) and bioenergy (including biofuels, electricity, heat) from biomass (Cherubini, 2010; de Jong, Higson, Walsh, & Wellisch, 2012; van Ree & de Jong, 2013). The fundamental aim of the biorefinery concept is a design for maximum profit and minimum energy use, and consequently, minimum environmental impact (Cherubini, 2010; Giuliano, Cerulli, Poletto, Raiconi, & Barletta, 2016). Several factors drive the trending sugarcane biorefinery concept. Firstly, only 17 % of a typical sugarcane plant is converted to the industry's main product, sugar (Moya et al., 2013). Furthermore, the remaining consists of by-products bagasse and molasses, as well as filter cakes, ash and the plant residue (tops, leaves and trash) are considered as waste products of the industry (Livison Mashoko, Mbohwa, & Thomas, 2010), in addition to surplus water. Bagasse, the most important sugar industry by-product is used as the primary fuel to generate steam and electricity for the industry, while molasses is used as an animal feed additive and for industrial production of beverage alcohols and other fermentation products (Eggleston & Lima, 2015). Currently, research efforts are being intensified towards producing value-added bio-products using both bagasse and molasses as economical feedstocks in a biorefinery concept.

Secondly, bio-based chemicals production has clear environmental advantages over petroleum-based ones due to environmental concerns over climate change associated with the use of fossil resources, which are limited in reserve (Ferone, Raganati, Olivieri, Salatino, & Marzocchella, 2017; Krajnc & Glavič, 2009; Vaswani, 2010). Over the years, the sugar industry, especially in Brazil (which is the largest in the world) has evolved from sugar only production to a combination of products – sugar, ethanol and electricity (Eggleston & Lima, 2015). Recently,

chemicals from sugarcane biomass have also received considerable research and commercial attention. However, there is much more to explore regarding modelling, energy and economic assessments aspects to establish the cost-effectiveness of the concept. Succinic acid has been identified as one of the 12 chemicals with near-term potential and increasing market (Biddu, Scarlata, Biddu, & Scarlata, 2016). In another vane, the increasing need for healthy food supplements present scFOS as another valuable sugarcane biorefinery product to consider.

2.5.1. Succinic acid (SA)

The potential of succinic acid as a platform chemical for a wide range of commercial products has made it an attractive biorefinery product to study.

2.5.1.1. Product description and market

Succinic acid (SA), also known as butanoic acid, is a four-carbon dicarboxylic acid naturally produced from plant, animal and microorganism metabolisms (Zeikus, Jain, & Elankovan, 1999). Industrially, bio-SA is produced by microbial fermentation of plant biomass sugars (McKinlay, Vieille, & Zeikus, 2007; Taylor et al., 2015). Succinic acid (SA) is widely used in the food, pharmaceuticals, polymers, paints, cosmetics and inks industries and a potential platform chemical for more than thirty (30) commercially important products (Barros et al., 2013; Vaswani, 2010). The global succinic acid market in 2011 was estimated at 40 000 tonnes per year and was projected to grow to 600 000 tonnes by 2020 (Klein et al., 2017). The main driver for the high rate of market growth is the potential of SA as a platform chemical to replace maleic anhydride for the production of bio-butanediol (BDO) (Klein et al., 2017). A wide range of feedstocks is available for the production of bio-succinic acid to meet the market demand.

2.5.1.2. Feedstock

Common feedstocks for large-scale fermentation production of succinic acid include starch, beet or cane molasses and refined sugars such as glucose, sucrose, and fructose (Jansen & van Gulik, 2014). Even though glucose is the most commonly used feedstock (Vaswani, 2010), current research efforts have been directed to utilizing cheaper feedstocks such as molasses and lignocellulosic residues (Borges & Pereira, 2011; S. Chan, Kanchanatawee, & Jantama, 2012; Jiang et al., 2013; R. Liu et al., 2013), considering that feedstock cost contributes the most to the manufacturing cost of biorefineries (Jiang et al., 2017). Both molasses and lignocellulosic residues are sugarcane resources, making the sugar industry a key player in the expansion of

the bio-succinic acid market. However, with such cheap feedstock, the SA purification deserves close attention (Efe, van der Wielen, & Straathof, 2013).

The use of molasses and lignocelluloses for a cost-effective succinate production gave comparable results with pure sugars. Chan et al. (2012) showed that fermentation 150 g/L molasses gave succinate titer, yield and productivity of 55.8 g/L, 0.96 g/g and 0.77 g/L/h at 72 h incubation compared to 53.4 g/L, 0.90 g/g and 0.74 g/L/h for the fermentation of 70 g/L sucrose at 48 h incubation in batch fermentation; both by a genetically modified *Escherichia coli* strain (KJ122-pKJSUC-24T). Their results indicated a slower succinic production in molasses than in pure sucrose, which was attributed to impurities in the molasses (S. Chan et al., 2012). Similarly, Ma et al. (2014) reported a slightly lower succinate titer (36.3 g/L) from the fermentation of molasses compared to sucrose (41 g/L) using engineered *E. coli* strain AFP111/pTrcC-cscA at 34 h in 3-L fermenter during dual-phase fermentation. Liu et al. (2008) reported that pre-treating the molasses to remove impurities before fermentation improved succinate production and also help with SA purification.

On the other hand, lignocellulosic feedstocks require expensive pre-treatment and hydrolysis steps to make the sugars present in them easily accessible. The pre-treatment step also yields various sugar degradation products that inhibit succinate fermentation (Jansen & van Gulik, 2014; Nieder-Heitmann et al., 2019) and may be carried through to SA product to complicate its purification. Detoxification is, therefore, carried out with activated carbon, among other processes to purify the sugars (R. Liu et al., 2013; Pateraki et al., 2016; Zhao et al., 2016). However, the fermentation of pre-treated sugarcane bagasse hydrolysate still produced a low final succinate concentration of 18.8 g/L after 24 h with a yield of 0.96 g/g, even after the detoxification (R. Liu et al., 2013). However, it is worth noting that inexpensive carbon sources and media for succinate fermentation are as important as high succinic acid titters with good productivity for a cost-effective bio-succinate industry (S. Chan et al., 2012). Therefore, a biorefinery that co-utilises molasses and lignocelluloses as first- and second-generation (1G-2G) feedstock will be interesting for an improved SA productivities compared to lignocelluloses only.

Alternatively, purer sugarcane-based feedstocks than final molasses and lignocelluloses but cheaper than sucrose such as clarified juice and intermediate (A and B) molasses could be explored for succinate production to avoid the expensive pre-treatment step, especially for lignocelluloses. Brazilian sugarcane biorefineries have demonstrated that the economics of

using these intermediate streams for the co-production of sugar and ethanol (M. O. de S. Dias et al., 2015a; Krajnc & Glavič, 2009) is superior to using the final (C-) molasses (Krajnc & Glavič, 2009). Furthermore, conversion of these intermediate sugar feedstocks in biorefineries will reduce the overall production of crystalline sucrose production, in response to the current oversupply in the international market leading to low market prices (OECD/FAO, 2016). The profitability of succinate production from cheap feedstocks is also a function of the processes employed for the production, recovery and purification of the product.

2.5.1.3. Pre-treatment and enzymatic hydrolysis of lignocellulosic feedstock

Lignocellulose biomass such sugarcane bagasse and residues contain structural carbohydrates, cellulose, embedded in structural biopolymers mainly hemicellulose and lignin, which require pre-treatment to improve its suitability for enzymatic digestion to simple sugars (Menon & Rao, 2012; Nieder-Heitmann et al., 2019). Several pre-treatment methods are presented in the literature with their advantages and disadvantages (Menon & Rao, 2012; O'Hara & Mundree, 2016; Santos et al., 2017). However, a cost-effective pre-treatment method is imperative among other selection criteria for a profitable biorefinery since pre-treatment, in general, is expensive and forms a major cost component of lignocellulose-based biorefineries (M. Mandegari, Farzad, & Görgens, 2018; Nieder-Heitmann et al., 2019). Dilute acid pre-treatment has been reported as one of the most cost-effective methods for pre-treating common lignocellulosic biomass (Humbird et al., 2011; Pachón, Vaskan, Raman, & Gnansounou, 2018).

The dilute acid pre-treated feedstock has two fractions. The liquid hydrolysate contains mainly xylose with some sugar degradation products such as furfural and 5-hydroxymethyl furfural (HMF), which inhibit fermentation. These need to be removed in a cost-effective detoxification step leaving a pure xylose-rich stream for fermentation (Salvachúa et al., 2016). Granulated activated carbon adsorption as a cheap and high efficient method for the detoxification (Nieder-Heitmann et al., 2019; Santos et al., 2017). The solid fraction containing cellulose and lignin is washed thoroughly to remove all inhibitors and xylose (Dong et al., 2017; Pachón et al., 2018) and the wash water is added to the xylose-rich liquid fraction before detoxification. The washed solid is then hydrolysed into glucose and lignin-rich solid using cellulase enzymes (Salvachúa et al., 2016). The conventional hydrolysis procedure used is the separate hydrolysis and fermentation (SHF) method (Bechara et al., 2018). The glucose is separated from the solid residue for the subsequent succinate fermentation.

2.5.1.4. Succinic acid production

Succinic acid can be synthesized industrially from appropriate sugars by microorganisms through anaerobic fermentation (Jiang et al., 2017; Morales et al., 2016). These sugars include glucose, fructose, arabinose, and xylose, which are the most abundant sugars in plant biomass (Vaswani, 2010). The production steps typically include fermentation, recovery, and purification (Vaswani, 2010).

Fermentation

Extensive studies on bio-based succinic acid production have been reported in the literature, using several microorganisms. The strains used currently include rumen bacteria *Actinobacillus succinogenes*, *Mannheimia succiniciproducens* and *Anaerobiospirillum succiniciproducens*, and gastrointestinal bacteria *E. coli* (Ahn, Jang, & Lee, 2017; Choi, Song, Shin, & Lee, 2015; Morales et al., 2016).

Among the rumen bacteria, *A. succinogenes* has been found as the best producer of succinic acid due to its peculiar potential to produce succinic acid from most carbon sources including disaccharides (sucrose, lactose, maltose) pentoses (arabinose, xylose) and hexoses (fructose, galactose, glucose), as well as polyols (sorbitol and glycerol) under anaerobic conditions (Nghiem, Kleff, & Schwegmann, 2017; Vaswani, 2010). Additionally, *A. succinogenes* tolerates high glucose concentrations which enhances fermentation (Vaswani, 2010). However, the growth and metabolism of rumen bacteria such as *A. succinogenes* highly depend on rich and complex nutrients, which increase production cost and hinder its commercial application (Dong et al., 2017; Jiang et al., 2017). Moreover, sufficient genetic tools are lacking to genetically engineer rumen bacteria strains for improved succinate production (Ahn et al., 2017).

Unlike the *A. succinogenes*, *E. coli* is easy to genetically engineer, has a fast growth rate and does not require complex and expensive nutrients, making it a promising strain for cost-effective production of bio-succinic acid (McKinlay et al., 2007; Morales et al., 2016; Vaswani, 2010). These properties reduce the cost and complexity of the *E. coli* fermentation process. Moreover, succinic acid has been produced commercially from commodity sugars using genetically modified *E. coli* (Biddu et al., 2016; Jansen & van Gulik, 2014; Vaswani, 2010). Even though *E. coli* naturally prefers glucose, several strains were engineered to utilise various

carbon sources including sucrose-based substrates for improved succinate production (Jiang et al., 2017; Nghiem et al., 2017).

Economic succinic acid production highly depends on the productivity, titre and yield of the fermentation process (Ahn et al., 2017). Reports showed that succinate production results of *E. coli* fermentation are comparable to the natural succinate producers (Jiang et al., 2017). Chan et al. (2012) reported succinate production from molasses sugars (sucrose, fructose and glucose) with titre, yield and productivity of 55.8 g/L, 0.96 g/g and 0.77 g/L/h, respectively using metabolically engineered *E. coli* strain KJ122-KJSUC-24T. Succinic acid productions of 18.8 g/L and 0.84 g/g were also reported from xylose-rich bagasse hydrolysate (cleaned using activated carbon) using *E. coli* BA204 under anaerobic conditions after 48 h (R. Liu et al., 2013).

Succinic production under anaerobic conditions has the advantage of fixing carbon dioxide as an additional carbon to produce more succinic acid, giving high theoretical yields by (Nghiem et al., 2017). Optimal temperature 37 °C and pH of 7.0 were reported for *E. coli* fermentations of various sugar substrates (S. Chan et al., 2012; Jiang et al., 2017; Thakker, Martínez, San, & Bennett, 2012). The pH is maintained using a base or a carbonate salt. For industrial applications, carbonate salts such as MgCO_3 , NaHCO_3 and Na_2CO_3 may be preferred to simultaneously neutralise the fermentation mixture and serve as the cheap carbon dioxide sources to reduce cost (Jiang et al., 2017). On the other hand, neutralization produces succinate salt requiring an acidification step for the recovery of the free succinic acid, thus increasing the recovery cost (Nghiem et al., 2017). Therefore, research efforts are underway to identify strains that can produce the free succinic acid under slightly acidic conditions to minimize the succinate salt production and consequently, the downstream processing cost (Jansen & van Gulik, 2014).

Recovery and purification

The downstream processing accounts for more than half of the total cost of microbial succinic acid production (Ahn et al., 2017; Cheng et al., 2012). The first step involves the removal of biomass cells from the fermentation broth by filtration or centrifugation (Nghiem et al., 2017). Cell recycle reactors have been proposed for continuous succinate production and subsequently more efficient separation and purification but require further studies for large scale application (Cao et al., 2013; Vaswani, 2010). The major challenge for commercial SA facilities after this step is how to efficiently recover SA from the fermentation broth in a cost-effectiveness way.

Various downstream processes have been investigated for the recovery and purification of succinic acid including precipitation, direct crystallization, extraction, electrodialysis and ion-exchange membrane processes (Alexandri et al., 2019; Morales et al., 2016). Among the processes studied in the literature, reactive extraction was widely reported as a simple and less energy-intensive method for the recovery of carboxylic acids such as succinic acid, and economical upon recycling of the reagents used (Ahn et al., 2017; Cheng et al., 2012; Morales et al., 2016). The recovery and purification of succinic acid using reactive extraction typically results in succinic acid purity of more than 99.5 % (Ahn et al., 2017).

The reactive extraction method uses tertiary amines to extract succinic acid from the fermentation broth into the organic phase (Alexandri et al., 2019). A mixture of tri-*n*-octylamine (TOA) and 1-octanol was reported as an effective reactive extraction system with high selectivity for succinic acid (Ahn et al., 2017; Huh et al., 2006). Then, the succinic acid is recovered from the organic phase into an aqueous phase composed of trimethylamine in water. The back extraction is necessary for the reuse of the reactive extraction system (Kurzrock, Schallinger, & Weuster-Botz, 2011). It is worth noting that recycling of the extractive agent as well as reduction of by-product impurities through strategic fermentation approach are necessary to minimize the cost of the downstream processing (Ahn et al., 2017; Alexandri et al., 2019; Cheng et al., 2012; Kurzrock et al., 2011). In addition to the reactive extraction, Huh et al. (2006) proposed subsequent distillation and crystallization for the effective purification of the succinic acid. The distillation process separates the trimethyl-amine for recycling to the back extraction column to reduce cost while eliminating volatile by-product impurities and concentrating the product, followed by crystallization (Ahn et al., 2017; Huh et al., 2006).

2.5.1.5. Techno-economic and exergy studies

Techno-economic assessment of biorefineries is important for the implementation of biorefineries of interest (M. A. Mandegari, Farzad, & Görgens, 2017). Majority of techno-economic studies on sugarcane biorefineries are reported for biofuels especially, ethanol (M. O. S. Dias et al., 2013; M. A. Mandegari et al., 2017; Petersen, Van der Westhuizen, Mandegari, & Görgens, 2018). It has been shown that integrating biorefineries into an existing sugar mill improves the economics (Ali Mandegari et al., 2017). However, only one study was found in the literature on SA production from pentoses integrated into sugarcane bioethanol plant (Klein et al., 2017).

Other techno-economic studies have been reported for stand-alone biorefineries producing SA from crystalline sucrose (Efe et al., 2013) and lignocellulose (Nieder-Heitmann et al., 2019) feedstocks. While the sucrose raw material was found to contribute 29 % to the yearly operational cost (Efe et al., 2013), cheaper lignocellulose feedstock required expensive pre-treatment and hydrolysis, which affected the economic performance of the biorefinery (Nieder-Heitmann et al., 2019). Molasses is a cheap sucrose source and is worth investigating for SA production in an integrated biorefinery aimed at improved economic performance.

Moreover, biorefineries integrating the use of first- and second-generation (1G-2G) have been reported to achieve notable performance compared to other configurations (Moncada et al., 2013). Therefore, the 1G-2G configuration can also be explored for SA production.

It is important to note that techno-economic evaluation alone is not enough to assess the sustainability and competitiveness of the future SA biorefinery. Other aspects including energy and environmental issues are equally important (Pinazo, Domine, Parvulescu, & Petru, 2015). Exergy analysis and its extensions are powerful tools proven to provide insight into the design of sustainable processes (Aghbashlo, Mandegari, et al., 2018a; Aghbashlo, Tabatabaei, & Karimi, 2016). Even though some exergy-based studies have been reported for sugarcane biorefineries (Aghbashlo, Mandegari, et al., 2018a; K Ojeda, Ávila, Suárez, & Kafarov, 2010; Karina Ojeda, Sánchez, & Kafarov, 2011; Soltanian et al., 2019), no such study was found for SA production.

2.5.2. Short-chain fructooligosaccharides (scFOS)

Increasing awareness to maintain physical health in response to the prevalence of diseases and health risks nowadays has spurred research interest in functional foods such as the short-chain fructooligosaccharides (scFOS). This is partly responsible for decreases in the global demand and prices for sugar as scFOS is a lower-calorie sweetener for replacement of sucrose, which is useful for individuals with diabetes.

2.5.2.1. Product description and market

Short-chain Fructooligosaccharides (scFOS) are fructose oligomers consisting of a terminal glucose group linked to 2 – 4 fructose units by β (1 \rightarrow 2) – glycosidic bonds, called 1-kestose (GF2), nystose (GF3) and fructofuranosyl nystose (GF4), respectively (Ur Rehman et al., 2016). ScFOS have lower calorific value (50-75 %) and are 30-60 % as sweeter as sucrose, making them suitable as low-calorie sweeteners in the food and pharmaceutical industries (Bali

et al., 2015; Clarisse Nobre, Teixeira, & Rodrigues, 2015). Furthermore, scFOS have functional properties as prebiotics with several health benefits, including prevention of colon cancer, reduction of cholesterol levels in the body and improved mineral absorption in the gut (Dominguez, Rodrigues, Lima, & Teixeira, 2014a). These functionalities and health benefits have led to increased consumer demand for scFOS. The global scFOS market was reported as 134 000 tons in 2015 and projected to almost double by 2024 with major applications in food and beverages (47 %), dairy products (21 %) and health products (17 %) (Grand View Research Inc., 2016). Over 50 % of the 2015 global scFOS market was supplied by synthesis from sucrose (Grand View Research Inc., 2016) offering great economic potential in the sugar industry as a biorefinery product (Bali et al., 2015; Ur Rehman et al., 2016).

2.5.2.2. ScFOS production

Short-chain fructooligosaccharides (scFOS) are naturally found in many food sources including onion, garlic, tomato, banana and Jerusalem artichoke (Dominguez et al., 2014a), but in small concentrations (Clarisse Nobre et al., 2015). Industrially, they are produced on commercial levels by the degradation of inulin or transformation of sucrose by enzymes isolated from bacteria, fungi or yeast sources (Bedzo, Mandegari, et al., 2019). The transformation of sucrose to produce scFOS is mostly done using fungal enzymes classified as β -D-fructofuranosidase (FFase, EC 3.2.1.26) or fructosyltransferases (FTase, EC 2.4.1.9) (Flores-Maltos et al., 2016).

FFases have high transfructosylase activity for scFOS production (Flores-Maltos et al., 2016) and can be produced by the yeast *Pichia pastoris* extracellularly into the supernatant (Bedzo, Trollope, Gottumukkala, Coetzee, & Görgens, 2019). The presence of the FFase in the supernatant allows for its direct use without having to break the yeast cells. Besides, the use of the free enzyme without immobilizing it was found to be more economical in scFOS production (Bedzo, Mandegari, et al., 2019), while producing the enzyme in the same facility with scFOS was noted to enhance the economic performance of the plant (Bedzo, Mandegari, et al., 2019; Katarína Vaňková, Onderková, Antošová, & Polakovič, 2008). Bedzo, Trollope, et al. (2019) reported FFase production using glycerol as a substrate with basal salt medium supplemented with *Pichia* trace elements and oxygen aeration. The yeast was grown for 72 h under pH condition of 5.0, maintained using ammonium hydroxide (Bedzo, Trollope, et al., 2019) and at the optimal growth and protein expression temperature of 30 °C (Bedzo, Mandegari, et al.,

2019). Xie et al. (2017) showed that FFase can be produced from molasses as a cheaper carbon source than glycerol to reduce the production cost.

The maximum yield of scFOS from sucrose using FFase (free enzyme) in a batch reactor was reported as 61 % (Bedzo, Trollope, et al., 2019). Similarly, Flores-Maltos et al. (2016) reported a range of scFOS yields (averagely 48-69 %) obtained from sucrose using FTase from various microbial sources in different processes (Dominguez et al., 2014a). The process type is one of the factors that affect the scFOS production yield. The batch process commonly used results in the accumulation of glucose from the sucrose transformation reaction, which inhibits the enzyme action and hence, further scFOS formation. Research efforts have been made to increase the scFOS yield for more profitable production.

Sheu et al. (2001) reported 90 % yield when the scFOS production enzyme, β -fructofuranosidase was mixed with glucose oxidase to convert the glucose to gluconic acid and later precipitated to calcium gluconate in solution. This process could also serve as a purification method (Clarisse Nobre et al., 2015). Moreover, Ur Rehman et al. (2016) studied a continuous type enzymatic membrane reactor (EMR) for scFOS production from molasses, which can remove the glucose as it is produced and also allow the recycling of the enzyme and reported 63 % scFOS yield compared to 58 % in a batch reactor. However, the economics of these processes have not been established.

On the other hand, a cost-effective scFOS production can be achieved with a cheaper sucrose source such as molasses (Shin, 2004). Shin (2004) investigated the most promising strain of *Aureobasidium pullulans* for scFOS production from molasses and reported 46 % conversion of molasses sucrose to scFOS after 24 h incubation at 55 °C and pH of 5.5, with strain KCCM 12017. In another study, 58 % scFOS from molasses in a batch reactor was reported (Ur Rehman et al., 2016). Slightly lower yields of scFOS were reported for molasses than for pure sucrose because of impurities in the molasses and higher glucose inhibition (Ur Rehman et al., 2016). In addition to the scFOS product, the reaction mixture contains sugar impurities consisting of largely glucose, fructose and unconverted sucrose that need to be removed (Clarisse Nobre et al., 2015).

2.5.2.3. ScFOS purification

To separate and purify scFOS from the reaction mixture, several technologies have been used including continuous adsorption of scFOS using activated charcoal fixed-bed (Bali et al., 2015;

Clarisse Nobre et al., 2015) or zeolite fixed-bed (Kuhn, Mazutti, & Filho, 2014) columns, membrane filtration, ion-exchange chromatography (Bali et al., 2015; Clarisse Nobre et al., 2015) and simulated moving bed (SMB) chromatography (Clarisse Nobre et al., 2015). The SMB chromatography uses a series of chromatographic columns in a continuous system to give efficient separation and high productivities with low solvent consumption for improved exergy efficiency. This makes it cost-effective for commercial scFOS purification (C. Nobre et al., 2016) as it is already being used in the sugar industry for the separation of fructose from corn syrup (Sá Gomes & Rodrigues, 2012).

Vaňková & Polakovič (2012) designed SMB chromatography for a large scale purification of scFOS from a reaction mixture containing 61.7 % total scFOS, 2.1 % fructose, 21.4 % glucose and 14.8 % unreacted sucrose. ScFOS yield and purity of 95 % and 90 % were recorded, respectively. The purified scFOS is either spray-dried into powder or concentrated a syrup as the final product (Soni & Tsai, 2016; Katarína Vaňková et al., 2008).

2.5.2.4. Techno-economic and exergy studies

The technical aspects of scFOS production presented earlier in this section are mainly driven by the need for an energy/exergy efficient and cost-effective process (Flores-Maltos et al., 2016). Therefore, it is important to assess the sustainability of the scFOS production process based on these criteria. However, such studies are not common for scFOS. Vaňková et al. (2008) analysed the economic feasibility of industrial production scFOS powder and syrup based on the conversion of sucrose by immobilised FTase. Other techno-economic studies were reported for scFOS production from pure crystalline sucrose as stand-alone plants focusing on the types of enzyme systems and fermentations used rather than the raw material (Bedzo, Mandegari, et al., 2019; Mussatto, Aguiar, Marinha, Jorge, & Ferreira, 2015). No techno-economic study was reported for scFOS production from cheaper sucrose sources such as molasses. Moreover, no exergy studies were found for scFOS production.

2.6. Summary of key shortcomings in the literature

In summary, a typical sugar mill involves energy-intensive processes designed for maximum crystalline sucrose recovery and are energy self-sustained by burning bagasse in cogeneration systems. The energy systems of traditional sugar mills were designed inefficiently to avoid the initial challenges with bagasse disposal. However, due to inefficiencies in process transients and the increasing importance of bagasse leading to its export for other uses, a typical South

African sugar mill now burns coal to supplement bagasse as fuel, which is not environmentally sustainable. To avoid the use of coal and to save bagasse for other important uses, it is imperative to improve the energy efficiency of the traditional sugar mill. Exergy analysis has been widely reported to reveal the true thermodynamic inefficiency of energy systems to improve them. However, no comprehensive exergy analysis has been reported for raw cane sugar production plant similar to a typical South African sugar mill. Following the gaps identified in the literature, objective 1 was set out **to evaluate the inefficiency of a typical South African sugar mill through exergy analysis** towards improving its energy efficiency and consequently, its sustainability and economic competitiveness. The contributions from this objective are reported in Chapter 3 for the raw sugar production processes and in Chapter 4 for the cogeneration system of the sugar mill.

Among the technologies reported in the literature to improve energy efficiency, absorption heat pump and organic Rankine cycle for waste heat recovery show to be sustainable technologies for this purpose. However, their sustainability will be better proven with exergy considerations in addition to their economic viability. Moreover, no report has been found in the literature to show the extent of efficiency improvement attained by these technologies in a sugar industry in terms of exergy. Objective 2 (**to select suitable energy-efficient technologies for integration into the base sugar mill towards energy/exergy efficiency improvement**) and objective 3 (**to assess the economic feasibility of integrating the selected technologies into the sugar mill**), accomplished in Chapter 4 of the thesis were set to fill these gaps in the literature, based on the exergy results from Chapter 3.

Furthermore, the ultimate goal of improving the energy efficiency of the sugar mill is to ensure cost-effective production and economic and environmental sustainability of the industry. The sugarcane biorefinery concept also seeks to attain the same goal. Succinic acid (SA) and short-chain fructooligosaccharides (scFOS) were reported as promising products for the sugarcane biorefinery. However, reports from the literature show that their sustainability has only been assessed based on simple economic principles and not exergy. Based on the knowledge of the inefficiencies identified by the exergy analysis of the sugar mill in Chapter 3 and considering the efficiency improvement opportunities presented in Chapter 4, it was also the target of this study **to design economically feasible biorefineries integrated into the sugar mill considering exergy principles** (objective 4). This objective was addressed in Chapter 5 of the thesis.

Finally, objective 5 was **to apply exergoeconomic analysis for the selection of the preferred biorefinery for a cost-effective integration into the sugar mill**. The literature shows that exergoeconomic analysis does not only reveal process inefficiencies but also evaluates the cost associated with them, providing insightful information for a cost-efficient design. Therefore, it is explored in Chapter 6 as a novel selection tool for the biorefineries developed in Chapter 5, based on the exergy outcomes of Chapter 3 and Chapter 4.

References

- Aghbashlo, M., Mandegari, M., Tabatabaei, M., Farzad, S., Mojarab Soufiyan, M., & Görgens, J. F. (2018). Exergy analysis of a lignocellulosic-based biorefinery annexed to a sugarcane mill for simultaneous lactic acid and electricity production. *Energy*, *149*, 623–638. <http://doi.org/10.1016/j.energy.2018.02.063>
- Aghbashlo, M., Tabatabaei, M., Jazini, H., & Ghaziaskar, H. S. (2018). Exergoeconomic and exergoenvironmental co-optimization of continuous fuel additives (acetins) synthesis from glycerol esterification with acetic acid using Amberlyst 36 catalyst. *Energy Conversion and Management*, *165*, 183–194. <http://doi.org/10.1016/j.enconman.2018.03.054>
- Aghbashlo, M., Tabatabaei, M., & Karimi, K. (2016). Exergy-based sustainability assessment of ethanol production via *Mucor indicus* from fructose, glucose, sucrose, and molasses. *Energy*, *98*, 240–252. <http://doi.org/10.1016/j.energy.2016.01.029>
- Aghbashlo, M., Tabatabaei, M., Soltanian, S., Ghanavati, H., & Dadak, A. (2019). Comprehensive exergoeconomic analysis of a municipal solid waste digestion plant equipped with a biogas genset. *Waste Management*, *87*, 485–498. <http://doi.org/10.1016/j.wasman.2019.02.029>
- Ahn, J. H., Jang, Y. S., & Lee, S. Y. (2017). Succinic Acid. In C. Wittmann & J. C. Liao (Eds.), *Industrial Biotechnology: Products and Processes* (pp. 505–544). Weinheim, Germany: Wiley ‐VCH Verlag GmbH & Co. KGaA. <http://doi.org/10.1002/9783527807833>
- Alexandri, M., Vlysidis, A., Papapostolou, H., Tverezovskaya, O., Tverezovskiy, V., Kookos, I. K., & Koutinas, A. (2019). Downstream separation and purification of succinic acid from fermentation broths using spent sulphite liquor as feedstock. *Separation and*

- Purification Technology*, 209(January 2018), 666–675.
<http://doi.org/10.1016/j.seppur.2018.08.061>
- Ali Mandegari, M., Farzad, S., & Görgens, J. F. (2017). Economic and environmental assessment of cellulosic ethanol production scenarios annexed to a typical sugar mill. *Bioresource Technology*, 224, 314–326. <http://doi.org/10.1016/j.biortech.2016.10.074>
- Alonso-Pippo, W., Luengo, C. a, Felfli, F. F., Garzone, P., & Cornacchia, G. (2009). Energy recovery from sugarcane biomass residues: Challenges and opportunities of bio-oil production in the light of second generation biofuels. *Journal of Renewable and Sustainable Energy*, 1(063102), 1–15. <http://doi.org/10.1063/1.3259170>
- Atmaca, A., & Yumrutaş, R. (2014). Thermodynamic and exergoeconomic analysis of a cement plant: Part I – Methodology. *Energy Conversion and Management*, 79, 790–798. <http://doi.org/10.1016/j.enconman.2013.11.053>
- Bakhtiari, B., Fradette, L., Legros, R., & Paris, J. (2010). Opportunities for the integration of absorption heat pumps in the pulp and paper process. *Energy*, 35(12), 4600–4606. <http://doi.org/10.1016/j.energy.2010.03.047>
- Bali, V., Panesar, P. S., Bera, M. B., & Panesar, R. (2015). Fructo-oligosaccharides: Production, Purification and Potential Applications. *Critical Reviews in Food Science and Nutrition*, 55(11), 1475–1490. <http://doi.org/10.1080/10408398.2012.694084>
- Barros, M. De, Freitas, S., Padilha, G. S., & Alegre, R. M. (2013). Biotechnological Production of Succinic Acid by *Actinobacillus Succinogenes* Using Different Substrate . *Chemical Engineering Transactions*, 32, 985–990.
- Bayrak, M., Midilli, A., & Nurveren, K. (2003). Energy and exergy analyses of sugar production stages. *International Journal of Energy Research*, 27(11), 989–1001. <http://doi.org/10.1002/er.916>
- Bechara, R., Gomez, A., Saint-Antonin, V., Schweitzer, J., Maréchal, F., & Ensinas, A. (2018). Review of design works for the conversion of sugarcane to first and second-generation ethanol and electricity. *Renewable and Sustainable Energy Reviews*, 91(January 2017), 152–164. <http://doi.org/10.1016/j.rser.2018.02.020>
- Bedzo, O. K. K., Mandegari, M., & Görgens, J. F. (2019). Comparison of immobilized and free enzyme systems in industrial production of short-chain fructooligosaccharides from

- sucrose using a techno-economic approach. *Biofuels, Bioproducts and Biorefining*, bbb.2025. <http://doi.org/10.1002/bbb.2025>
- Bedzo, O. K. K., Trollope, K., Gottumukkala, L. D., Coetzee, G., & Görgens, J. F. (2019). Amberlite IRA 900 versus calcium alginate in immobilization of a novel, engineered β -fructofuranosidase for short-chain fructooligosaccharide synthesis from sucrose. *Biotechnology Progress*, 35(3), e2797. <http://doi.org/10.1002/btpr.2797>
- Biddy, M. J., Scarlata, C., Biddy, M. J., & Scarlata, C. (2016). *Chemicals from Biomass : A Market Assessment of Bioproducts with Near-Term Potential*. National Renewable Energy Laboratory (NREL), Denver.
- Birru, E., Martin, A., & Erlich, C. (2016). *Sugar cane industry overview and energy efficiency considerations*. Stockholm, Sweden.
- Bocci, E., Di Carlo, a., & Marcelo, D. (2009). Power plant perspectives for sugarcane mills. *Energy*, 34(5), 689–698. <http://doi.org/10.1016/j.energy.2009.02.004>
- Boch Andersen, T., & Thomsen, K. (2009). Separation of water through gas hydrate formation. *International Sugar Journal*, 111(1330), 632–636.
- Borges, E. R., & Pereira, N. (2011). Succinic acid production from sugarcane bagasse hemicellulose hydrolysate by *Actinobacillus succinogenes*. *Journal of Industrial Microbiology & Biotechnology*, 38(8), 1001–1011. <http://doi.org/10.1007/s10295-010-0874-7>
- Cao, Y., Zhang, R., Sun, C., Cheng, T., Liu, Y., & Xian, M. (2013). Fermentative Succinate Production: An Emerging Technology to Replace the Traditional Petrochemical Processes. *BioMed Research International*, 2013, 723412. <http://doi.org/10.1155/2013/723412>
- Carrier. (n.d.). Lithium Bromide Absorption Chiller. Retrieved from [http://www.carrier.com.hk/comm/comm_new2010/2013_Cat/Absorption_Chiller/16JLR_CN\(C\).pdf](http://www.carrier.com.hk/comm/comm_new2010/2013_Cat/Absorption_Chiller/16JLR_CN(C).pdf)
- Chan, C. W., Ling-Chin, J., & Roskilly, A. P. (2013). A review of chemical heat pumps, thermodynamic cycles and thermal energy storage technologies for low grade heat utilisation. *Applied Thermal Engineering*, 50(1), 1257–1273. <http://doi.org/10.1016/j.applthermaleng.2012.06.041>

- Chan, S., Kanchanatawee, S., & Jantama, K. (2012). Production of succinic acid from sucrose and sugarcane molasses by metabolically engineered *Escherichia coli*. *Bioresource Technology*, *103*(1), 329–336. <http://doi.org/10.1016/j.biortech.2011.09.096>
- Chen, H., Goswami, D. Y., & Stefanakos, E. K. (2010). A review of thermodynamic cycles and working fluids for the conversion of low-grade heat. *Renewable and Sustainable Energy Reviews*, *14*(9), 3059–3067. <http://doi.org/10.1016/j.rser.2010.07.006>
- Cheng, K.-K., Zhao, X.-B., Zeng, J., Wu, R.-C., Xu, Y.-Z., Liu, D.-H., & Zhang, J.-A. (2012). Downstream processing of biotechnological produced succinic acid. *Applied Microbiology and Biotechnology*, *95*(4), 841–850. <http://doi.org/10.1007/s00253-012-4214-x>
- Cherubini, F. (2010). The biorefinery concept: Using biomass instead of oil for producing energy and chemicals. *Energy Conversion and Management*, *51*(7), 1412–1421. <http://doi.org/10.1016/j.enconman.2010.01.015>
- Choi, S., Song, C. W., Shin, J. H., & Lee, S. Y. (2015). Biorefineries for the production of top building block chemicals and their derivatives. *Metabolic Engineering*, *28*, 223–239. <http://doi.org/10.1016/j.ymben.2014.12.007>
- CIAB. (2010). *Power generation from coal: Measuring and reporting efficiency performance and CO2 emissions*. Paris, France.
- Clarke, S. J. (1999). Outlook for Emerging Technologies in Sugar Processing. Florida, USA.
- Cleveland, C. J., & Ayres, R. U. (2004). Encyclopedia of energy. In *Energy*.
- Conningarth Economists. (2013). *Investigation and evaluation of alternative uses and products from sugar cane: A Cost benefit and macroeconomic impact analysis*. (Vol. 27). Pretoria, South Africa. Retrieved from [http://www.namc.co.za/upload/Document 5 Alternative Uses.pdf](http://www.namc.co.za/upload/Document%205%20Alternative%20Uses.pdf)
- Coustel, J., & Journet, G. (2009). Falling-film evaporator performance results from two Reunion Island factories. *Sugar Industry*, *134*(4), 225–229.
- de Jong, E., Higson, A., Walsh, P., & Wellisch, M. (2012). *Bio-based chemicals: Value added products from biorefineries*. Retrieved from www.ieabioenergy.com
- Deshmukh, R., Jacobson, A., Chamberlin, C., & Kammen, D. (2013). Thermal gasification or

- direct combustion? Comparison of advanced cogeneration systems in the sugarcane industry. *Biomass and Bioenergy*, 55, 163–174. <http://doi.org/10.1016/j.biombioe.2013.01.033>
- Dias, M. O. de S., Maciel Filho, R., Mantelatto, P. E., Cavalett, O., Rossell, C. E. V., Bonomi, A., & Leal, M. R. L. V. (2015). Sugarcane processing for ethanol and sugar in Brazil. *Environmental Development*, 15, 35–51. <http://doi.org/10.1016/j.envdev.2015.03.004>
- Dias, M. O. S., Junqueira, T. L., Cavalett, O., Pavanello, L. G., Cunha, M. P., Jesus, C. D. F., Bonomi, A. (2013). Biorefineries for the production of first and second generation ethanol and electricity from sugarcane. *Applied Energy*, 109, 72–78. <http://doi.org/10.1016/j.apenergy.2013.03.081>
- Dogbe, E. S., Mandegari, M. A., & Görgens, J. F. (2018). Exergetic diagnosis and performance analysis of a typical sugar mill based on Aspen Plus® simulation of the process. *Energy*, 145, 614–625. <http://doi.org/10.1016/j.energy.2017.12.134>
- Dominguez, A. L., Rodrigues, L. R., Lima, N. M., & Teixeira, J. A. (2014). An Overview of the Recent Developments on Fructooligosaccharide Production and Applications. *Food and Bioprocess Technology*, 7(2), 324–337. <http://doi.org/10.1007/s11947-013-1221-6>
- Dong, W., Xue, M., Zhang, Y., Xin, F., Wei, C., Zhang, W., Jiang, M. (2017). Characterization of a β -glucosidase from *Paenibacillus* species and its application for succinic acid production from sugarcane bagasse hydrolysate. *Bioresource Technology*, 241, 309–316. <http://doi.org/10.1016/j.biortech.2017.05.141>
- Efe, Ç., van der Wielen, L. A. M., & Straathof, A. J. J. (2013). Techno-economic analysis of succinic acid production using adsorption from fermentation medium. *Biomass and Bioenergy*, 56, 479–492. <http://doi.org/10.1016/j.biombioe.2013.06.002>
- Eggleston, G., & Lima, I. (2015). Sustainability Issues and Opportunities in the Sugar and Sugar-Bioprocess Industries. *Sustainability*, 7(9), 12209–12235. <http://doi.org/10.3390/su70912209>
- Ensinas, A. V., Modesto, M., Nebra, S. A., & Serra, L. (2009). Reduction of irreversibility generation in sugar and ethanol production from sugarcane. *Energy*, 34(5), 680–688. <http://doi.org/10.1016/j.energy.2008.06.001>
- Ensinas, A. V., Nebra, S. A., Lozano, M. A., & Serra, L. M. (2007). Analysis of process steam

- demand reduction and electricity generation in sugar and ethanol production from sugarcane. *Energy Conversion and Management*, 48(11), 2978–2987. <http://doi.org/10.1016/j.enconman.2007.06.038>
- Ensinas, A. V, Nebra, S. a, Lozano, M. a, & Serra, L. (2006). Analysis of cogeneration systems in sugar cane factories - Alternatives of steam and combined cycle power plants. In *Proceedings of ECOS* (pp. 1177–1184). Aghia Pelagia, Crete, Greece.
- Eslamimanesh, A., Mohammadi, A. H., Richon, D., Naidoo, P., & Ramjugernath, D. (2012). Application of gas hydrate formation in separation processes: A review of experimental studies. *The Journal of Chemical Thermodynamics*, 46, 62–71. <http://doi.org/10.1016/j.jct.2011.10.006>
- Eyidogan, M., Canka, F., Kaya, D., & Coban, V. (2016). Investigation of organic Rankine cycle (ORC) technologies in Turkey from the technical and economic point of view. *Renewable and Sustainable Energy Reviews*, 58, 885–895. <http://doi.org/10.1016/j.rser.2015.12.158>
- Farzad, S., Mandegari, M. A., & Gorgens, J. F. (2015). Biorefinery annexed to South African sugar mill, Part II: Energy sufficiency analysis. In *International Congress and Expo on Biofuels & Bioenergy*. Valencia, Spain. Retrieved from <http://dx.doi.org/10.4172/2090-4541.S1.002>
- Ferone, M., Raganati, F., Olivieri, G., Salatino, P., & Marzocchella, A. (2017). Biosuccinic Acid from Lignocellulosic-Based Hexoses and Pentoses by *Actinobacillus succinogenes*: Characterization of the Conversion Process. *Applied Biochemistry and Biotechnology*, 1–13. <http://doi.org/10.1007/s12010-017-2514-4>
- Finlayson, B. A. (2015). AspenTech. AspenTech.
- Flores-Maltos, D. A., Mussatto, S. I., Contreras-Esquivel, J. C., Rodríguez-Herrera, R., Teixeira, J. A., & Aguilar, C. N. (2016). Biotechnological production and application of fructooligosaccharides. *Critical Reviews in Biotechnology*, 36(2), 259–267. <http://doi.org/10.3109/07388551.2014.953443>
- Garg, A., Sharma, M. P., & Sharma, V. (2016). Exergy and energy analyses of a sugarcane juice production and clarification unit. *International Journal of Exergy*, 19(1), 78. <http://doi.org/10.1504/IJEX.2016.074268>

- Giuliano, A., Cerulli, R., Poletto, M., Raiconi, G., & Barletta, D. (2016). Process Pathways Optimization for a Lignocellulosic Biorefinery Producing Levulinic Acid, Succinic Acid, and Ethanol. *Industrial & Engineering Chemistry Research*, 55(40), 10699–10717. <http://doi.org/10.1021/acs.iecr.6b01454>
- Gourmelon, S., They-Hetreux, R., Floquet, P., Baudouin, O., Baudet, P., & Campagnolo, L. (2015). Exergy analysis in ProSimPlus ® simulation software: A focus on exergy efficiency evaluation. *Computers & Chemical Engineering*, 79, 91–112. <http://doi.org/10.1016/j.compchemeng.2015.02.014>
- Grand View Research Inc. (2016). *Fructooligosaccharides (FOS) Market Analysis By Source (Inulin, Sucrose) By Application (Food & Beverages, Infant Formula, Dietary Supplements, Animal Feed, Pharmaceuticals) And Segment Forecasts To 2024*. US. Retrieved from <https://www.grandviewresearch.com/industry-analysis/fructooligosaccharides-market>
- Guest, K., Stark, A., & Starzak, M. (2019). Development and verification of an ASPEN Plus ® model of a sugarcane biorefinery. *Proc S Afr Sug Technol Ass*, 92(August), 254–273.
- Gul, S., & Harasek, M. (2012). Energy saving in sugar manufacturing through the integration of environmental friendly new membrane processes for thin juice pre-concentration. *Applied Thermal Engineering*, 43, 128–133. <http://doi.org/10.1016/j.applthermaleng.2011.12.024>
- Gurumurthy, B. S. (2011). Advancement in sugar processing technologies in India emerging innovations. *Sugar Tech*, 13(4), 378–393. <http://doi.org/10.1007/s12355-011-0106-0>
- Héctor, V. I., Carolina Mesa, M., & Sergio, G. A. A. (2013). Energy and exergy analysis of the combined production process of sugar and ethanol from sugarcane (a Colombia case study). In *Proceedings of ECOS* (pp. 1–14).
- Heluane, H., Colombo, M., Hernández, M. R., Graells, M., & Puigjaner, L. (2007). Enhancing sugar cane process performance through optimal production scheduling. *Chemical Engineering and Processing: Process Intensification*, 46(3), 198–209. <http://doi.org/10.1016/j.cep.2006.05.015>
- Hess, S., Beukes, H., Smith, G., & Dinter, F. (2016). Initial study on solar process heat for South African sugar mills. In *Proc S Afr Sug Technol Ass* (pp. 324–349). Durban, South

Africa.

- Higa, M., Freitas, a. J., Bannwart, a. C., & Zemp, R. J. (2009). Thermal integration of multiple effect evaporator in sugar plant. *Applied Thermal Engineering*, 29(2–3), 515–522. <http://doi.org/10.1016/j.applthermaleng.2008.03.009>
- Hinderink, A. P., Kerkhof, F. P. J. M., Lie, A. B. K., & Van Der Kooi, H. J. (1996). Exergy analysis with a flowsheeting simulator - I . Theory ; Calculating exergies of material streams. *Chemical Engineering Science*, 51(20), 4693–4700. [http://doi.org/10.1016/0009-2509\(96\)00220-5](http://doi.org/10.1016/0009-2509(96)00220-5)
- Hofsetz, K., & Silva, M. A. (2012). Brazilian sugarcane bagasse: Energy and non-energy consumption. *Biomass and Bioenergy*, 46, 564–573. <http://doi.org/10.1016/j.biombioe.2012.06.038>
- Hugot, E. (1972). *Handbook of Cane Sugar Engineering* (2nd ed.). Elsevier Science Ltd.
- Huh, Y. S., Jun, Y.-S., Hong, Y. K., Song, H., Lee, S. Y., & Hong, W. H. (2006). Effective purification of succinic acid from fermentation broth produced by *Mannheimia succiniciproducens*. *Process Biochemistry*, 41(6), 1461–1465. <http://doi.org/10.1016/j.procbio.2006.01.020>
- Humbird, D., Davis, R., Tao, L., Kinchin, C., Hsu, D., & Aden, A. (2011). *Process Design and Economics for Biochemical Conversion of Lignocellulosic Biomass to Ethanol. Technical Report NREL/TP-5100-47764*. Golden, Colorado.
- Ibarra-Bahena, J., & Romero, R. (2014). Performance of different experimental absorber designs in absorption heat pump cycle technologies: A review. *Energies*, 7(2), 751–766. <http://doi.org/10.3390/en7020751>
- International Sugar Organization. (2016). About sugar. Retrieved August 28, 2017, from <http://www.isosugar.org>
- Jacobs Consultancy. (2015). What is ExerCom? Retrieved October 20, 2015, from http://www.cocos.nl/en/548/ccs/Wat_is_ExerCom.html
- Jansen, M. La, & van Gulik, W. M. (2014). Towards large scale fermentative production of succinic acid. *Current Opinion in Biotechnology*, 30, 190–197. <http://doi.org/10.1016/j.copbio.2014.07.003>

- Jiang, M., Ma, J., Wu, M., Liu, R., Liang, L., Xin, F., Dong, W. (2017). Progress of succinic acid production from renewable resources: Metabolic and fermentative strategies. *Bioresource Technology*, 245, 1710–1717. <http://doi.org/10.1016/j.biortech.2017.05.209>
- Jiang, M., Xu, R., Xi, Y.-L., Zhang, J.-H., Dai, W.-Y., Wan, Y.-J., Wei, P. (2013). Succinic acid production from cellobiose by *Actinobacillus succinogenes*. *Bioresource Technology*, 135, 469–474. <http://doi.org/10.1016/j.biortech.2012.10.019>
- Kamate, S. C., & Gangavati, P. B. (2009a). Cogeneration in Sugar Industries: Technology Options and Performance Parameters—A Review. *Cogeneration & Distributed Generation Journal*, 24(4), 6–33. <http://doi.org/10.1080/15453660909595148>
- Kamate, S. C., & Gangavati, P. B. (2009b). Exergy analysis of cogeneration power plants in sugar industries. *Applied Thermal Engineering*, 29(5–6), 1187–1194. <http://doi.org/10.1016/j.applthermaleng.2008.06.016>
- Keil, C., Plura, S., Radspieler, M., & Schweigler, C. (2008). Application of customized absorption heat pumps for utilization of low-grade heat sources. *Applied Thermal Engineering*, 28(16), 2070–2076. <http://doi.org/10.1016/j.applthermaleng.2008.04.012>
- Khawaji, A. D., Kutubkhanah, I. K., & Wie, J.-M. (2008). Advances in seawater desalination technologies. *Desalination*, 221(1–3), 47–69. <http://doi.org/10.1016/j.desal.2007.01.067>
- Khoodaruth, A. (2015). Use of falling thin film evaporator for increasing cogenerated electricity in cane flexi-factory in Mauritius. *Energy Procedia*, 75, 1021–1026. <http://doi.org/10.1016/j.egypro.2015.07.349>
- Klein, B. C., Silva, J. F. L., Junqueira, T. L., Rabelo, S. C., Arruda, P. V., Ienczak, J. L., Bonomi, A. (2017). Process development and techno-economic analysis of bio-based succinic acid derived from pentoses integrated to a sugarcane biorefinery. *Biofuels, Bioproducts and Biorefining*, 11(6), 1051–1064. <http://doi.org/10.1002/bbb.1813>
- Kotas, T. J. (1995). *The Exergy Method of Thermal Plant Analysis* (Reprint Ed). Florida, USA: Krieger Publishing Company, Malabar. <http://doi.org/10.1016/B978-0-408-01350-5.50003-9>
- Krajnc, D., & Glavič, P. (2009). Assessment of different strategies for the co-production of bioethanol and beet sugar. *Chemical Engineering Research and Design*, 87(9), 1217–1231. <http://doi.org/10.1016/j.cherd.2009.06.014>

- Kuhn, R. C., Mazutti, M. A., & Filho, F. M. (2014). Separation and purification of fructooligosaccharides on a zeolite fixed-bed column. *Journal of Separation Science*. <http://doi.org/10.1002/jssc.201300979>
- Kumar, U., & Karimi, M. N. (2014). Low grade waste heat recovery for optimized energy efficiencies and enhanced sustainability in process industries : A comprehensive review. *International Journal of Multidisciplinary Sciences and Engineering*, 5(4), 15–26.
- Kurzrock, T., Schallinger, S., & Weuster-Botz, D. (2011). Integrated separation process for isolation and purification of biosuccinic acid. *Biotechnology Progress*, 27(6), 1623–1628. <http://doi.org/10.1002/btpr.673>
- Larson, E. D., Williams, R. H., & Leal, M. R. L. V. (2001). A review of biomass integrated-gasifier/gas turbine combined cycle technology and its application in sugarcane industries, with an analysis for Cuba. *Energy for Sustainable Development*, 5(1), 54–76. [http://doi.org/10.1016/S0973-0826\(09\)60021-1](http://doi.org/10.1016/S0973-0826(09)60021-1)
- Lazzaretto, A., & Tsatsaronis, G. (2006). SPECO: A systematic and general methodology for calculating efficiencies and costs in thermal systems. *Energy*, 31(8–9), 1257–1289. <http://doi.org/10.1016/j.energy.2005.03.011>
- Lecompte, S., Huisseune, H., Van Den Broek, M., Vanslambrouck, B., & De Paepe, M. (2015). Review of organic Rankine cycle (ORC) architectures for waste heat recovery. *Renewable and Sustainable Energy Reviews*, 47, 448–461. <http://doi.org/10.1016/j.rser.2015.03.089>
- Lehnberger, A., Brahim, F., & Mallikarjun, S. S. (2014). Falling-film evaporator plant for a cane sugar factory : Presentation of the concept and operating results. *International Sugar Journal*, 116(August), 604–609. Retrieved from www.internationalsugarjournal.com
- Li, S., Shen, Y., Liu, D., Fan, L., Zhang, Z., & Li, W. (2014). A novel orange juice concentration method based on C₂H₄ clathrate hydrate formation. *Advanced Journal of Food Science Technology*, 6(6), 1–13.
- Liu, R., Liang, L., Cao, W., Wu, M., Chen, K., Ma, J., Ouyang, P. (2013). Succinate production by metabolically engineered *Escherichia coli* using sugarcane bagasse hydrolysate as the carbon source. *Bioresource Technology*, 135, 574–577. <http://doi.org/10.1016/J.BIORTECH.2012.08.120>
- Liu, Y., Zheng, P., Sun, Z., Ni, Y., Dong, J., & Zhu, L. (2008). Economical succinic acid

- production from cane molasses by *Actinobacillus succinogenes*. *Bioresource Technology*, 99(6), 1736–1742. <http://doi.org/10.1016/j.biortech.2007.03.044>
- Ma, J., Li, F., Liu, R., Liang, L., Ji, Y., Wei, C., Ouyang, P. (2014). Succinic acid production from sucrose and molasses by metabolically engineered *E. coli* using a cell surface display system. *Biochemical Engineering Journal*, 91, 240–249. <http://doi.org/10.1016/j.bej.2014.08.014>
- Madaeni, S. S., & Zereshki, S. (2010). Energy consumption for sugar manufacturing. Part I: Evaporation versus reverse osmosis. *Energy Conversion and Management*, 51(6), 1270–1276. <http://doi.org/10.1016/j.enconman.2010.01.002>
- Mahamud, R., Khan, M. M. K., Rasul, M. G., & Leinster, M. G. (2013). Exergy Analysis and Efficiency Improvement of a Coal Fired Thermal Power Plant in Queensland. *Thermal Power Plants - Advanced Applications*, 3–28.
- Mandegari, M. A., Farzad, S., & Görgens, J. F. (2017). Recent trends on techno-economic assessment (TEA) of sugarcane biorefineries. *Biofuel Research Journal*, 4(3), 704–712. <http://doi.org/10.18331/BRJ2017.4.3.7>
- Mandegari, M., Farzad, S., & Görgens, J. F. (2018). A new insight into sugarcane biorefineries with fossil fuel co-combustion: Techno-economic analysis and life cycle assessment. *Energy Conversion and Management*, 165, 76–91. <http://doi.org/10.1016/j.enconman.2018.03.057>
- Martínez-Guido, S. I., Betzabe González-Campos, J., Ponce-Ortega, J. M., Nápoles-Rivera, F., & El-Halwagi, M. M. (2016). Optimal reconfiguration of a sugar cane industry to yield an integrated biorefinery. *Clean Technologies and Environmental Policy*, 18(2), 553–562. <http://doi.org/10.1007/s10098-015-1039-1>
- Mashoko, L., Mbohwa, C., & Thomas, V. M. (2010). LCA of the South African sugar industry. *Journal of Environmental Planning and Management*, 53(6), 793–807. <http://doi.org/10.1080/09640568.2010.488120>
- Mashoko, L., Mbohwa, C., & Thomas, V. M. (2013). Life cycle inventory of electricity cogeneration from bagasse in the South African sugar industry. *Journal of Cleaner Production*, 39, 42–49. <http://doi.org/10.1016/j.jclepro.2012.08.034>
- Matelli, J. A., Perrella Perrella, J. A., & Fonseca Filho, V. F. d. (2016). Carbon exergy tax

- applied to biomass integrated gasification combined cycle in sugarcane industry. *Energy*, 103, 715–724. <http://doi.org/10.1016/j.energy.2016.02.161>
- Mbohwa, C. (2013). Energy Management in the South African Sugar Industry. *Proceedings of the World Congress on Engineering, I(Vol I)*, 3–8.
- McKinlay, J. B., Vieille, C., & Zeikus, J. G. (2007). Prospects for a bio-based succinate industry. *Applied Microbiology and Biotechnology*, 76(4), 727–740. <http://doi.org/10.1007/s00253-007-1057-y>
- Meadows, D. M., Schumann, G. T., & Soji, C. (1998). Farewell to filters: The recycle of clarifier mud to the diffuser. *Proc S Afr Sug Technol Ass*, 72, 198–203.
- Menon, V., & Rao, M. (2012). Trends in bioconversion of lignocellulose: Biofuels, platform chemicals & biorefinery concept. *Progress in Energy and Combustion Science*, 38(4), 522–550. <http://doi.org/10.1016/j.pecs.2012.02.002>
- Meyer, J., Rein, P., Turner, P., & Mathias, K. (2011). *Good management practices manual for the cane sugar industry (Final)*. Johannesburg, South Africa.
- Modesto, M., Zemp, R. J., & Nebra, S. A. (2009). Ethanol production from sugar cane: Assessing the possibilities of improving energy efficiency through exergetic cost analysis. *Heat Transfer Engineering*, 30(4), 272–281. <http://doi.org/10.1080/01457630802380386>
- Moncada, J., El-Halwagi, M. M., & Cardona, C. A. (2013). Techno-economic analysis for a sugarcane biorefinery: Colombian case. *Bioresource Technology*, 135, 533–543. <http://doi.org/10.1016/J.BIORTECH.2012.08.137>
- Morales, M., Ataman, M., Badr, S., Linster, S., Kourlimpinis, I., Papadokonstantakis, S., Hungerbühler, K. (2016). Sustainability assessment of succinic acid production technologies from biomass using metabolic engineering. *Energy & Environmental Science*, 9(9), 2794–2805. <http://doi.org/10.1039/C6EE00634E>
- Moya, C., Domínguez, R., Van Langenhove, H., Herrero, S., Gil, P., Ledón, C., & Dewulf, J. (2013). Exergetic analysis in cane sugar production in combination with Life Cycle Assessment. *Journal of Cleaner Production*, 59, 43–50. <http://doi.org/10.1016/j.jclepro.2013.06.028>
- Mussatto, S. I., Aguiar, L. M., Marinha, M. I., Jorge, R. C., & Ferreira, E. C. (2015). Economic analysis and environmental impact assessment of three different fermentation processes

- for fructooligosaccharides production. *Bioresource Technology*, 198, 673–681. <http://doi.org/10.1016/j.biortech.2015.09.060>
- Narasimha Rao, G. R., & Nagarajan, M. (2012). Energy Efficiency in Indian Sugar Industries. *FAPCCI Review*, 9–15.
- Nghiem, N., Kleff, S., & Schwegmann, S. (2017). Succinic Acid: Technology Development and Commercialization. *Fermentation*, 3(2), 26. <http://doi.org/10.3390/fermentation3020026>
- Nieder-Heitmann, M., Haigh, K., & Görgens, J. F. (2019). Process design and economic evaluation of integrated, multi-product biorefineries for the co-production of bio-energy, succinic acid, and polyhydroxybutyrate (PHB) from sugarcane bagasse and trash lignocelluloses. *Biofuels, Bioproducts and Biorefining*, 13(3), 599–617. <http://doi.org/10.1002/bbb.1972>
- Nobre, C., Castro, C. C., Hantson, A.-L., Teixeira, J. A., De Weireld, G., & Rodrigues, L. R. (2016). Strategies for the production of high-content fructo-oligosaccharides through the removal of small saccharides by co-culture or successive fermentation with yeast. *Carbohydrate Polymers*, 136, 274–281. <http://doi.org/10.1016/j.carbpol.2015.08.088>
- Nobre, C., Teixeira, J. A., & Rodrigues, L. R. (2015). New Trends and Technological Challenges in the Industrial Production and Purification of Fructo-oligosaccharides. *Critical Reviews in Food Science and Nutrition*, 55(10), 1444–1455. <http://doi.org/10.1080/10408398.2012.697082>
- O'Hara, I. M., & Mundree, S. G. (2016). *Sugarcane-Based Biofuels and Bioproducts*. Retrieved from www.copyright.com. <http://www.wiley.com/go/permission>. <http://lccn.loc.gov/2016007511>
- OECD/FAO. (2016). Sugar. In *OECD-FAO Agricultural Outlook 2016-2025* (pp. 1–13). Paris: OECD Publishing, Paris. http://doi.org/10.1787/agr_outlook-2016-9-en
- Ogden, J. M., Hochgreb, S., & Hylton, M. (1989). Process energy efficiency and cogeneration in cane sugar factories. *Energy*, 441–451. Retrieved from <http://www.princeton.edu/pei/energy/publications/texts/>
- Ogden, J. M., Hochgreb, S., & Hylton, M. (1990). Steam economy and cogeneration in cane

- sugar factories. *International Sugar Journal*, 92(1099), 131–143.
- Ojeda, K., Ávila, O., Suárez, J., & Kafarov, V. (2010). Evaluation of technological alternatives for process integration of sugarcane bagasse for sustainable biofuels production — Part 1. *Chemical Engineering Research and Design*, 89(3), 270–279. <http://doi.org/10.1016/j.cherd.2010.07.007>
- Ojeda, K., Sánchez, E., & Kafarov, V. (2011). Sustainable ethanol production from lignocellulosic biomass - Application of exergy analysis. *Energy*, 36(4), 2119–2128. <http://doi.org/10.1016/j.energy.2010.08.017>
- Pachón, E. R., Vaskan, P., Raman, J. K., & Gnansounou, E. (2018). Transition of a South African sugar mill towards a biorefinery. A feasibility assessment. *Applied Energy*, 229, 1–17. <http://doi.org/10.1016/J.APENERGY.2018.07.104>
- Pateraki, C., Almqvist, H., Ladakis, D., Lidén, G., Koutinas, A. A., & Vlysidis, A. (2016). Modelling succinic acid fermentation using a xylose based substrate. *Biochemical Engineering Journal*, 114, 26–41. <http://doi.org/10.1016/j.bej.2016.06.011>
- Pellegrini, L. F., & de Oliveira Jr, S. (2011). Combined production of sugar, ethanol and electricity: Thermoeconomic and environmental analysis and optimization. *Energy*, 36(6), 3704–3715. <http://doi.org/10.1016/j.energy.2010.08.011>
- Pellegrini, L. F., de Oliveira Jr, S., & Burbano, J. C. (2010). Supercritical steam cycles and biomass integrated gasification combined cycles for sugarcane mills. *Energy*, 35(2), 1172–1180. <http://doi.org/10.1016/j.energy.2009.06.011>
- Petersen, A. M., Van der Westhuizen, W. A., Mandegari, M. A., & Görgens, J. F. (2018). Economic analysis of bioethanol and electricity production from sugarcane in South Africa. *Biofuels, Bioproducts and Biorefining*, 12(2), 224–238. <http://doi.org/10.1002/bbb.1833>
- Pinazo, J. M., Domine, M. E., Parvulescu, V., & Petru, F. (2015). Sustainability metrics for succinic acid production: A comparison between biomass-based and petrochemical routes. *Catalysis Today*, 239, 17–24. <http://doi.org/10.1016/J.CATTOD.2014.05.035>
- Purwanto, Y. A., Oshita, S., Seo, Y., & Kawagoe, Y. (2001). Concentration of liquid foods by the use of gas hydrate. *Journal of Food Engineering*, 47, 133–138. [http://doi.org/10.1016/S0260-8774\(00\)00109-6](http://doi.org/10.1016/S0260-8774(00)00109-6)

- Purwanto, Y. A., Oshita, S., Seo, Y., & Kawagoe, Y. (2014). Separation process of nonpolar gas hydrate in food solution under high pressure apparatus. *International Journal of Chemical Engineering*, 2014, 8.
- Querol, E., Gonzalez-Regueral, B., & Perez-Benedito, J. L. (2013). Exergy Concept and Determination. In *Practical Approach to Exergy and Thermo-economic Analyses of Industrial Processes* (pp. 9–28). SpringerBriefs in Energy. http://doi.org/10.1007/978-1-4471-4622-3_2
- Querol, E., Gonzalez-Regueral, B., Ramos, A., & Perez-Benedito, J. L. (2011). Novel application for exergy and thermo-economic analysis of processes simulated with Aspen Plus®. *Energy*, 36(2), 964–974. <http://doi.org/10.1016/j.energy.2010.12.013>
- Rajnauth, J., Barrufet, M., & Falcone, G. (2013). Potential industry applications using gas hydrate technology. *The West Indian Journal of Engineering*, 35(2), 15–21.
- Rein, P. (2007). *Cane Sugar Engineering*. (A. Bartens, Ed.). Berlin.
- Rosen, M. A. (2002). Clarifying thermodynamic efficiencies and losses via exergy. *Exergy, An International Journal*, 2(1), 3–5. [http://doi.org/10.1016/S1164-0235\(01\)00054-1](http://doi.org/10.1016/S1164-0235(01)00054-1)
- Rosen, M. A., & Bulucea, C. A. (2009). Using exergy to understand and improve the efficiency of electrical power technologies. *Entropy*, 11(4), 820–835. <http://doi.org/10.3390/e11040820>
- Sá Gomes, P., & Rodrigues, A. E. (2012). Simulated Moving Bed Chromatography: From Concept to Proof-of-Concept. *Chemical Engineering & Technology*, 35(1), 17–34. <http://doi.org/10.1002/ceat.201100281>
- Şahin, H. M., Acir, A., Altunok, T., Baysal, E., & Koçyiğit, E. (2010). Analysis of exergy and energy of sugar production process in sugar plant. *Journal of the Energy Institute*, 83(3), 178–185. <http://doi.org/10.1179/014426010X12759937396911>
- Saidur, R., Ahamed, J. U., & Masjuki, H. H. (2010). Energy, exergy and economic analysis of industrial boilers. *Energy Policy*, 38(5), 2188–2197. <http://doi.org/10.1016/j.enpol.2009.11.087>
- Salvachúa, D., Mohagheghi, A., Smith, H., Bradfield, M. F. A., Nicol, W., Black, B. A., Beckham, G. T. (2016). Succinic acid production on xylose-enriched biorefinery streams by *Actinobacillus succinogenes* in batch fermentation. *Biotechnology for Biofuels*, 9(1),

28. <http://doi.org/10.1186/s13068-016-0425-1>

Santos, C. I., Silva, C. C., Mussatto, S. I., Osseweijer, P., van der Wielen, L. A. M., & Posada, J. A. (2017). Integrated 1st and 2nd generation sugarcane bio-refinery for jet fuel production in Brazil: Techno-economic and greenhouse gas emissions assessment. *Renewable Energy*. <http://doi.org/10.1016/j.renene.2017.05.011>

Sciubba, E. (2014). A Critical interpretation and quantitative extension of the Sama-Szargut second law rules in an extended exergy perspective. *Energies*, 7(8), 5357–5373. <http://doi.org/10.3390/en7085357>

Seebaluck, V., Mohee, R., Sobhanbabu, P. R. K., Leal, M. R. L. V., & Johnson, F. X. (2008). *Bioenergy for sustainable development and global competitiveness : The case of sugar cane in Southern Africa*. Stockholm, Sweden.

Sheu, D. C., Lio, P. J., Chen, S. T., Lin, C. T., & Duan, K. J. (2001). Production of fructooligosaccharides in high yield using a mixed enzyme system of β -fructofuranosidase and glucose oxidase. *Biotechnology Letters*, 23, 1499–1503. Retrieved from <https://link.springer.com/article/10.1023/A:1011689531625>

Shin, H. (2004). Production of fructo-oligosaccharides from molasses by *Aureobasidium pullulans* cells. *Bioresource Technology*, 93(1), 59–62. <http://doi.org/10.1016/j.biortech.2003.10.008>

Sogut, Z., Ilten, N., & Oktay, Z. (2010). Energetic and exergetic performance evaluation of the quadruple-effect evaporator unit in tomato paste production. *Energy*, 35(9), 3821–3826. <http://doi.org/10.1016/j.energy.2010.05.035>

Soltanian, S., Aghbashlo, M., Farzad, S., Tabatabaei, M., Mandegari, M., & Görgens, J. F. (2019). Exergoeconomic analysis of lactic acid and power cogeneration from sugarcane residues through a biorefinery approach. *Renewable Energy*, 143, 872–889. <http://doi.org/10.1016/j.renene.2019.05.016>

Soni, M. G., & Tsai, H. (2016). GRAS notification for Fructooligosaccharides. *CFR*, 17036. Retrieved from <https://www.fda.gov/downloads/Food/IngredientsPackagingLabeling/GRAS/NoticeInventory/ucm504609.pdf>

Starzak, M. (2016). *Biorefinery techno-economic modelling: Validation of a standalone sugar*

mill model using factory data.

- Starzak, M., & Davis, S. (2015). Mass and Energy Balance Modelling of a Sugar Mill . A comparison of MATLAB ® and Association Andrew van Hook. In *22nd International Symposium* (pp. 1–22). Mauritius.
- Starzak, M., & Davis, S. (2016). MATLAB ® modelling of a sugar mill : model development and validation. In *Proc S Afr Sug Technol Ass* (pp. 517–536).
- Starzak, M., & Zizhou, N. (2015). *Biorefinery techno-economic modelling: Sugar mill and ethanol distillery process model*. Durban.
- Taner, T., & Sivrioglu, M. (2015). Energy-exergy analysis and optimisation of a model sugar factory in Turkey. *Energy*, *93*, 641–654. <http://doi.org/10.1016/j.energy.2015.09.007>
- Taylor, R., Natrass, L., Alberts, G., Robson, P., Chudziak, C., Bauen, A., Ree, R. van. (2015). *From the Sugar Platform to biofuels and biochemicals*. Retrieved from [https://ec.europa.eu/energy/sites/ener/files/documents/EC Sugar Platform final report.pdf](https://ec.europa.eu/energy/sites/ener/files/documents/EC_Sugar_Platform_final_report.pdf)
- Tekin, T., & Bayramoğlu, M. (1998). Exergy analysis of the sugar production process from sugar beets. *International Journal of Energy Research*, *22*(7), 591–601. [http://doi.org/10.1002/\(SICI\)1099-114X\(19980610\)22:7<591::AID-ER360>3.0.CO;2-D](http://doi.org/10.1002/(SICI)1099-114X(19980610)22:7<591::AID-ER360>3.0.CO;2-D)
- Thakker, C., Martínez, I., San, K.-Y., & Bennett, G. N. (2012). Succinate production in *Escherichia coli*. *Biotechnology Journal*, *7*(2), 213–24. <http://doi.org/10.1002/biot.201100061>
- Tsatsaronis, G. (2007). Definitions and nomenclature in exergy analysis and exergoeconomics. *Energy*, *32*(4), 249–253. <http://doi.org/10.1016/j.energy.2006.07.002>
- Tsatsaronis, G., & Cziesla, F. (1999). Strengths and limitations of exergy analysis. In *Thermodynamic Optimization of Complex Energy Systems* (Vol. I, pp. 93–100). Dordrecht: Springer Netherlands. http://doi.org/10.1007/978-94-011-4685-2_6
- Tsatsaronis, G., & Winhold, M. (1985). Exergoeconomic analysis and evaluation of energy-conversion plants—I. A new general methodology. *Energy*, *10*(1), 69–80. [http://doi.org/10.1016/0360-5442\(85\)90020-9](http://doi.org/10.1016/0360-5442(85)90020-9)
- U.S. Department of Energy. (2011). *Cooling towers : Understanding key components of cooling towers and how to improve water efficiency*. FEMP, Washington.

- Ur Rehman, A., Kovacs, Z., Quitmann, H., Ebrahimi, M., & Czermak, P. (2016). Enzymatic production of fructo-oligosaccharides from inexpensive and abundant substrates using a membrane reactor system. *Separation Science and Technology*, 01496395.2016.1167740. <http://doi.org/10.1080/01496395.2016.1167740>
- van Ree, R., & de Jong, E. (2013). *Biorefineries: Co-production of fuels, chemicals, power and materials from biomass*. Cape Town, South Africa. Retrieved from http://www.iea-bioenergy.task42-biorefineries.com/upload_mm/b/f/a/4898a5a0-ed3b-435e-bec3-17d297957f60_Final Report IEA Bioenergy Task42 Triennium 2010 2012.pdf
- Vaňková, K., Onderková, Z., Antošová, M., & Polakovič, M. (2008). Design and economics of industrial production of fructooligosaccharides. *Chemical Papers*, 62(4), 375–381. <http://doi.org/10.2478/s11696-008-0034-y>
- Vaňková, K., & Polakovič, M. (2012). Design of Fructooligosaccharide Separation Using Simulated Moving-Bed Chromatography. *Chemical Engineering & Technology*, 35(1), 161–168. <http://doi.org/10.1002/ceat.201100254>
- Vaswani, S. (2010). *Bio-based succinic acid*. Menlo Park, California 94025. Retrieved from https://www.ihs.com/pdf/RW2010-14_220240110917062932.pdf
- Wall, G. (2009). *Exergetics*. Bucaramanga: Exergy, Ecology, Democracy.
- Xie, Y., Zhou, H., Liu, C., Zhang, J., Li, N., Zhao, Z., Zhong, Y. (2017). A molasses habitat-derived fungus *Aspergillus tubingensis* XG21 with high β -fructofuranosidase activity and its potential use for fructooligosaccharides production. *AMB Express*, 7(1), 128. <http://doi.org/10.1186/s13568-017-0428-8>
- Zeikus, J. G., Jain, M. K., & Elankovan, P. (1999). Biotechnology of succinic acid production and markets for derived industrial products. *Applied Microbiology and Biotechnology*, 51(5), 545–552. <http://doi.org/10.1007/s002530051431>
- Zhang, C., Yang, M., Lu, M., Shan, Y., & Zhu, J. (2011). Experimental research on LiBr refrigeration – Heat pump system applied in CCHP system. *Applied Thermal Engineering*, 31(17–18), 3706–3712. <http://doi.org/10.1016/j.applthermaleng.2011.02.004>
- Zhao, Y., Cao, W., Wang, Z., Zhang, B., Chen, K., & Ouyang, P. (2016). Enhanced succinic acid production from corncob hydrolysate by microbial electrolysis cells. *Bioresource Technology*, 202, 152–157. <http://doi.org/10.1016/j.biortech.2015.12.002>

Chapter 3

Exergetic diagnosis and performance analysis of a typical sugar mill based on Aspen Plus® simulation of the process

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Objective of dissertation in this chapter and summary of findings

The exergy analysis methodology reported for the true thermodynamic assessment of thermal processes was applied to evaluate the energy inefficiencies of a typical South African sugarcane mill (objective 1) for improvement. Previous sugar industry exergy studies were mostly on beet sugar production processes, and combined ethanol and cane sugar mills, using factory data and numerical equations. Moreover, the previous studies mostly focused on the cogeneration system, as the principal exergy destruction unit in a sugar mill. The extensive simulation-based exergy analysis focusing on the production processes of raw sugar from sugarcane has not been reported, which is, therefore, the aim of this chapter.

The study was based on the Aspen plus® simulation of a typical South African sugar mill processes including juice extraction, clarification, evaporation, crystallization, and drying units as well as the cooling tower. The exergy methodology was developed basically to evaluate sources of inefficiencies and prioritize inefficient units for improvement. The cogeneration unit, which has received much attention in previous studies was assessed separately in Chapter 4.

The key outcome was that the crystallization has the lowest exergy efficiency due to the complexity of the three-stage crystallization in the current sugar mill. Consequently, the crystallization unit had the highest potential for improvement, closely followed by the

evaporation unit. Therefore, a simpler crystallization process such as a single-stage crystallization was recommended to improve its efficiency.

The results and principles obtained from this Chapter were applied in the selection of technologies to be integrated into the sugar mill to improve to improve the efficiency (Chapter 4) while the single-stage crystallization resulting in a sucrose-rich A-molasses as by-product informs the valorisation of A-molasses for integrated biorefineries towards economic sustainability of the mill (Chapter 5). Moreover, since exergy since a trade-off always exists between exergy improvements and the economics of doing so, exergoeconomics was recommended to evaluate the suggested exergy solution before their implementation. This was explored in Chapter 6.

Declaration by the candidate

With regards to Chapter 3, pages 72 – 110, the nature and scope of my contribution were as follows:

Nature of contribution	Extent of contribution (%)
Project and scope definition, analysis of data, interpretation of results and writing of manuscript	80

The following co-authors have contributed to Chapter 3, pages 72 – 110:

Name	e-mail address	Nature of contribution	Extent of contribution (%)
J.F. Görgens	jgorgens@sun.ac.za	Provided writing assistance through review and proof reading of manuscript and general discussion	5
M.A. Mandegari	mandegari@sun.ac.za	Provided writing assistance through scope suggestions, continuous review and proof reading of manuscript and general discussions	15

Signature of candidate:

Date:

Declaration by co-authors:

The undersigned hereby confirm that

1. the declaration above accurately reflects the nature and extent of the contributions of the candidate and the co-authors to Chapter 3, pages 72 – 110,
2. no other authors contributed to Chapter 3, pages 72 – 110 besides those specified above, and
3. potential conflicts of interest have been revealed to all interested parties and that the necessary arrangements have been made to use the material in Chapter 3, pages 72 – 110, of this dissertation.

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Exergetic diagnosis and performance analysis of a typical sugar mill based on Aspen Plus[®] simulation of the process

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Abstract

The sugar industry is the second largest agro-industry in the world, with more than 80 % of sugar produced from sugarcane. Sugar mills are energy-intensive and historically not designed to be energy efficient, even though they may be energy self-sufficient. This study presents a comprehensive exergy analysis of cane sugar production processes to identify inefficient components for improvement. The exergy analysis was based on rigorous mass and energy balances calculated from an Aspen Plus[®] simulation of a typical 250 ton per hour sugar mill, along with an appropriate exergy methodology. The exergy analysis of the cogeneration system, which has been found to be the principal sugar mill exergy destruction unit, is conducted separately and will be presented in a subsequent paper. The overall sugar mill irreversibility and functional exergy efficiency were 217.3 MJ per ton of cane crushed and 9.7 %, respectively. The evaporation unit generated the highest irreversibility of 100 MJ/ton of cane, while crystallization unit had the lowest functional exergy efficiency of 9.6 % and the highest potential for improvement of 47.0 MJ/ton of cane. The exergetic performance of the mill may be improved by adopting a single stage crystallization with an integrated biorefinery.

Keywords: Sugarcane mill; Exergy analysis; Functional exergy efficiency; Exergetic improvement potential; Grassmann diagram; Aspen Plus[®].

3.1. Introduction

The sugar industry is an important agro-industry in the world. Its raw materials, sugarcane and sugar beet are produced in more than 130 countries. Sugarcane, the world's largest cash crop is primarily grown for sugar production in over 90 countries around the world, and account for about 80% of global sugar production (International Sugar Organization, 2016; Pippo & Luengo, 2013). World cane sugar production in 2015 was more than 75 million metric tons

(International Sugar Organization, 2016), with energy consumption estimated at 325 – 550 kg steam /ton of cane and 25 – 32 kWh electricity /ton of cane (Seebaluck et al., 2008). In recent times, the cane sugar industry has turned toward increasing production of economical by-products such as ethanol and electricity for sale. The cane residue (bagasse), which is the main source of sugar mill energy, is also the feedstock for electricity and several bio-chemicals, whose market sizes are expected to increase. Therefore, improving sugar mill energy efficiency will help save lignocellulosic biomass residues for such further valorisation through biorefineries (Ali Mandegari et al., 2017; Somayeh Farzad, Mandegari, Guo, et al., 2017).

Global challenges with environmental degradation and limited fossil reserves have called for the sustainable use of energy, which demands a thorough assessment of energy use in energy-intensive systems, in order to improve them (de Oliveira Jr, 2013). Such assessments also enhance the sustainability and profitability of the production processes. The most commonly used method for this purpose is the conventional energy analysis based on the first law of thermodynamics, which only accounts for the quantity of energy through processes and not the quality (Hevert & Hevert, 1980).

Exergy analysis, based on the second law of thermodynamics, on the other hand adequately identifies locations, magnitudes, and causes of thermodynamic inefficiencies in energy conversion processes (Aghbashlo et al., 2016). Exergy quantifies the maximum possible work that can be obtained from a reversible process when a system is brought to equilibrium with its environment. Process improvements through exergy analysis lead to the sustainable use of limited natural resources, and hence the sustainability of process industry and the environment (Aghbashlo et al., 2016; Luis & Van der Bruggen, 2014). Exergy analysis has been widely studied for the design, evaluation and optimization of energy conversion systems in industries such as the cement (Atmaca & Yumrutaş, 2014b), power (Kamate & Gangavati, 2009b), pulp and paper (Assari, Basirat, Najafpour, Ahmadi, & Jafari, 2014), steel (Costa, Schaeffer, & Worrell, 2001), chemical (Hosseini, Aghbashlo, Tabatabaei, Younesi, & Najafpour, 2015) and food (Dowlati, Aghbashlo, & Mojarab, 2017) production.

Previous reports on sugar mill exergy analysis were based on factory data and mostly focused on beet sugar production, with different processes to those used in traditional raw sugarcane processes, such as chemical clarification. Tekin and Bayramoğlu (1998a) applied exergy analysis to a factory producing white refined sugar from beet and found that the steam power plant accounted for 74.4 % of the total plant exergy losses, mainly due to the irreversible

combustion reaction and waste exergy of combustion products. They suggested thermal recuperation in the steam power plant as a measure to improve the factory's exergetic performance. Bayrak et al (2003) conducted energy and exergy analyses on beet sugar production stages, comprising juice production, clarification, concentration and crystallization. The juice production stage had the lowest exergy efficiency, while the crystallization stage recorded the highest exergy loss due to heat transfer over high temperature differences. They concluded that minimizing energy loss and heat transfer temperature differences will reduce process irreversibility, resulting in a more efficient sugar production. Sahin et al (2010) also conducted exergy analysis on an old beet sugar factory in Turkey, with the aim of rebuilding the factory, and obtained similar results as that obtained by Bayrak et al. The most recent beet sugar exergy study by Taner and Sivrioglu (2015) focused on optimizing the power plant, which was the most inefficient component of the factory.

The few exergy studies of the sugarcane industry reported in literature focus either on the co-production of sugar and ethanol (A. V. Ensinas et al., 2009; Héctor et al., 2013), with results showing the cogeneration unit as the least efficient in exergy terms, or a stand-alone cogeneration plant (Kamate & Gangavati, 2009b). Recently, Garg, Sharma, and Sharma (Garg et al., 2016) evaluated the exergy efficiencies of only the juice production (59.27 %) and juice clarification (71.23 %) processes of a sugarcane plant, claiming both have large exergy losses that should be minimized. To the authors' best knowledge, no complete and simulation-based exergy analysis of raw sugar production processes from sugarcane has been reported in open literature. Therefore, the aim of this study is to conduct a comprehensive exergy analysis of a typical sugarcane mill producing raw sugar based on an Aspen Plus® simulation of the process.

This paper applies a comprehensive exergy methodology to assess actual thermodynamic performances of raw cane sugar production processes such as extraction, clarification, evaporation, crystallization, and drying, as well as the water cooling process, and to rank them for improvement. Cogeneration system is not included in this paper since it has received considerable attention as the major exergy destructive unit in a sugar mill (A. V. Ensinas et al., 2009; Kamate & Gangavati, 2009b; Taner & Sivrioglu, 2015, 2017; T Tekin & Bayramoğlu, 1998a).

3.2. Materials and methods

3.2.1. Process description

Sugar production involves a series of physical processes including juice extraction, clarification, evaporation, crystallization, and drying. In addition to the main process units, a typical sugar mill has a cogeneration system where process steam and power are generated, and a cooling tower for cooling process warm water. Figure 3-1 depicts a block diagram of the main production processes, with the main resource and products. Figure 3-2 shows the process flow diagram (including the cooling tower).

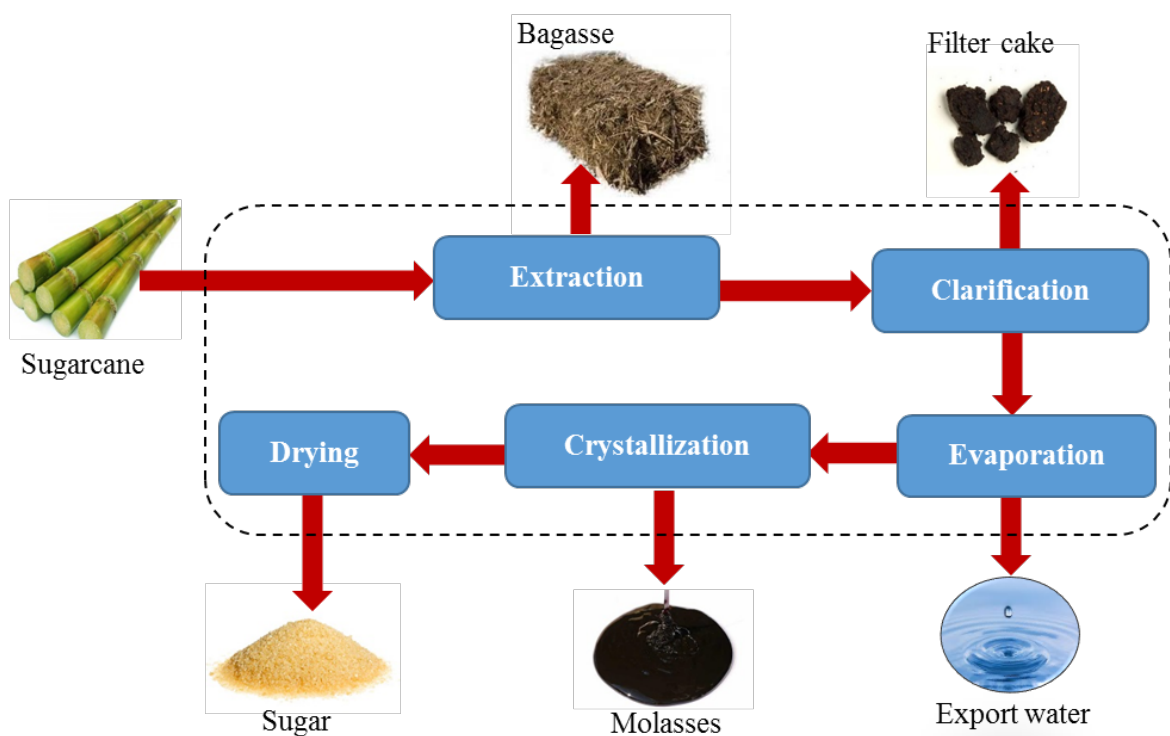


Figure 3-1: Process block diagram of sugar mill

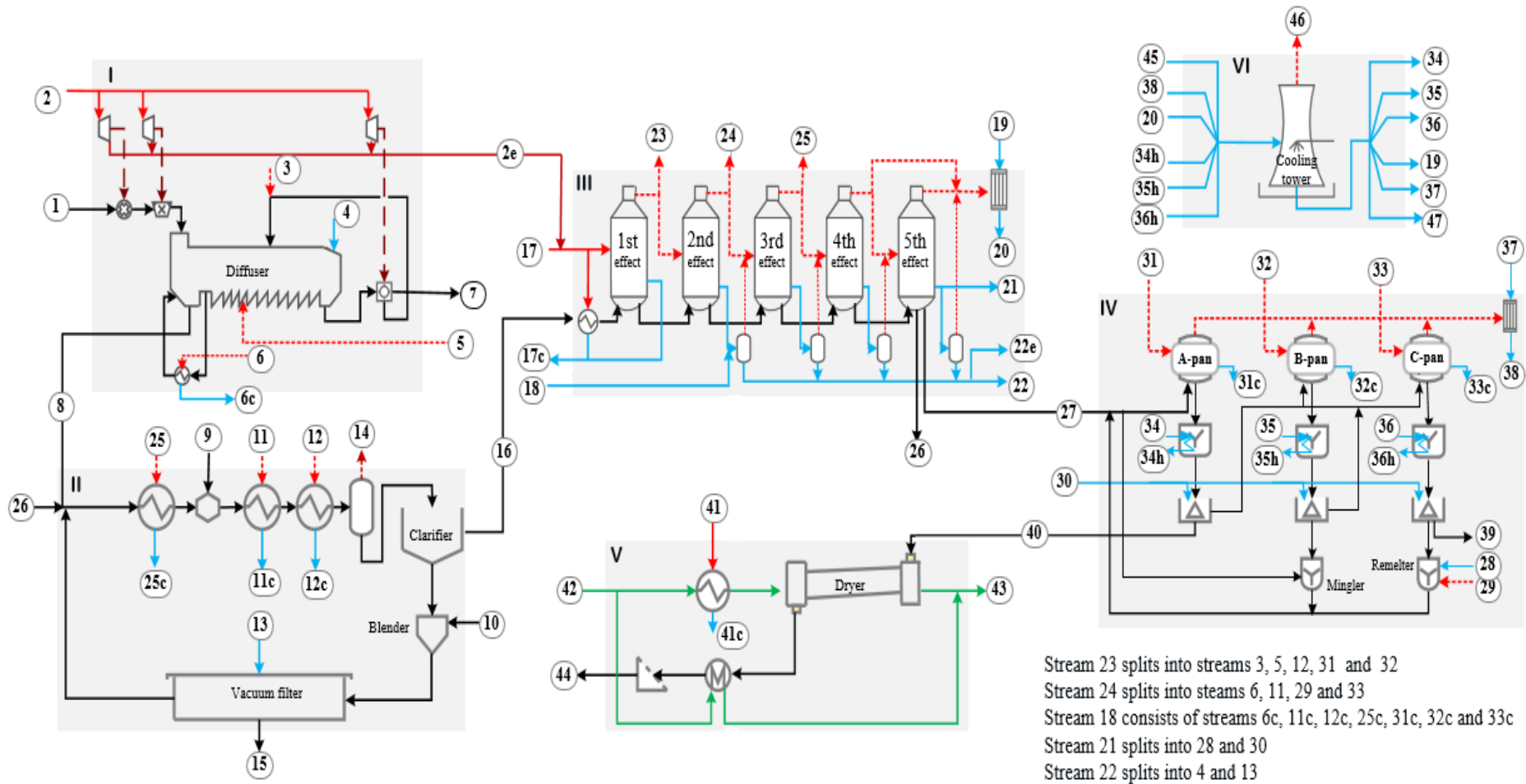


Figure 3-2: Process flow diagram of sugar mill process units; I – extraction; II – clarification; III – evaporation; IV – crystallization; V – drying; VI – cooling tower

Juice extraction: Sugarcane from the farm is first cut and shredded using knives and shredders to break the hard structure and rupture sucrose cell for easy juice extraction (Hugot, 1972; Seebaluck et al., 2008). Juice extraction is usually done either by compressing shredded cane in roller mills to squeeze out the juice (milling method), or by the diffusion method through a counter-current hot water leaching process at a suitable temperature (about 70 – 75 °C), to keep a balance between maintaining low microorganisms growth and low sucrose inversion (Seebaluck et al., 2008). Milling requires about 60–65 % of sugar mill energy requirement, consequently, diffusion is preferred due to its lower energy consumption, even though it has higher imbibition water volumes than milling (Hugot, 1972; Mbohwa, 2013). Figure 3-2 (I) depicts a continuous diffusion process (Gurumurthy, 2011; Seebaluck et al., 2008) where extracted weak juice and imbibition water are passed over the prepared cane in a counter-current manner to ensure maximum juice extraction (Rein, 2007). Two sets of bagasse (dewatering) mills accompany the diffuser to help further juice extraction and bring bagasse moisture to about 50% (Modesto et al., 2009).

Juice clarification: In the clarification unit shown in Figure 3-2 (II) the draft juice is first heated and limed to enhance precipitation and separation of some non-sucrose impurities and fine bagasse from the juice. Liming also provides a nearly neutral pH, suitable for clarification and restricting sucrose inversion in subsequent evaporation. Floc settling is enhanced by flashing juice in flash drums to remove dissolved air from the mixture, prior to clarifiers (Rein, 2007; Starzak & Davis, 2016). A suspension of precipitated impurities (mud) in clear juice settles at the bottom of clarifiers and is filtered in a rotary vacuum filter to recover juice, leaving a filter cake. Alternatively, the filtration step could be eliminated by recycling the mud back to the diffuser, with the cane bed serving as a filter (Gurumurthy, 2011; Meadows et al., 1998).

Juice concentration (evaporation): Clarified juice contains about 85 % (by weight) of water (A. V Ensinas et al., 2006; Higa et al., 2009), most of which is removed by boiling, to concentrate the sugar. The evaporation unit, shown in Figure 3-2 (III) mainly comprises a heat exchanger to preheat clear juice to about 110 °C, a five-effect evaporator train and a barometric condenser. The heat exchangers and first-effect evaporator make use of exhaust steam from steam turbines, while the remaining effects utilize vapour from preceding effects, to concentrate clarified juice to approx. 65–68 °Brix² (Rein, 2007). The evaporator train operates with decreasing pressure – from exhaust steam pressure of about 2 bar in the first effect calandria (steam side) to about 0.15 bar (vacuum) in the body (juice side) of the last effect, to prevent sucrose inversion (Rein, 2007). The reduced pressure in each

² Brix is juice concentration expressed as grams of solids per 100 g of water

successive effect allows the temperature differences required for evaporation to be established (A. V. Ensinas et al., 2009). Vacuum vapours from the last effect evaporators are condensed in a barometric condenser, which also maintains the vacuum in the system.

Crystallization and centrifugation: The crystallization process starts by boiling syrup (concentrated juice) in conventional batch or continuous vacuum pans while adding a slurry of fine sugar crystals to seed crystals formation. The hot mixture of crystals and the mother liquor (molasses), called massecuite is then discharged from the vacuum pan into cooling crystallizers for further crystallization. Subsequently, the sugar crystals are filtered and washed out of the liquor in centrifuges. Vapours produced in the vacuum pans are condensed in a barometric condenser, to maintain the system vacuum. To ensure minimal loss of sugar in the final molasses, the whole crystallization process is carried out in three stages (A, B and C), as shown in Figure 3-2 (IV). Lower grade sugars produced from the B and C centrifuges are either melted (dissolved in water or clarified juice) in the remelter or mingled to serve as seed in the vacuum pans, while A and B molasses are recycled to the B and C vacuum pans (Starzak & Davis, 2016). Finally, C molasses comes out as a by-product, while the moist and hot A sugar (raw sugar) is passed on to the dryer (Seebaluck et al., 2008).

Sugar drying: Drying is the last process in raw sugar production (Rein, 2007), and is traditionally done in a rotary drum dryer tilted at an angle of about 5 ° to the horizontal to allow continuous flow of the sugar through the drier (see Figure 3-2 (V)). Hot air is fed into the dryer counter current to the sugar flow and by continuously lifting and dropping the sugar crystals, they are cooled and dried (Seebaluck et al., 2008).

Cooling tower: Warm water from cooling crystallizers and barometric condensers is sent to the cooling tower shown in Figure 3-2 (VI), where a portion of water is evaporated into the air passing through the tower and escapes through the top of the tower. Consequently, the rest of the water is cooled and recycled to the process. Some cold water may be added to the warm water tank to cater for the evaporative losses (U.S. Department of Energy, 2011).

3.2.2. Modelling and simulation

The steady-state operation of a typical sugar mill has been modelled in MATLAB computational tool based on steady-state mass and energy balances of all process units, and was subsequently validated with simulations results of a sugar mill model developed by Sugars International Software Company (Sugars™) for the same input data of 250 tonnes of cane crushed per hour (Starzak & Zizhou, 2015). Later, some industry partners developed Aspen simulation models to complement the MATLAB

model. The Aspen simulations provides the thermophysical properties of stream flows, necessary for exergy analysis and validation of existing mass and energy balances.

The authors of this work verified the mass and energy balances of the Aspen model using the industry-verified MATLAB model and modified it as much as possible to ensure that it adequately reflects the sugar mill operations.

3.2.3. Exergy analysis formulations

Exergy calculations and analysis in this study were carried out in Aspen plus and Excel. It is important to note that, exergy analysis must be preceded by mass and energy balances of the system accomplished in Aspen. Aspen stream results were exported to Excel using Aspen simulation workbook in Aspen V8.6 (Finlayson, 2015). Subsequently, further exergy calculations and analysis were completed in the Excel[®] spread sheet using external data (from sugar industry and literature). Exergy calculations and analysis were carried out based on the following assumptions (Z. Wang, Fan, Zhang, & Dong, 2016):

1. Each of the units was evaluated as a steady state flow process.
2. Changes in potential and kinetic energy and exergy of the system were neglected.
3. Environmental temperature and pressure were assumed constant.

Reference temperature and pressure of 25 °C and 1.01325 bar, respectively, were considered for exergy calculations in Aspen.

3.2.3.1. Exergy calculations

Exergy is the maximum work possible obtained in a reversible process, when a stream of matter or energy at a specified state is brought to equilibrium with the environment with which it interacts. The total exergy transfer through a flow system or across a control volume (flow exergy) is carried out by material and energy (work and heat) streams (Şahin et al., 2010) as expressed in equation (3-1).

$$\dot{E}x_{system} = \dot{E}x_{material} + \dot{E}x_{heat} + \dot{E}x_{work} \quad (3-1)$$

Exergy of all material streams was determined according to equation (3-2).

$$\dot{E}x_{material} = \dot{E}x_{ph} + \dot{E}x_{ch} \quad (3-2)$$

The physical exergy ($\dot{E}x_{ph}$), defined by equation (3-3) was obtained through Aspen prop-set, EXERGYFL (exergy flow rate).

$$\dot{E}x_{ph} = \dot{m}[(h - h_o) - T_o(s - s_o)] \quad (3-3)$$

The chemical exergies were computed by equation (3-4), from the specific chemical exergies of the stream components and their molar fractions (Enrique Querol et al., 2013).

$$\dot{E}x_{ch} = \dot{m} \left[\sum_{component,i} x_i ex_i \right] \quad (3-4)$$

Specific chemical exergy, ex_i of chemical components are available in thermodynamic tables, otherwise, it can be calculated from the standard chemical exergy of their constituent elements and the component's free energy of formation, g_f as in equation (3-5).

$$ex_{component,i} = g_{f,i} + \sum_{element,j} x_j ex_j \quad (3-5)$$

Specific chemical exergies of chemical components found in the sugar mill simulation are presented in Table 3-1.

Table 3-1: Specific chemical exergies of stream components involve the sugar mill process

Component	Chemical formula	Specific chemical exergy [kJ/mol]	References
Water ^a	H ₂ O	0.9	(Aghbashlo et al., 2016; Karina Ojeda et al., 2011; Wall, 2009)
Sucrose	C ₁₂ H ₂₂ O ₁₁	6007.8	(Aghbashlo et al., 2016; Nebra, 2005)
D-Fructose ^b	C ₆ H ₁₂ O ₆	2793.2	(Karina Ojeda et al., 2011)
Cellulose ^b	(C ₆ H ₁₀ O ₅) _n	2749.9	(Karina Ojeda et al., 2011; Sohel & Jack, 2011)
Lime	CaO	110.2	(Karina Ojeda et al., 2011; Wall, 2009)
Crystal ^c	C ₁₂ H ₂₂ O ₁₁	6007.8	-
Air	N ₂ , atmospheric	0.69	(Kotas, 1995; Wall, 2009)

^a It is important to note that, the specific chemical exergy of liquid water is used for both liquid and gaseous water streams (Kotas, 1995), since chemical exergy is the exergy of a substance at the environmental state, at which water is liquid (Frangopoulos, 2009).

^b In the sugar mill Aspen model, D-fructose represents all non-sucrose sugars while cellulose represents fibre.

^c Specific chemical exergy of sucrose crystals is assumed to be that of sucrose since both have the same chemical components

Exergy transfer by work, either electrical or mechanical, is equivalent to the work itself (Cornelissen, 1997; Ghannadzadeh, 2012) as shown in equation (3-6) below.

$$\dot{E}x_{work} = \dot{W} \quad (3-6)$$

Therefore, exergy of electricity input to each process unit was taken as its energy value.

On the other hand, heat is however not entirely convertible to work, according to the second law. Therefore, the exergy of heat flow is its useful work potential (Çengel, 2004; T Tekin & Bayramoğlu, 1998b) as given by equation (3-7).

$$\dot{E}x_{heat} = \dot{Q} \left(1 - \frac{T_o}{T} \right) \quad (3-7)$$

All heat transfers across the mill boundaries are heat losses to the environment. Since the environment is significantly larger than the system, the temperature at which heat is transferred, T is considered to be the environmental temperature, T_o (Gutowski & Sekulic, 2011), resulting in zero exergy value (Madloul, Saidur, Rahim, Islam, & Hossian, 2012) from equation (3-7).

3.2.3.2. Exergy performance indicators

Generally, exergy analysis is carried out to evaluate the thermodynamic performance parameters of a process which include process irreversibility, exergy efficiency, and exergetic improvement potential.

Irreversibility

Process irreversibility is the difference in exergy of input and output streams (equation (3-8)), showing how much exergy is destroyed through each process unit.

$$\dot{I} = \dot{E}x_{destroyed} = \sum \dot{E}x_{in} - \sum \dot{E}x_{out} \quad (3-8)$$

Table 3-2 summarises the exergy balance equations used to evaluate irreversibility of each sugar mill process unit. Equation (3-8) was also applied to individual components of each process unit to determine the sources of irreversibility. The principal phenomena taking place in each component were adjudged the cause of irreversibility evaluated in that component.

Table 3-2: Equations of irreversibility calculation for sugar mill process units

Process unit	Exergy balance equation
I	$\dot{I}_I = (\dot{E}x_1 + \dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + \dot{E}x_6) - (\dot{E}x_{2e} + \dot{E}x_{6c} + \dot{E}x_7 + \dot{E}x_8 + \dot{W}_I^{out^a})$
II	$\dot{I}_{II} = (\dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{11} + \dot{E}x_{12} + \dot{E}x_{13} + \dot{E}x_{25} + \dot{E}x_{26} + \dot{W}_{II}^{in}) - (\dot{E}x_{11c} + \dot{E}x_{12c} + \dot{E}x_{14}^w + \dot{E}x_{15} + \dot{E}x_{16} + \dot{E}x_{25c})$
III	$\dot{I}_{III} = (\dot{E}x_{2e} + \dot{E}x_{16} + \dot{E}x_{17} + \dot{E}x_{18} + \dot{E}x_{19} + \dot{W}_{III}^{in}) - (\dot{E}x_{17c} + \dot{E}x_{20} + \dot{E}x_{21} + \dot{E}x_{22} + \dot{E}x_{22e} + \dot{E}x_{23} + \dot{E}x_{24} + \dot{E}x_{25} + \dot{E}x_{26} + \dot{E}x_{27})$
IV	$\dot{I}_{IV} = (\dot{E}x_{27} + \dot{E}x_{28} + \dot{E}x_{29} + \dot{E}x_{30} + \dot{E}x_{31} + \dot{E}x_{32} + \dot{E}x_{33} + \dot{E}x_{34} + \dot{E}x_{35} + \dot{E}x_{36} + \dot{E}x_{37} + \dot{W}_{IV}^{in}) - (\dot{E}x_{31c} + \dot{E}x_{32c} + \dot{E}x_{33c} + \dot{E}x_{34h} + \dot{E}x_{35h} + \dot{E}x_{36h} + \dot{E}x_{38} + \dot{E}x_{39} + \dot{E}x_{40})$
V	$\dot{I}_V = (\dot{E}x_{40} + \dot{E}x_{41} + \dot{E}x_{42}) - (\dot{E}x_{41c} + \dot{E}x_{43} + \dot{E}x_{44})$
VI	$\dot{I}_{VI} = (\dot{E}x_{20} + \dot{E}x_{34h} + \dot{E}x_{35h} + \dot{E}x_{36h} + \dot{E}x_{38} + \dot{E}x_{45}) - (\dot{E}x_{19} + \dot{E}x_{34} + \dot{E}x_{35} + \dot{E}x_{36} + \dot{E}x_{37} + \dot{E}x_{46}^w + \dot{E}x_{47}^w)$

^a Due to the limitations of the Aspen model, the turbine electricity meant to drive knives, shredders and mills were modelled as outputs. The effect of this on the result is discussed in the results and discussions section.

“ \dot{W} ” and “w” denote exergy transferred by electricity and waste streams, respectively.

Exergy efficiency

Exergy efficiency is a yardstick that indicates the closeness of a system to the ideal (i.e. fully reversible energy conversions), and it provides a more meaningful evaluation of the performance of a process than conventional energy efficiency (Rosen, 2002). Among various exergy efficiency formulations proposed in literature (Gourmelon et al., 2015), the most commonly used is the simple (or universal) exergy efficiency, which is defined as a ratio of the useful output exergy to input exergy of a process expressed by equation (3-9) (Cornelissen, 1997; Gong & Wall, 1997).

$$\psi_{universal} = \frac{\dot{E}x_{out}^{useful\ product}}{\dot{E}x_{in}} \quad (3-9)$$

However, the rational (or functional) exergy efficiency defined in equation (3-10) is a more specific formulation with respect to each system (Cornelissen, 1997; Kotas, 1995).

$$\psi_{functional} = \frac{\dot{E}x_{desired\ output}}{\dot{E}x_{used}} \quad (3-10)$$

The desired exergy output, $Ex_{desired\ output}$ and the necessary exergy input, Ex_{used} are determined in relation to the purpose of each process and both terms account for all exergy transfers into and out of the control volume of that process (Moran, Shapiro, Boettner, & Bailey, 2011). Table 3-3 presents functional exergy efficiency equations developed for each sugar mill process unit in this study.

Table 3-3: Functional exergy efficiency formulations for sugar mill process units

Process unit	Exergy efficiency equation
I	$\psi_I = \frac{(\dot{E}x_7 + \dot{E}x_8 + \dot{E}x_{Elect\ out}) - (\dot{E}x_1 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5)}{(\dot{E}x_6 - \dot{E}x_{6c}) + (\dot{E}x_2 - \dot{E}x_{2e})}$
II	$\psi_{II} = \frac{(\dot{E}x_{15} + \dot{E}x_{16}) - (\dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{26})}{(\dot{E}x_{25} - \dot{E}x_{25c}) + (\dot{E}x_{11} - \dot{E}x_{11c}) + (\dot{E}x_{12} - \dot{E}x_{12c}) + \dot{E}x_{Elect\ in}}$
III	$\psi_{III} = \frac{(\dot{E}x_{20} + \dot{E}x_{21} + \dot{E}x_{22} + \dot{E}x_{22a} + \dot{E}x_{23} + \dot{E}x_{24} + \dot{E}x_{25} + \dot{E}x_{26} + \dot{E}x_{27}) - (\dot{E}x_{16} + \dot{E}x_{19} + \dot{E}x_{34})}{(\dot{E}x_{2e} + \dot{E}x_{17} - \dot{E}x_{17c}) + \dot{E}x_{Elect\ in}}$
IV	$\psi_{IV} = \frac{(\dot{E}x_{38} + \dot{E}x_{39} + \dot{E}x_{40}) - (\dot{E}x_{30} + \dot{E}x_{28} + \dot{E}x_{37} + \dot{E}x_{29} + \dot{E}x_{27}) + (\dot{E}x_{34h} - \dot{E}x_{34}) + (\dot{E}x_{35h} - \dot{E}x_{35}) + (\dot{E}x_{36h} - \dot{E}x_{36})}{(\dot{E}x_{31} - \dot{E}x_{31c}) + (\dot{E}x_{32} - \dot{E}x_{32c}) + (\dot{E}x_{33} - \dot{E}x_{33c}) + \dot{E}x_{Elect\ in}}$
V	$\psi_V = \frac{(\dot{E}x_{43} + \dot{E}x_{44}) - (\dot{E}x_{42} + \dot{E}x_{40})}{(\dot{E}x_{41} - \dot{E}x_{41c})}$

Exergetic improvement potential

Improving exergy efficiency as well as minimizing irreversibility are limited by technological and economic constraints (Cleveland & Ayres, 2004; Kotas, 1995), therefore the exergetic improvement potential is evaluated to indicate the magnitude of, and compare possible improvement potentials (Sogut et al., 2010) of processes. Exergetic improvement potential is a derivative of exergy efficiency and irreversibility as expressed in equation (3-11).

$$\dot{I}P = (1 - \psi)(\dot{E}x_{in} - \dot{E}x_{out}) \quad (3-11)$$

3.3. Results and discussions

Table 3-4 presents detailed stream results from Aspen process simulation and exergy calculations, as well as the stream descriptions. Total exergy is the sum of physical exergy determined in Aspen and chemical exergy calculated from equation (3-4). Mole fractions of multicomponent streams (extracted from Aspen simulation) for chemical exergy calculations are also presented in Table 3-5.

Table 3-4: Description of sugar mill material streams, and their simulation and exergy results

Stream tag	Description	m [kg/s]	T [°C]	P [bar]	Physical Exergy [kW]	Total Exergy [kW]
1	Raw cane	69.44	27	1.01	0.93	401389.89
2	Live steam to mill turbines	5.43	360	28.60	6264.05	6535.43
2e	Total exhaust steam from extraction plant	5.43	130	2.62	3452.13	3723.35
3	1 st effect vapour injected to press water heater	0.80	123	2.13	482.18	522.09
4	Imbibition water to cane diffuser	31.29	55	3.08	207.26	1770.32
5	1 st effect vapour injected to diffuser	1.20	123	2.13	723.17	783.02
6	2 nd effect vapour to diffuser heaters	2.45	115	1.69	1395.10	1517.74
6c	Diffuser heaters condensate	2.45	115	1.69	132.52	255.16
7	Total bagasse	21.49	72	1.01	176.60	204037.02
8	Draft juice	81.24	60	1.01	611.27	199802.56
9	Milk of lime	0.84	20	1.01	0.81	204.72
10	Bagacillo	1.03	72	1.01	8.40	8574.54
11	2 nd effect vapour to secondary heater	3.48	115	1.69	1978.53	2152.45
11c	Condensate from secondary heater	3.48	115	1.69	187.93	361.84
12	1 st effect vapour to tertiary heater	1.76	123	2.13	1063.21	1151.20
12c	Condensate from tertiary heater	1.76	122	2.13	110.90	198.90
13	Wash water to vacuum filter	1.85	55	3.08	12.27	104.80
14	Vapour from flash drum	0.81	100	1.01	398.97	439.48
15	Filter cake	3.79	93	1.01	76.62	17401.22
16	Clarified juice	80.41	100	2.39	2674.97	193779.83
17	Boiler exhaust steam to evaporation plant	22.25	130	2.62	14144.54	15255.89
17c	Boiler feed water from evaporators	27.68	129	2.62	1993.28	3375.86
19	Cooling water to evaporator barometric condenser	500.60	25	3.00	0.00	25031.13

Table 3-4: Description of sugar mill material streams, and their simulation and exergy results (continued 1)

Stream tag	Description	m [kg/s]	T [°C]	P [bar]	Physical Exergy [kW]	Total Exergy [kW]
20	Water from evaporator barometric condenser	513.49	40	1.01	903.15	26578.66
22e	Condensate exported from sugar mill	14.28	55	3.08	94.62	808.21
25	3 rd effect vapour to primary heater	1.71	105	1.19	885.14	970.74
25c	Condensate from primary heater	1.71	105	1.19	72.85	158.45
26	Sludge from syrup	0.04	59	0.16	1.08	417.18
27	Final syrup from syrup filter	16.76	59	0.16	46.13	187557.23
28	Condensate to remelter	0.46	89	0.68	12.80	35.68
29	2 nd effect vapour to remelter	0.02	115	1.69	11.04	12.01
30	Total centrifuge wash water	0.81	89	0.68	22.63	63.08
31	1 st effect vapour to A-pan	5.83	123	2.13	3518.18	3809.33
31c	A-pan condensate	5.83	122	2.13	363.58	654.74
32	1 st effect vapour to B-pan	0.78	123	2.13	470.30	509.22
32c	B-pan condensate	0.78	122	2.13	49.05	87.97
33	2 nd effect vapour to C-pan	0.80	115	1.69	456.74	496.89
33c	C-pan condensate	0.80	115	1.69	43.38	83.53
34	Cold water to A-cooler	2.61	25	3.00	0.00	130.53
34h	Outlet water from A- cooler	2.61	53	1.01	14.78	145.31
35	Cold water to B-cooler	0.98	25	3.00	0.00	48.81
35h	Outlet water from B- cooler	0.98	51	1.01	4.96	53.78
36	Cold water to C-cooler	0.95	25	3.00	0.00	47.70
36h	Outlet water from C- cooler	0.95	44	1.01	2.73	50.43
37	Cooling water to pan barometric condenser	256.39	25	3.00	0.00	12820.25
38	Water from pan barometric condenser	262.80	40	3.00	453.92	13595.13
39	C-molasses from C-centrifuge	2.68	52	1.01	-1.81	31857.62
40	A-sugar from A-centrifuge	8.96	56	1.01	0.27	155395.29
41	Exhaust steam to drying plant	0.29	130	2.62	181.92	196.21
41c	Condensate from drying plant	0.29	129	2.62	20.59	34.89

Table 3-4: Description of sugar mill material streams, and their simulation and exergy data (continued 2)

Stream tag	Description	m [kg/s]	T [°C]	P [bar]	Physical Exergy [kW]	Total Exergy [kW]
42	Drying air	46.48	26	1.01	-0.91	1125.91
43	Air from drying plant	46.56	40	1.01	17.21	1158.57
44	Sugar from dryer	8.88	35	1.01	-0.23	155380.25
45	Cooling tower make-up water	0.07	25	1.01	0.00	3.62
46	Cooling tower vapours	16.83	27	0.04	315.11	1155.77
47	Waste water	2.55	25	3.00	0.00	127.49
	\dot{W}_I^{out}					2456.25
	\dot{W}_{II}^{in}					17.80
	\dot{W}_{III}^{in}					19.12
	\dot{W}_{IV}^{in}					0.10

Stream 18 = {6c, 11c, 12c, 25c, 31c, 32c and 33c}; Stream 21 = {28 and 30}; Stream 22 = {4 and 13}; Stream 23 = {3, 5, 12, 31 and 32}; Stream 24 {6, 11, 29 and 33}; \dot{W} = electricity rate

Table 3-5: Mole fractions of components in multicomponent streams

Stream tag	Water	Sucrose	D-Fructose	Cellulose	Lime	Crystal	Air
1	0.963	0.011	0.003	0.024	-	-	-
7	0.905	0.001	0.001	0.093	-	-	-
8	0.990	0.007	0.002	0.001	-	-	-
9	0.966	-	-	-	0.034	-	-
10	0.905	0.001	0.001	0.093	-	-	-
15	0.951	-	0.006	0.033	0.010	-	-
16	0.991	0.007	0.002	-	-	-	-
26	0.919	0.053	0.027	-	-	-	-
27	0.902	0.078	0.020	-	-	-	-
39	0.781	0.055	0.164	-	-	-	-
40	0.152	0.015	0.016	-	-	0.816	-
42	0.025	-	-	-	-	-	0.975
43	0.028	-	-	-	-	-	0.972
44	0.012	0.002	0.019	-	-	0.967	-

3.3.1. Mass, energy and exergy balances

Mass and energy balances are the basis on which the exergy analysis was conducted. Table 3-6 presents the total mass, energy and exergy balances around the control volume of the sugar plant. Basically, the plant is in mass and energy balance, according to the law of conservation of mass and the first law of thermodynamics. A relative difference of less than 0.002 % in both cases shows a

good balance. Exergy balance was split into three categories for the purposes of this study. A balanced chemical exergy (with a relative difference of less than 0.001 %) indicate that chemical exergy remains unchanged through the sugar mill processes (Gourmelon et al., 2015). The difference between input and output physical exergy and the total exergy is the exergy destroyed in the entire sugar plant. Exergy destruction termed irreversibility is discussed in details in the next section.

Table 3-6: Results of mass, energy and exergy balances of the sugar plant

	\dot{m} (kg/s)	H [kW]	Ex_{ph} [kW]	Ex_{ch} [kW]	Ex_{tot} [kW]
Total material input	145.828	-1268000.2	20596.7	388275.5	408872.2
Power in		37.0	37.0		37.0
Total material output	125.639	-1010289.8	2376.9	387266.0	389642.9
Waste material output	20.188	-277021.9	714.1	1008.7	1722.8
Power out		2456.2	2456.2		2456.2
Heat loss		16839.1			
Difference	0.001	20.8	15086.5	0.8	15087.3

3.3.2. Irreversibility study

3.3.2.1. Exergy accounting of the sugar mill

Accounting for exergy resources distribution in a process is important for assessing its performance from a thermodynamic point of view. It has been established from exergy balances that the total chemical exergy entering the sugar mill is equal to that exiting it, considering negligible or no sucrose inversion in the process. This section presents the account of all exergy resources (without chemical exergy) of the sugar mill and its individual units. Figure 3-3 shows that only 23.4 % of exergy input to the sugar mill (21 MW) exits the process through product streams, 3.5 % was lost through waste streams while 73.1 % was destroyed in the process, due to irreversibilities. The boiler of the cogeneration unit is reported as the major sugar mill exergy destruction unit (A. V. Ensinas et al., 2009; Taner & Sivrioglu, 2015), the analysis of which is excluded from this paper. Therefore, in the absence of the boiler the evaporation unit generates the most irreversibility, destroying 33.8 % of the total exergy resource.

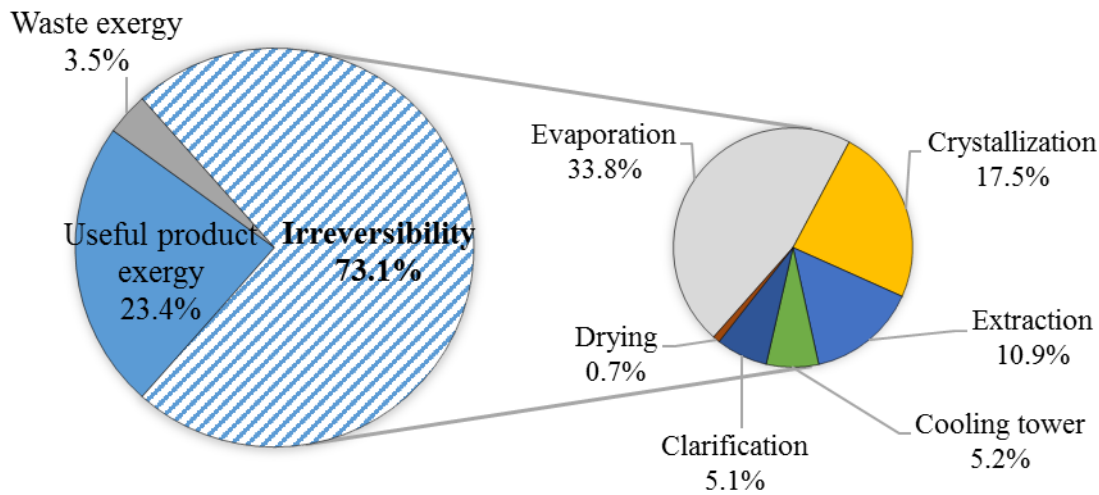


Figure 3-3: Distribution of sugar mill exergy resources and contributions of process unit to sugar mill total irreversibility

Figure 3-4 presents a visual exergy accounting of each sugar mill process unit in the form of Grassmann diagrams. Irreversibilities of the extraction (8064.94 kW), evaporation (25108.84 kW), crystallization (12985.34 kW) and drying (135.27 kW) units represent 24.70 %, 32.82 %, 79.46 % and 77.16 %, respectively of their individual exergy resources, while the remaining exited through the products. In the clarification unit, 23.07 % (3802.53 kW) of exergy resource was destroyed, 8.71 % wasted in the flash vapour and the remaining flowed through the products. On the other hand, 77.16 % (3831.96 kW) of cooling tower exergy resource was destroyed in the process while the remaining was wasted through the vapours. The causes of these irreversibilities, as well as possible engineering solutions, are discussed in the following sections. A reduction in the process irreversibilities in sugar mills, as well as all energy-intensive industries is required to minimize energy resources utilization, and hence lessen global energy and environmental challenges. Figure 3-5 shows the Grassmann diagram accounting for exergy flows through the entire sugar mill.

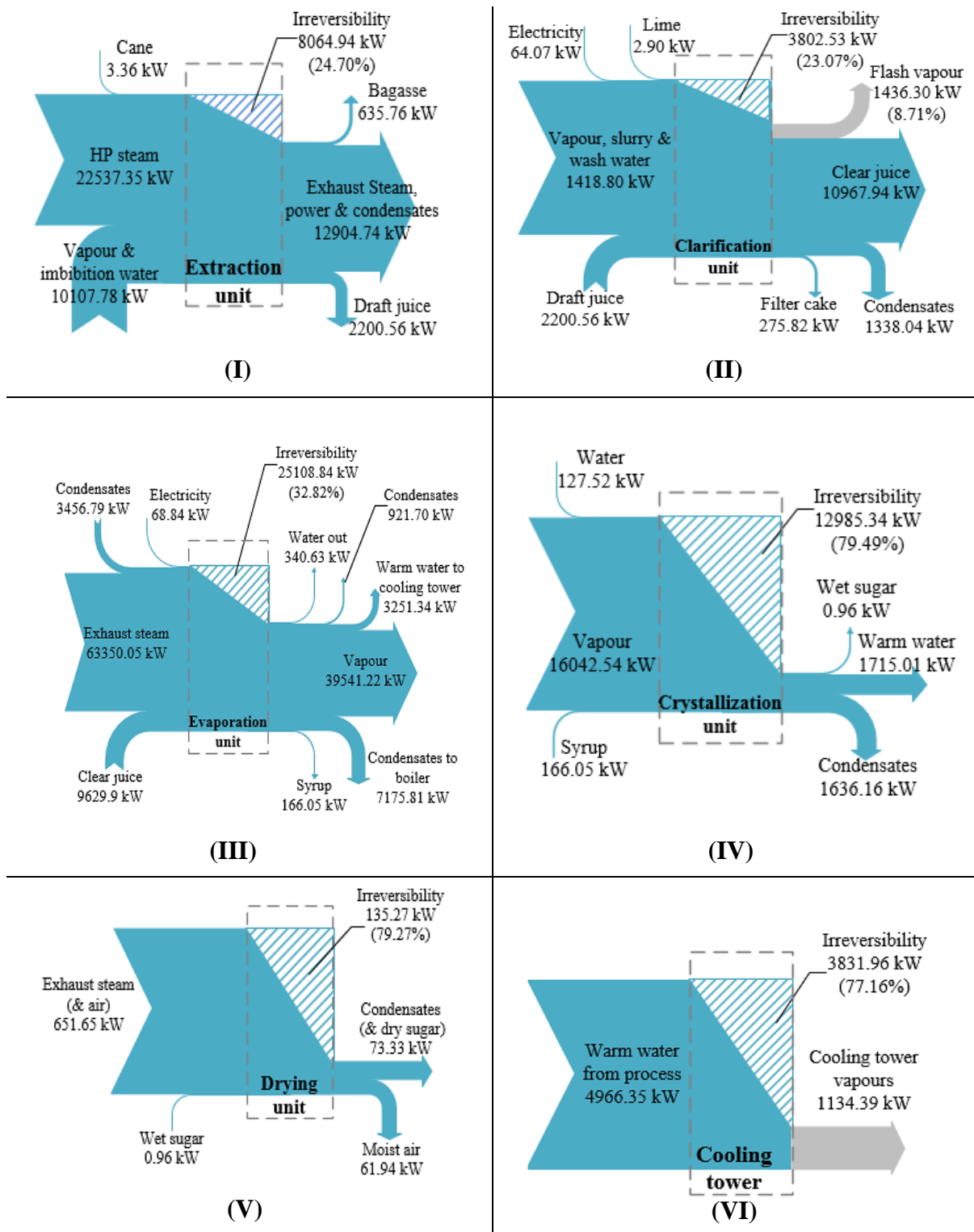


Figure 3-4: Grassmann diagrams of sugar mill process units

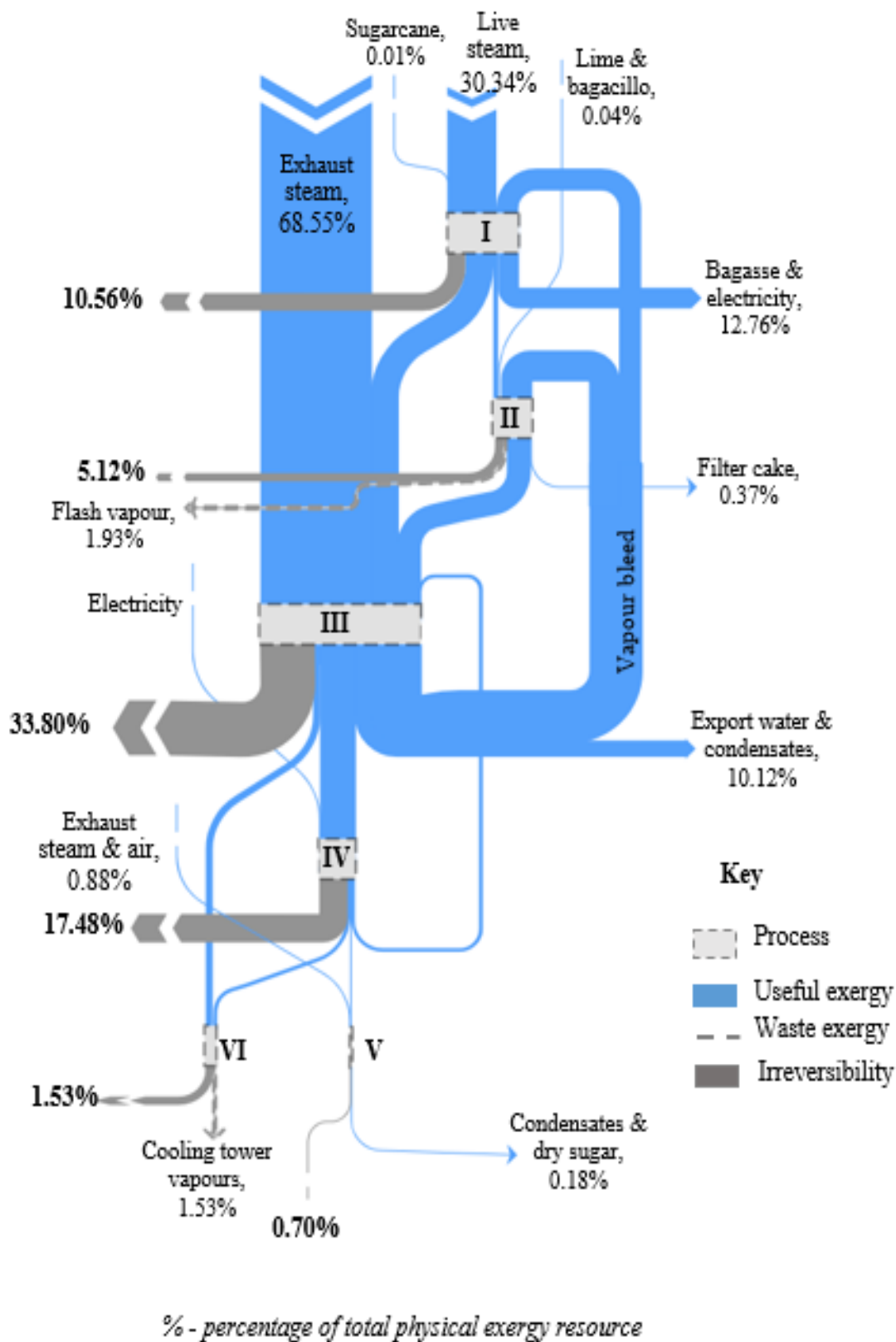


Figure 3-5: Overall Grassmann diagram of sugar mill; (I) extraction, (II) clarification, (III) evaporation, (IV) crystallization, (V) drying, and (VI) cooling tower

3.3.2.2. Sources of Irreversibility

Real processes are characterized by phenomena that generate irreversibilities. These include work dissipation to heat by solid or fluid friction, spontaneous chemical reaction, heat transfer over finite temperature differences, unrestricted expansion and equalization of temperature in mixing (Çengel, 2004; Kotas, 1995). Sugar mill processes considered here involve one or more of these phenomena taking place in the various components making up each process unit, except spontaneous chemical reaction since no chemical processes were analysed.

Figure 3-6 (a) presents the principal components responsible for extraction unit irreversibility. The diffuser heat exchanger generated the highest irreversibility due to heat transfer over 74.9 °C temperature difference. In addition, the diffuser and the press water heater caused irreversibility resulting from temperature equalizing of mixing streams. Irreversibility of the diffuser is more than four times that of the press water heater, even though the temperature gradient in the former (71 °C) was almost the same as the latter (68 °C). This may be due to the higher juice draft and mixing friction as well as heat loss in the diffuser (Taner Tekin & Bayramoğlu, 2001). Another cause of irreversibility is an unrestrained expansion of steam in the low-efficiency prime mover turbines. Replacing turbines with electric drivers for shredders, knives and dewatering mills could minimize this irreversibility (A. V Ensinas & Nebra, 2009).

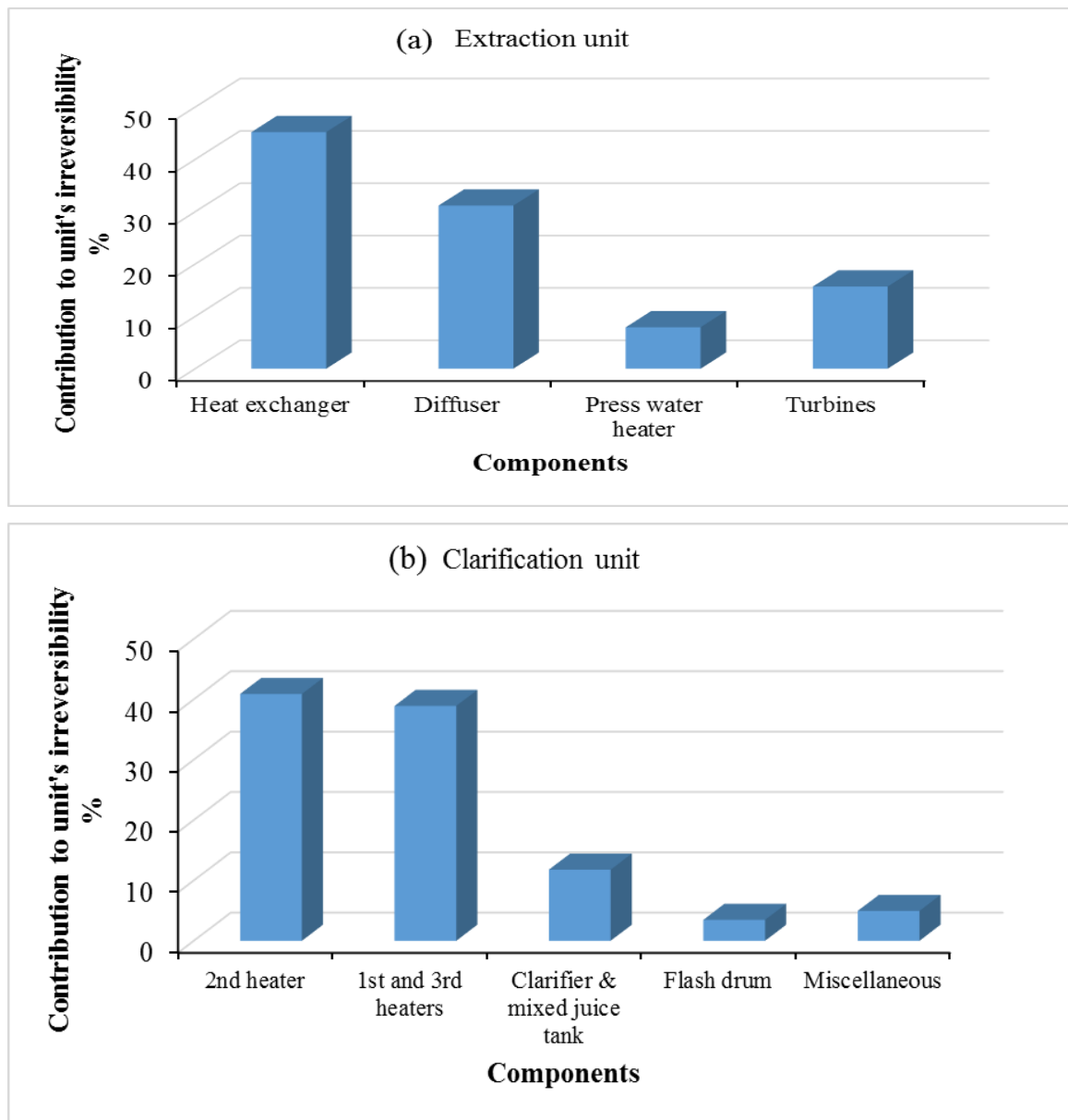


Figure 3-6: Sources of irreversibility in (a) extraction and (b) clarification units

Reducing the driving force (temperature difference) of indirect heat transfer components such as heat exchangers to the lowest value possible would minimize their irreversibility, since such irreversibility is proportional to the temperature difference (Ghannadzadeh & Sadeqzadeh, 2016). However, this measure would require larger heat exchangers which are not favourable economically and as such, this irreversibility could be considered unavoidable. Nevertheless, a trade-off exists between improving the system thermodynamics and economics in order to improve such processes (Cleveland & Ayres, 2004).

Mixing irreversibility may be minimized by isothermal mixing (Ghannadzadeh & Sadeqzadeh, 2016), that is, ensuring stream temperatures are brought close to each other before mixing. Once again, there is a trade-off between economics and irreversibility reduction in this case.

Figure 3-6 (b) shows the major components responsible for clarification unit irreversibility. Juice heaters altogether generate almost 80 % percent of the total unit irreversibility, due to indirect heat transfer over finite temperature difference between juice and heating steam. However, the irreversibility of the secondary heater is higher than in both primary and tertiary heaters because by the nature of the clarification process, the secondary heater has a higher heat transfer rate. Contrariwise, where carbonation liming is used (as in the beet sugar industry) instead of the simpler liming process used in the sugar mill analysed in this study, chemical reactions could be the principal source of irreversibility in the clarification unit (T Tekin & Bayramoğlu, 1998a). Other irreversibility causes include equalizing the temperature of streams in the mixed juice tank and the clarifier and stream pressure reduction in the flash drum (Kotas, 1995). Other minor components (miscellaneous) together contribute 5 % to the unit's irreversibility.

Figure 3-7 (a) shows the contributions of evaporation unit components to the unit's irreversibility. More than 50 % of evaporation unit irreversibility is generated in the evaporators. Various degrees of temperatures differences between heating steam (or vapour), and the concentrating juice cause the evaporators' irreversibility (T Tekin & Bayramoğlu, 1998a). Another principal cause of evaporators' irreversibility is the reduction in juice pressure in successive evaporator effects. The 5th effect evaporator alone produces 27 % of the unit's irreversibility principally due to the higher heat transfer temperature difference (32 °C) than the first four effects (ranging from 7 – 15 °C) and the vacuum in this effect (Kotas, 1995; Pellegrini & de Oliveira, 2006).

The barometric condenser is the highest source of irreversibility (29 %). The principal cause may be mixing of vapour from the last effect (55 °C, 0.16 bar) and cold water from cooling tower (25 °C, 3 bar) at different temperature and pressure conditions (Ghannadzadeh & Sadeqzadeh, 2016). Yet, the barometric condenser is important to control the evaporator station pressure profile. Various other components labelled miscellaneous contribute less than 10 % to the unit's irreversibility.

The evaporation unit irreversibility may be reduced by increasing the number of evaporator effects, resulting in lower temperature and pressure gradients in each effect (Pellegrini & de Oliveira, 2006; Sogut et al., 2010). Alternatively, introducing falling film evaporators to replace the conventional Robert type evaporator currently used, may reduce irreversibility. Falling film evaporators operate at low residence time and high heat transfer rates, and require lower temperature difference for the same amount of water evaporated using the Robert type (Gurumurthy, 2011).

For the barometric condenser, cooling water with a temperature close to the last effect vapour is required to reduce irreversibility. However, this practice is undesirable as it reduces vacuum in the system, and increases boiling temperatures, which are unfavourable in the sugar industry because of

sucrose inversion (Chouhan, 2014). Moreover, the volume of water required would also increase with increased pumping costs. Alternatively, mechanical vapour recompression could be exploited to increase the pressure of vapours off the 5th effect evaporator and vacuum pans in order to reuse them, provided this does not result in surplus of exhaust steam. Thus, eliminating the highly irreversible barometric condensers and the cooling tower (Sciubba & Wall, 2007; T Tekin & Bayramoğlu, 1998a). Reid & Rein (1983) suggested mechanical vapour compression for sugar mill evaporators and pans for reduction of steam consumption. Further studies may investigate exergy savings using vapour compression and their economic feasibility in comparison to the barometric condensers.

Figure 3-7 (b) presents the principal components responsible for crystallization unit irreversibility. Sugar boiling in the A-pan generates 55% of the unit's irreversibility due to the temperature difference between heating vapour and syrup. Barometric condenser is the next major irreversibility source (25 %), due to the reasons discussed previously. B-pan and C-pan together contribute only 19 % to the units' irreversibility as water evaporated from each of them is six times less than the A-pan vapour, again indicating the effect of process size on irreversibility. Therefore, the specific irreversibility (as indicated by the percentage irreversibility value in Figure 3-4) might give a better indication of thermodynamic performance than the absolute irreversibility value, when comparing similar processes in different plants. Other components such as cooling crystallizers, centrifuges, mingler and remelter, altogether contribute only 1 % to the unit's irreversibility.

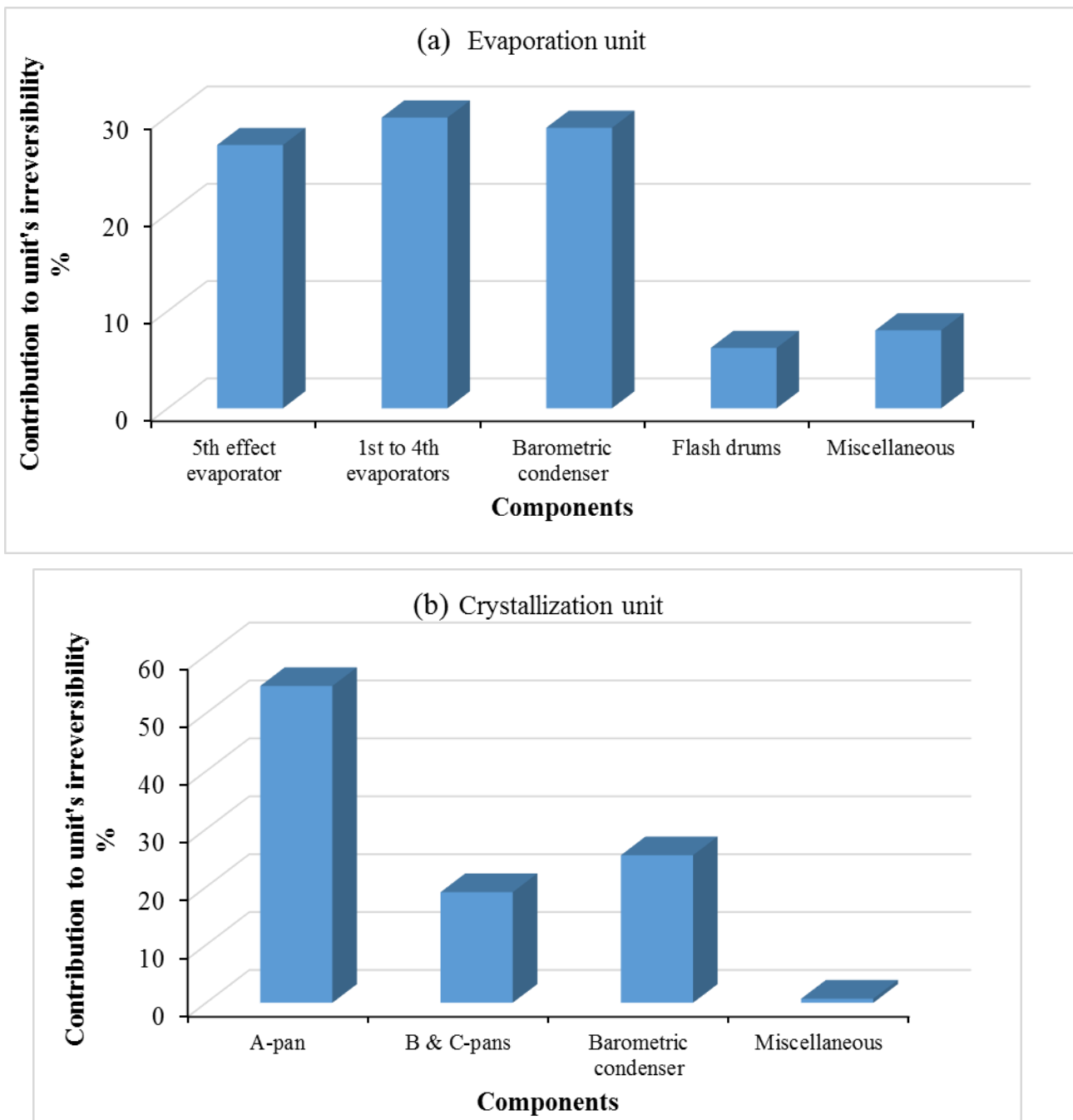


Figure 3-7: Sources of irreversibility in (a) evaporation and (b) crystallization units

As shown in Figure 3-8 (a), more than 75 % of drying unit irreversibility occurs in the air heater due to indirect heat transfer between drying air and steam over a temperature difference of 73 °C. Other locations of irreversibility include the sugar dryer and air mixer. The dryer irreversibility could be due to high air temperature (80 °C). Finally, equalizing temperatures of drying air (60 °C) and cooling air (34 °C) in the air mixer also generates irreversibility. Therefore, isothermal mixing of the exit drying and cooling air would be necessary to minimize such irreversibility (Ghannadzadeh & Sadeqzadeh, 2016). In order to reduce temperature difference in air heat and hence reduce its irreversibility, waste streams such as flash drum vapour could be used to pre-heat the drying air.

Figure 3-8 (b) shows that 98 % of the cooling tower irreversibility is generated in the cooling column due to direct contact heat and mass transfer from process warm water to air passing through the tower. Equalizing temperatures of the various process water in the mixing tank contributes the remaining 2 % to the cooling tower irreversibility. The only way to avoid cooling tower irreversibility is, to cool process water by other means such as using absorption heat pump (Chouhan, 2014) to make use of the process hot water exergy which is otherwise, wasted in the cooling tower.

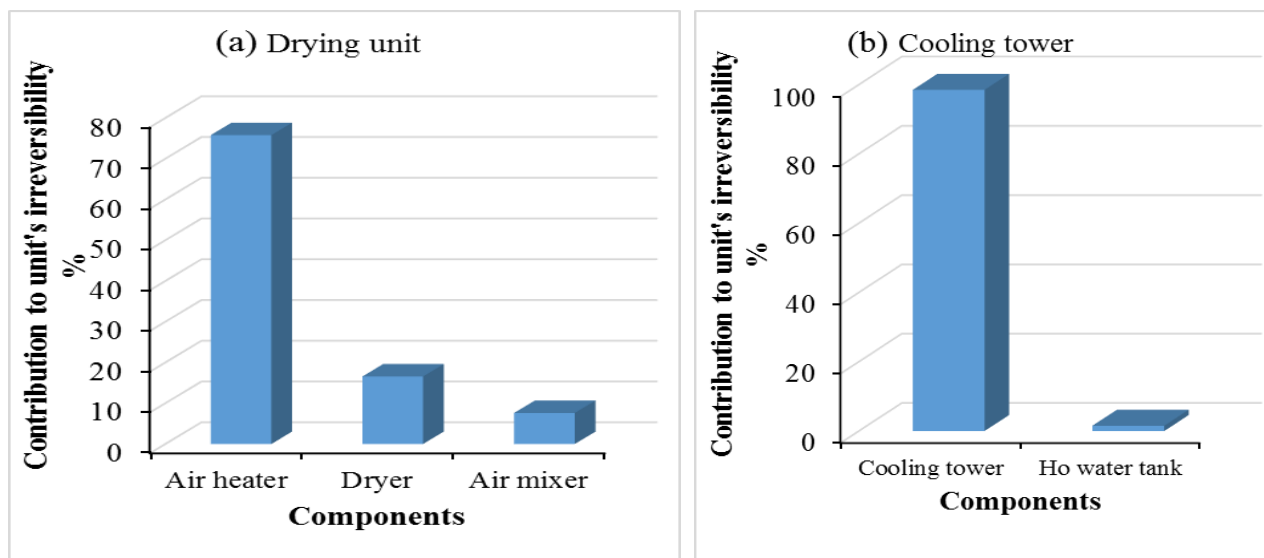


Figure 3-8: Sources of irreversibility in (a) drying unit and (b) cooling tower

3.3.3. Exergy efficiency

In previous studies (Bayrak et al., 2003; A. V. Ensinas et al., 2009; T Tekin & Bayramoğlu, 1998a), the universal exergy efficiency (formulated in equation (3-9)) results for physical processes were very high (over 90 %) giving exaggerated performances (T Tekin & Bayramoğlu, 1998a) and might wrongly represent the system. Therefore in this study, only the functional exergy efficiencies were evaluated. The functional exergy efficiency gives a “rational” indication of the process because it is specific to each process and defines the purpose of the process (Gourmelon et al., 2015; Kotas, 1995) as formulated in Table 3-3, and not just a general input-output formulation like the universal exergy efficiency. Even though the exergy balances in Table 3-6 favours the assumption of neglecting chemical material exergy for physical processes, the functional exergy efficiency was evaluated in two ways using (i) physical material exergy only and (ii) the total material exergy, in order to investigate the effect of the assumption on exergy efficiency indication of such processes.

Figure 3-9 shows that the results for each unit using both analysis methods (physical and total exergy) are the same, except the clarification unit with the highest functional exergy efficiency, as well as the overall plant. Unlike in the irreversibility analysis, where output streams include waste streams for

which reason the difference in the chemical exergy of input and output streams is zero and can be ignored, the output stream in exergy efficiency evaluation does not include waste streams (in the clarification and cooling tower) as products. The result is that there is a small difference in the chemical exergy of resources and product streams, which is ignored in the physical exergy only analysis. Hence, the exergy efficiency by total exergy analysis is the better representation of the system performance.

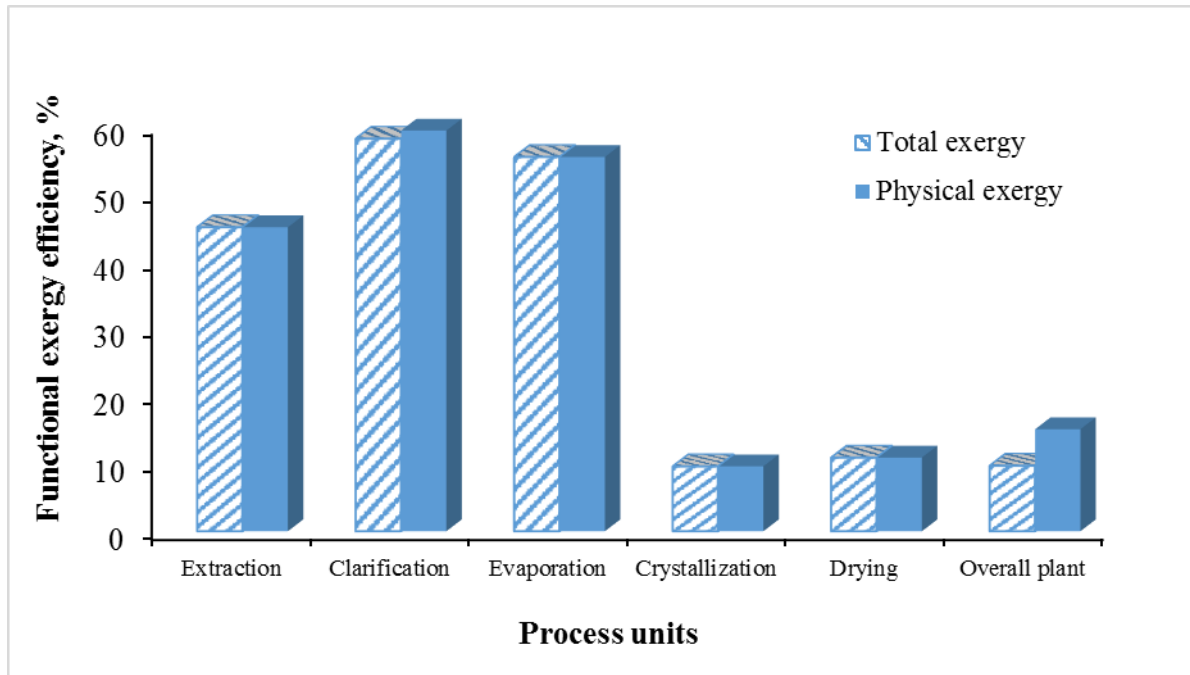


Figure 3-9: Comparison of functional exergy efficiencies evaluated by physical and total exergy analysis

The functional exergy efficiency of the overall plant was 9.7 %. The efficiencies of clarification (58.1 %) and evaporation (55.4 %) units reflect how effectively their input exergies were used in raising the juice temperature and concentrating the juice, respectively, thus producing high exergy products. Likewise, the extraction unit (45 %) has relatively good exergy utilization. However, its efficiency would be a little lower if electricity produced by mill turbines was coupled with the prime movers in the simulation. The reason is that part of the electricity exergy would be expended on frictional irreversibilities in the prime movers.

Crystallization unit has the lowest exergy efficiency of 9.6 % followed by the drying unit with 10.9 %. These unusually low functional exergy efficiencies may be due to relatively high exergy demands of the processes and low exergy products. Complex crystallization processes such as concentration, crystal nucleation, re-melting, centrifuging (Rein, 2007), make the most use of the exergy resource, while the main exergy resource of the drying unit is expended more on irreversibilities of air heating, rather than the drying product - sugar.

Further analysis of crystallization unit revealed that B and C-pans have lower functional exergy efficiencies than A-pans, even though their irreversibilities are far lower than that of the A-pans (see Figure 3-7 (b)). This may be due to higher boiling point elevation, requiring an additional exergy expended to evaporate water from the B and C-pans. Thus, a possible solution to the unusually low functional efficiency of the crystallization unit may be to produce sugar in one stage sugar boiling and crystallization, while sucrose remaining in the A-molasses could be used for a bio-refinery product such as ethanol (Diederichs, Mandegari, Farzad, & Görgens, 2016).

The cooling tower has no functional exergy efficiency because it has no useful output to be expressed in terms of exergy (Kotas, 1995). Its purpose is to release heat to the environment which has zero exergy. Therefore the exergy resource is expended on evaporating vapours, which is of no exergetic use, leading to cooling of the process water.

Comparing irreversibility and functional exergy efficiency results, it is clear that prioritizing processes for improvement exclusively based on irreversibility and exergy efficiency becomes a difficult task. Ideally, the part of a system with the highest irreversibility should be considered first for improvement. This is, however, not always the case since the systems are interconnected and trade-offs may exist (Cleveland & Ayres, 2004). Therefore, the entire process improvement task requires a more tactical approach (Gong & Wall, 1997), which would result in significant cost reduction or profit rise, as well as greater ease of operation.

3.3.4. Exergetic improvement potential

Evaluating exergetic improvement potential provides a distinct factor for making improvement priority decisions, as it combines the effects of both irreversibility and functional exergy efficiency in the improvement process. Figure 3-10 presents the exergetic improvement potential of each process unit relative to its irreversibility. The crystallization unit has the highest potential for improvement of 3.3 MW followed by the evaporating unit with 3.1 MW, representing about 22 % and 21 % of the sugar mill's total irreversibility, respectively (compare Figure 3-10 with Table 3-6). The drying unit has the least potential for improvement, less than 1 % of mill's total irreversibility, indicating relatively no need to improve it. Usually, high exergetic improvement potential is associated with high irreversibility units (Dowlati et al., 2017; Sogut et al., 2010) as can be seen in this study. Further studies may be directed towards investigating the rate at which such improvements can be obtained to enhance the decision-making process for improvement.

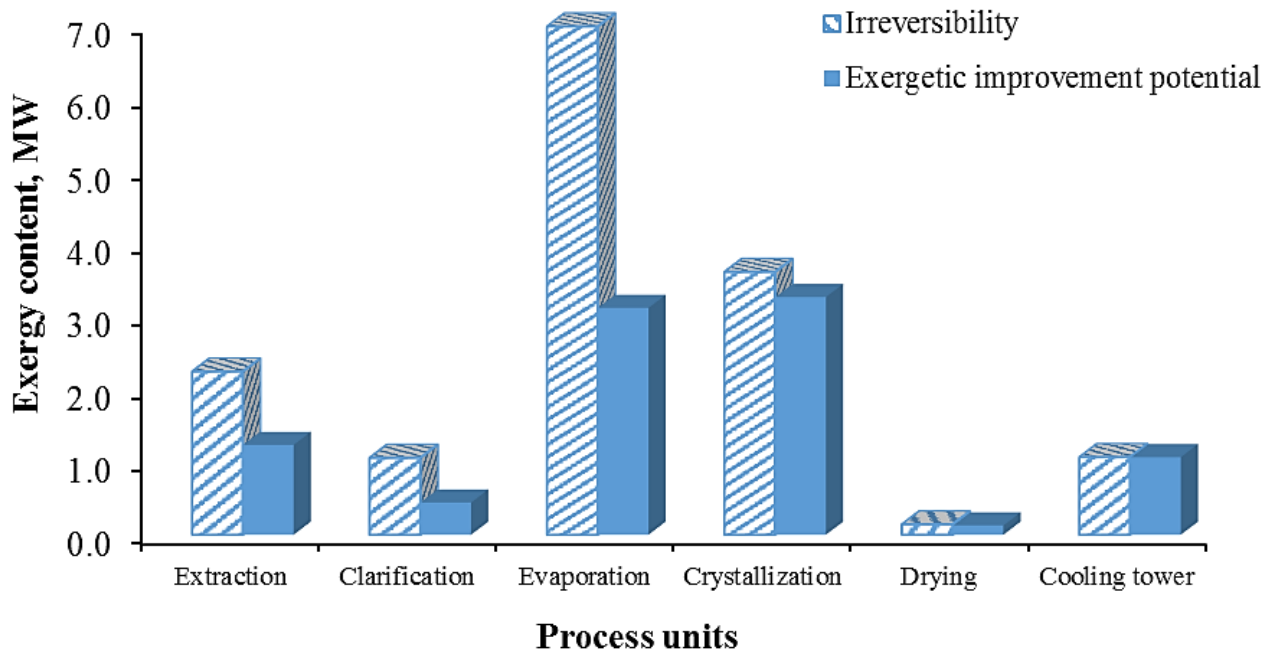


Figure 3-10: Exergetic improvement potentials of sugar mill process units compared to their irreversibilities

3.4. Conclusion

The study presented a comprehensive exergy analysis of raw cane sugar production processes on the basis of Aspen simulation of a typical sugar mill, and functional exergy efficiency formulations for each process unit. The overall plant irreversibility and functional exergy efficiency were 15 MW and 9.7 %, respectively. The evaporation unit had the highest irreversibility, representing 46 % of the global plant irreversibility due to indirect heat transfers over finite temperature differences and temperature equalization through mixing. Irreversibility was also found to depend on the process “size”, therefore the specific irreversibility may be a better performance indicator when comparing similar processes of different plants. The crystallization unit of the sugar mill analysed had lowest functional exergy efficiency of 9.6 % due to process complexity and was ranked first for improvement based on the exergetic improvement potential. As per literature, the study showed that chemical exergy may be neglected in exergy analysis of physical processes. However, where there are a substantial number of waste streams, the exergy efficiency obtained by physical exergy only analysis could overestimate the process performance. Based on the results of the sugar mill exergy analysis, the following engineering solutions can be suggested for improving exergy performance of the mill; vapour recompression to replace barometric condensers and cooling towers, one-stage crystallization with an integrated biorefinery, heat transfer over small temperature differences and isothermal mixing of streams. However, it is necessary to assess the economic implications of these exergy solutions, through exergoeconomics studies for instance, prior to their implementation.

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References

- Aghbashlo, M., Tabatabaei, M., & Karimi, K. (2016). Exergy-based sustainability assessment of ethanol production via *Mucor indicus* from fructose, glucose, sucrose, and molasses. *Energy*, 98, 240–252. <http://doi.org/10.1016/j.energy.2016.01.029>
- Ali Mandegari, M., Farzad, S., & Görgens, J. F. (2017). Economic and environmental assessment of cellulosic ethanol production scenarios annexed to a typical sugar mill. *Bioresource Technology*, 224, 314–326. <http://doi.org/10.1016/j.biortech.2016.10.074>
- Assari, M., Basirat, T., Najafpour, E., Ahmadi, A., & Jafari, I. (2014). Exergy modeling and performance evaluation of pulp and paper production process of bagasse, a case study. *Thermal Science*, 18(4), 1399–1412. <http://doi.org/10.2298/TSCI110715141A>
- Atmaca, A., & Yumrutaş, R. (2014). Thermodynamic and exergoeconomic analysis of a cement plant: Part II – Application. *Energy Conversion and Management*, 79, 799–808. <http://doi.org/10.1016/j.enconman.2013.11.054>
- Bayrak, M., Midilli, A., & Nurveren, K. (2003). Energy and exergy analyses of sugar production stages. *International Journal of Energy Research*, 27(11), 989–1001. <http://doi.org/10.1002/er.916>
- Çengel, Y. A. (2004). *Thermodynamics: An Engineering Approach*. McGraw-Hill.
- Chouhan, P. (2014). Performance enhancement of sugar mill by alternate cooling system for condenser. *IOSR Journal of Mechanical and Civil Engineering (IOSR-JMCE)*, 11(5), 18–25. Retrieved from <http://www.iosrjournals.org/iosr-jmce/papers/vol11-issue5/Version-1/C011511825.pdf>
- Cleveland, C. J., & Ayres, R. U. (2004). Encyclopedia of energy. In *Energy*.
- Cornelissen, R. L. (1997). *Thermodynamics and Sustainable Development: The use of Exergy*

Analysis and the Reduction of Irreversibility. University of Twente, The Netherlands. Retrieved from

<http://scholar.google.com/scholar?hl=en&btnG=Search&q=intitle:Thermodynamics+and+sustainable+development#5>

- Costa, M. M., Schaeffer, R., & Worrell, E. (2001). Exergy accounting of energy and materials flows in steel production systems. *Energy*, 26(4), 363–384. [http://doi.org/10.1016/S0360-5442\(01\)00004-4](http://doi.org/10.1016/S0360-5442(01)00004-4)
- de Oliveira Jr, S. (2013). *Exergy: Production, Cost and Renewability*. London: Springer. <http://doi.org/10.1007/978-1-4471-4165-5>
- Diederichs, G. W., Mandegari, M. A., Farzad, S., & Görgens, J. F. (2016). Techno-economic comparison of biojet fuel production from lignocellulose, vegetable oil and sugar cane juice. *Bioresource Technology*, 216, 331–339. <http://doi.org/10.1016/j.biortech.2016.05.090>
- Dowlati, M., Aghbashlo, M., & Mojarab, M. (2017). Exergetic performance analysis of an ice-cream manufacturing plant: A comprehensive survey. *Energy*, 123, 445–459. <http://doi.org/10.1016/j.energy.2017.02.007>
- Ensinas, A. V., Modesto, M., Nebra, S. A., & Serra, L. (2009). Reduction of irreversibility generation in sugar and ethanol production from sugarcane. *Energy*, 34(5), 680–688. <http://doi.org/10.1016/j.energy.2008.06.001>
- Ensinas, A. V., & Nebra, S. A. (2009). Exergy analysis as a tool for sugar and ethanol process improvement. In G. Pelissier & A. Calvet (Eds.), *Handbook of Exergy, Hydrogen and Hydropower Research* (pp. 125–160).
- Ensinas, A. V., Nebra, S. a, Lozano, M. a, & Serra, L. (2006). Analysis of cogeneration systems in sugar cane factories - Alternatives of steam and combined cycle power plants. In *Proceedings of ECOS* (pp. 1177–1184). Aghia Pelagia, Crete, Greece.
- Farzad, S., Mandegari, M. A., Guo, M., Haigh, K. F., Shah, N., & Görgens, J. F. (2017). Multi-product biorefineries from lignocelluloses: a pathway to revitalisation of the sugar industry? *Biotechnology for Biofuels*, 10. <http://doi.org/10.1186/s13068-017-0761-9>
- Finlayson, B. A. (2015). AspenTech. AspenTech.
- Frangopoulos, C. A. (2009). *Exergy, energy system and optimization - Volume 1: Exergy and thermodynamic analysis*. (C. A. Frangopoulos, Ed.). Oxford, United Kingdom: Eolss Publishers Co. Ltd.

- Garg, A., Sharma, M. P., & Sharma, V. (2016). Exergy and energy analyses of a sugarcane juice production and clarification unit. *International Journal of Exergy*, 19(1), 78. <http://doi.org/10.1504/IJEX.2016.074268>
- Ghannadzadeh, A. (2012). *Exergetic Balances and Analysis in a Process Simulator: A Way to Enhance Process Energy Integration*. Université de Toulouse.
- Ghannadzadeh, A., & Sadeqzadeh, M. (2016). Exergy analysis as a scoping tool for cleaner production of chemicals : a case study of an ethylene production process. *Journal of Cleaner Production*. <http://doi.org/10.1016/j.jclepro.2016.04.018>
- Gong, M., & Wall, G. (1997). On exergetics, economics and optimization of technical processes to meet environmental conditions. *Thermodynamic Analysis and Improvement of Energy Systems*, 10–13(June), 453–461.
- Gourmelon, S., They-Hetreux, R., Floquet, P., Baudouin, O., Baudet, P., & Campagnolo, L. (2015). Exergy analysis in ProSimPlus ® simulation software: A focus on exergy efficiency evaluation. *Computers & Chemical Engineering*, 79, 91–112. <http://doi.org/10.1016/j.compchemeng.2015.02.014>
- Gurumurthy, B. S. (2011). Advancement in sugar processing technologies in India emerging innovations. *Sugar Tech*, 13(4), 378–393. <http://doi.org/10.1007/s12355-011-0106-0>
- Gutowski, T. G., & Sekulic, D. P. (2011). Thermodynamic analysis of resources used in manufacturing processes. In *Thermodynamics and the Destruction of Resources* (pp. 1–43). Cambridge University Press.
- Héctor, V. I., Carolina Mesa, M., & Sergio, G. A. A. (2013). Energy and exergy analysis of the combined production process of sugar and ethanol from sugarcane (a Colombia case study). In *Proceedings of ECOS* (pp. 1–14).
- Hevert, H. W., & Hevert, S. C. (1980). Second law analysis: An alternative indicator of system efficiency. *Energy*, 5(8–9), 865–873. [http://doi.org/10.1016/0360-5442\(80\)90102-4](http://doi.org/10.1016/0360-5442(80)90102-4)
- Higa, M., Freitas, a. J., Bannwart, a. C., & Zemp, R. J. (2009). Thermal integration of multiple effect evaporator in sugar plant. *Applied Thermal Engineering*, 29(2–3), 515–522. <http://doi.org/10.1016/j.applthermaleng.2008.03.009>
- Hosseini, S. S., Aghbashlo, M., Tabatabaei, M., Younesi, H., & Najafpour, G. (2015). Exergy analysis of biohydrogen production from various carbon sources via anaerobic photosynthetic bacteria (*Rhodospirillum rubrum*). *Energy*, 93, 730–739. <http://doi.org/10.1016/j.energy.2015.09.060>

- Hugot, E. (1972). *Handbook of Cane Sugar Engineering* (2nd ed.). Elsevier Science Ltd.
- International Sugar Organization. (2016). About sugar. Retrieved August 28, 2017, from <http://www.isosugar.org>
- Kamate, S. C., & Gangavati, P. B. (2009). Exergy analysis of cogeneration power plants in sugar industries. *Applied Thermal Engineering*, 29(5–6), 1187–1194. <http://doi.org/10.1016/j.applthermaleng.2008.06.016>
- Kotas, T. J. (1995). *The Exergy Method of Thermal Plant Analysis* (Reprint Ed). Florida, USA: Krieger Publishing Company, Malabar. <http://doi.org/10.1016/B978-0-408-01350-5.50003-9>
- Luis, P., & Van der Bruggen, B. (2014). Exergy analysis of energy-intensive production processes: Advancing towards a sustainable chemical industry. *Journal of Chemical Technology and Biotechnology*, 89(9), 1288–1303. <http://doi.org/10.1002/jctb.4422>
- Madloul, N. A., Saidur, R., Rahim, N. A., Islam, M. R., & Hossian, M. S. (2012). An exergy analysis for cement industries: An overview. *Renewable and Sustainable Energy Reviews*, 16(1), 921–932. <http://doi.org/10.1016/j.rser.2011.09.013>
- Mbohwa, C. (2013). Energy Management in the South African Sugar Industry. *Proceedings of the World Congress on Engineering, I*(Vol I), 3–8.
- Meadows, D. M., Schumann, G. T., & Soji, C. (1998). Farewell to filters: The recycle of clarifier mud to the diffuser. *Proc S Afr Sug Technol Ass*, 72, 198–203.
- Modesto, M., Zemp, R. J., & Nebra, S. A. (2009). Ethanol production from sugar cane: Assessing the possibilities of improving energy efficiency through exergetic cost analysis. *Heat Transfer Engineering*, 30(4), 272–281. <http://doi.org/10.1080/01457630802380386>
- Moran, M. J., Shapiro, H. N., Boettner, D. D., & Bailey, M. B. (2011). *Fundamentals of Engineering Thermodynamics* (7th ed.). John Wiley & Sons, Inc.
- Nebra, S. A. (2005). The Exergy of Sucrose-Water Solutions : Proposal of a calculation method. In *Proceedings of ECOS* (pp. 385–392). Trondheim, Norway.
- Ojeda, K., Sánchez, E., & Kafarov, V. (2011). Sustainable ethanol production from lignocellulosic biomass - Application of exergy analysis. *Energy*, 36(4), 2119–2128. <http://doi.org/10.1016/j.energy.2010.08.017>
- Pellegrini, L. F., & de Oliveira, S. (2006). Exergy Analysis of Different Configurations of Multiple-effect Evaporators in Sugarcane Mills. In *Proceedings of the 11th Brazilian Congress of Thermal*

Sciences and Engineering.

- Pippo, W. A., & Luengo, C. A. (2013). Sugarcane energy use : Accounting of feedstock energy considering current agro-industrial trends and their feasibility. *International Journal of Energy and Environmental Engineering*, 4(10), 1–13.
- Querol, E., Gonzalez-Regueral, B., & Perez-Benedito, J. L. (2013). Exergy Concept and Determination. In *Practical Approach to Exergy and Thermo-economic Analyses of Industrial Processes* (pp. 9–28). SpringerBriefs in Energy. http://doi.org/10.1007/978-1-4471-4622-3_2
- Reid, B. M. J., & Rein, P. W. (1983). Steam balance for the new Felixton II mill. In *Proceedings of The South African Sugar Technologists' Association* (pp. 85–91).
- Rein, P. (2007). *Cane Sugar Engineering*. (A. Bartens, Ed.). Berlin.
- Rosen, M. A. (2002). Clarifying thermodynamic efficiencies and losses via exergy. *Exergy, An International Journal*, 2(1), 3–5. [http://doi.org/10.1016/S1164-0235\(01\)00054-1](http://doi.org/10.1016/S1164-0235(01)00054-1)
- Şahin, H. M., Acir, A., Altunok, T., Baysal, E., & Koçyiğit, E. (2010). Analysis of exergy and energy of sugar production process in sugar plant. *Journal of the Energy Institute*, 83(3), 178–185. <http://doi.org/10.1179/014426010X12759937396911>
- Sciubba, E., & Wall, G. (2007). A brief commented history of exergy from the beginnings to 2004. *International Journal of Thermodynamics*, 10(1), 1–26. <http://doi.org/10.5541/ijot.184>
- Seebaluck, V., Mohee, R., Sobhanbabu, P. R. K., Leal, M. R. L. V., & Johnson, F. X. (2008). *Bioenergy for sustainable development and global competitiveness : The case of sugar cane in Southern Africa*. Stockholm, Sweden.
- Sogut, Z., Ilten, N., & Oktay, Z. (2010). Energetic and exergetic performance evaluation of the quadruple-effect evaporator unit in tomato paste production. *Energy*, 35(9), 3821–3826. <http://doi.org/10.1016/j.energy.2010.05.035>
- Sohel, M. I., & Jack, M. W. (2011). Thermodynamic analysis of lignocellulosic biofuel production via a biochemical process : Guiding technology selection and research focus. *Bioresource Technology*, 102(3), 2617–2622. <http://doi.org/10.1016/j.biortech.2010.10.032>
- Starzak, M., & Davis, S. (2016). MATLAB ® modelling of a sugar mill : model development and validation. In *Proc S Afr Sug Technol Ass* (pp. 517–536).
- Starzak, M., & Zizhou, N. (2015). *Biorefinery techno-economic modelling: Sugar mill and ethanol distillery process model*. Durban.

- Taner, T., & Sivrioglu, M. (2015). Energy-exergy analysis and optimisation of a model sugar factory in Turkey. *Energy*, 93, 641–654. <http://doi.org/10.1016/j.energy.2015.09.007>
- Taner, T., & Sivrioglu, M. (2017). A techno-economic & cost analysis of a turbine power plant: A case study for sugar plant. *Renewable and Sustainable Energy Reviews*, 78, 722–730. <http://doi.org/10.1016/j.rser.2017.04.104>
- Tekin, T., & Bayramoğlu, M. (1998a). Exergy analysis of the sugar production process from sugar beets. *International Journal of Energy Research*, 22(7), 591–601. [http://doi.org/10.1002/\(SICI\)1099-114X\(19980610\)22:7<591::AID-ER360>3.0.CO;2-D](http://doi.org/10.1002/(SICI)1099-114X(19980610)22:7<591::AID-ER360>3.0.CO;2-D)
- Tekin, T., & Bayramoğlu, M. (1998b). Exergy loss minimization analysis of sugar production process from sugar beet. *Trans IchemE*, 76(September), 149–154.
- Tekin, T., & Bayramoğlu, M. (2001). Exergy and structural analysis of raw juice production and steam-power units of a sugar production plant. *Energy*, 26(3), 287–297. [http://doi.org/10.1016/S0360-5442\(00\)00068-2](http://doi.org/10.1016/S0360-5442(00)00068-2)
- U.S. Department of Energy. (2011). *Cooling towers : Understanding key components of cooling towers and how to improve water efficiency*. FEMP, Washington.
- Wall, G. (2009). *Exergetics*. Bucaramanga: Exergy, Ecology, Democracy.
- Wang, Z., Fan, W., Zhang, G., & Dong, S. (2016). Exergy analysis of methane cracking thermally coupled with chemical looping combustion for hydrogen production. *Applied Energy*, 168, 1–12. <http://doi.org/10.1016/j.apenergy.2016.01.076>

Appendix

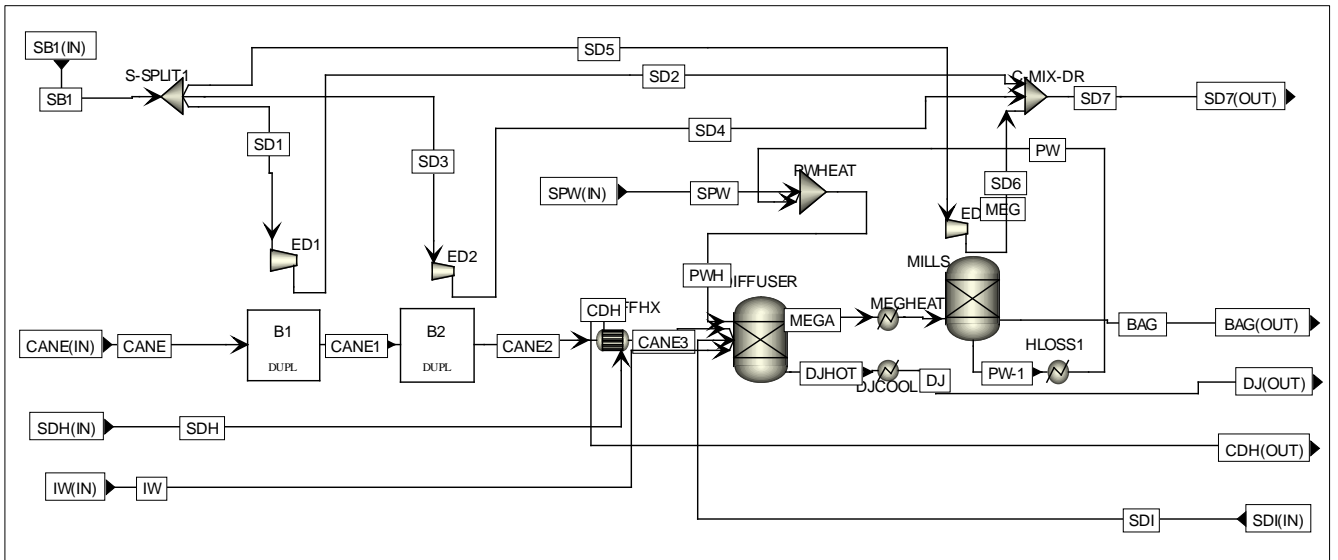


Figure A-3-11: Aspen flowsheet of sugar mill extraction unit

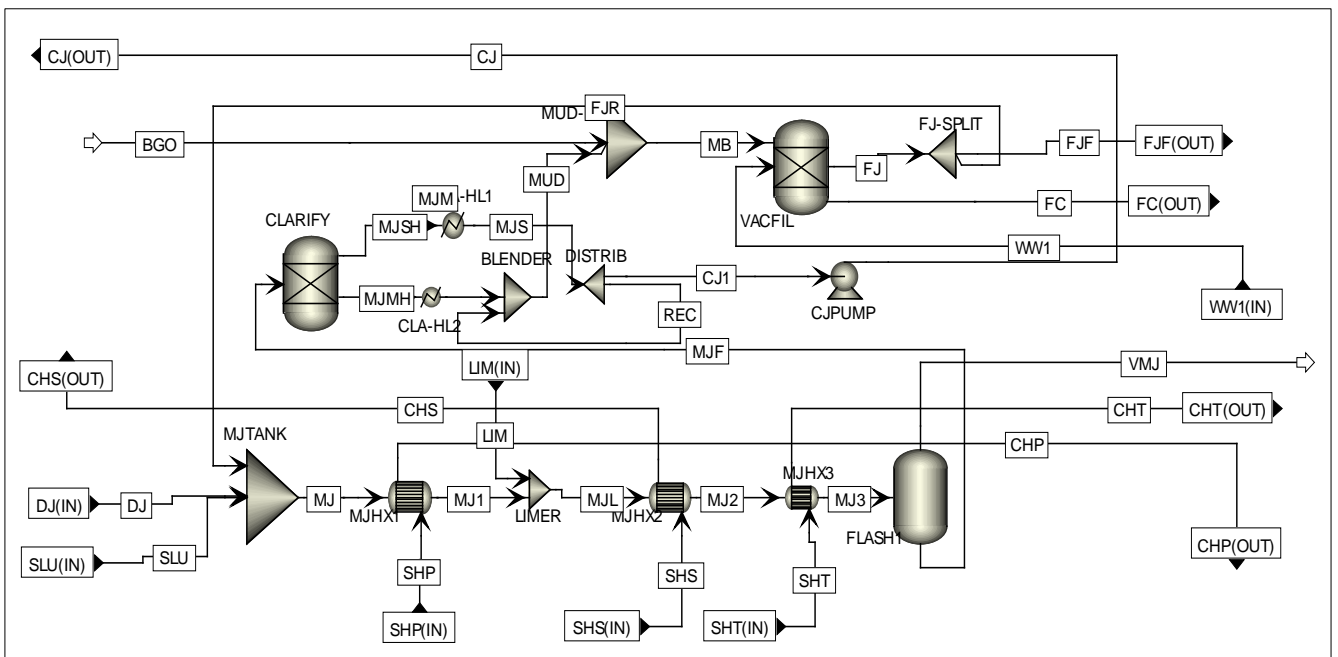


Figure A-3-12: Aspen flowsheet of sugar mill clarification unit

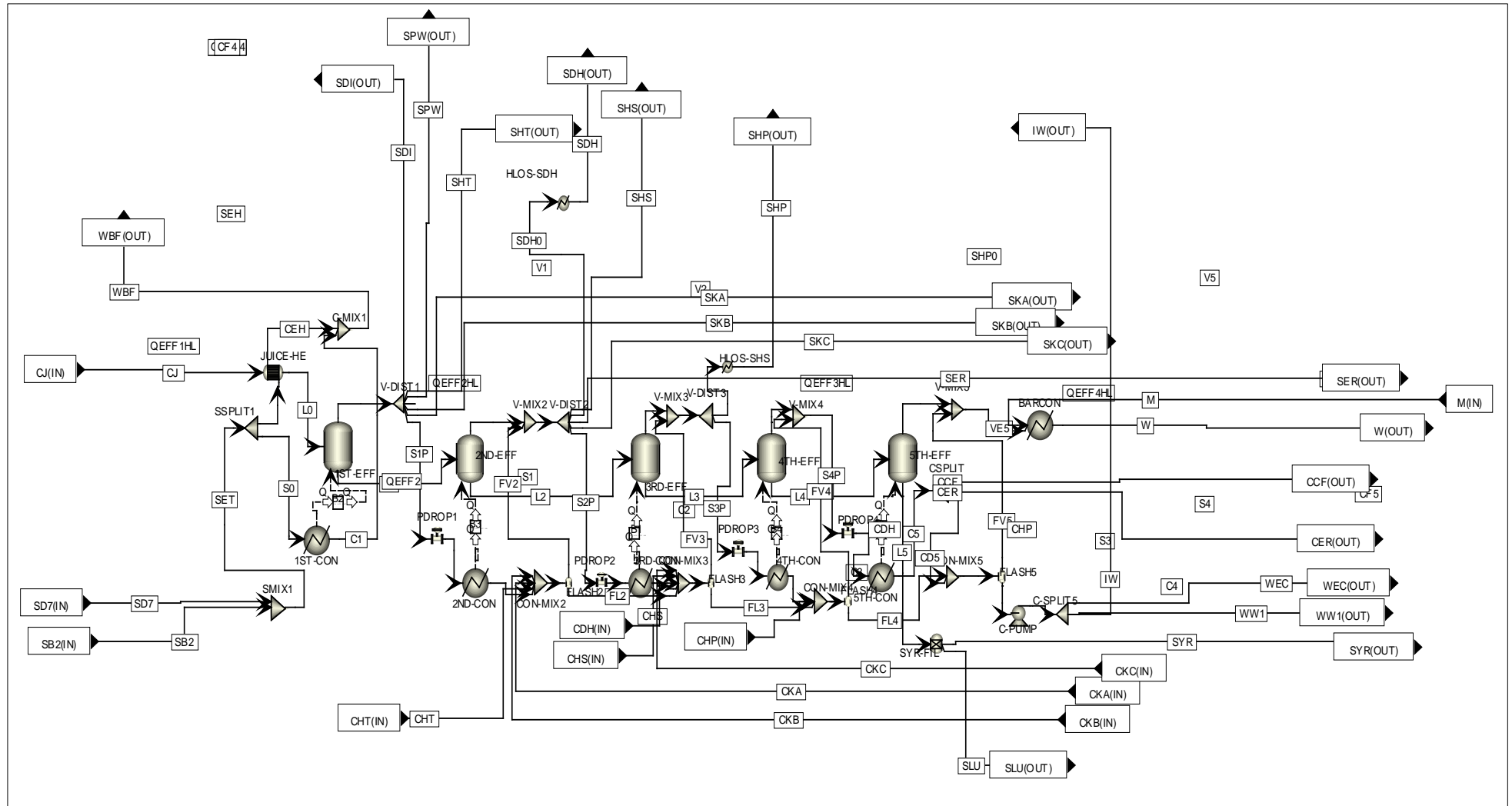


Figure A-3-13: Aspen flowsheet of sugar mill evaporation unit

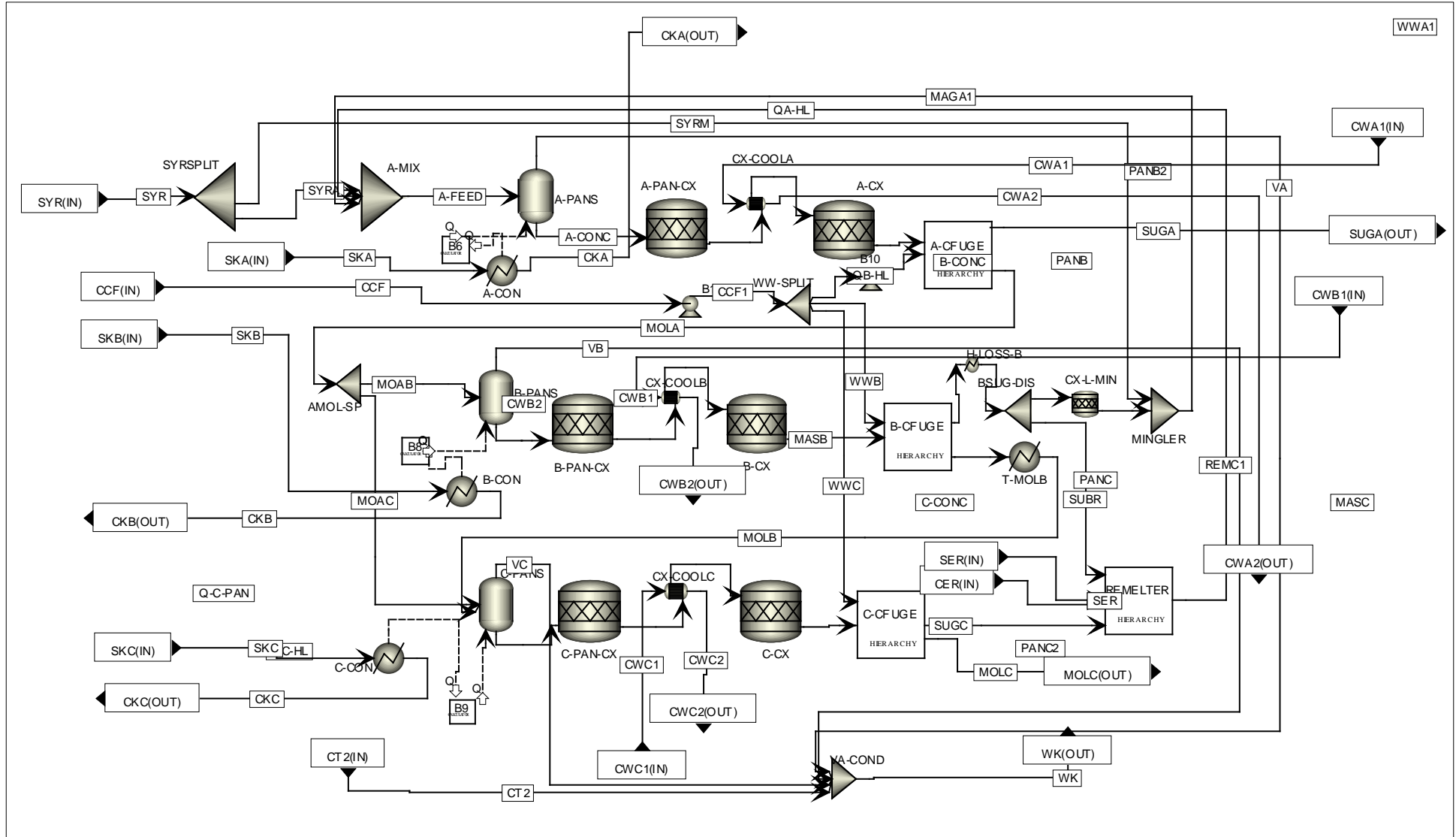


Figure A-3-14: Aspen flowsheet of sugar mill crystallization unit

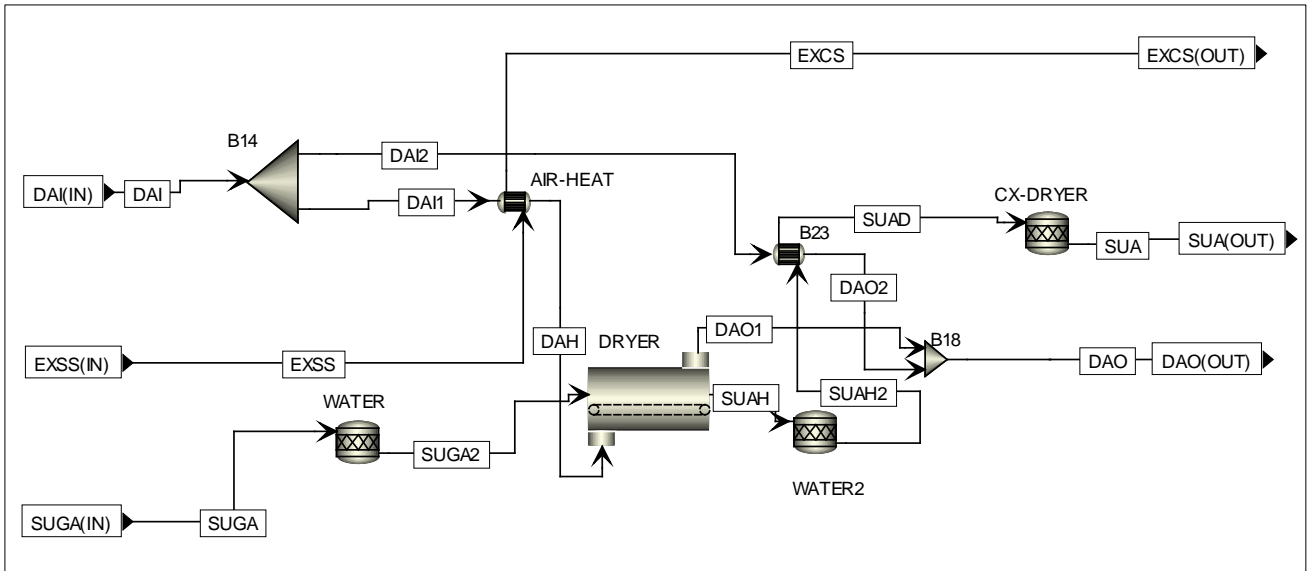


Figure A-3-15: Aspen flowsheet of sugar mill drying unit

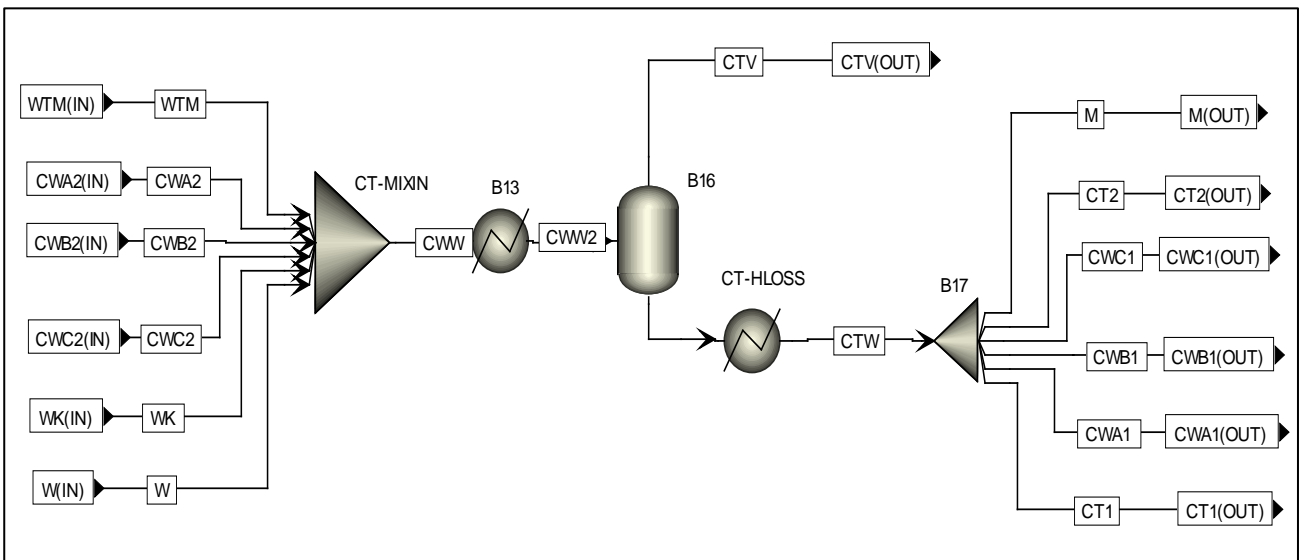


Figure A-3-16: Aspen flowsheet of sugar mill cooling tower unit

Chapter 4

Assessment of the thermodynamic performance improvement of a typical sugar mill through the integration of waste-heat recovery technologies

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Objective of dissertation in this chapter and summary of findings

In Chapter 3, the inefficiencies of the processes in a raw sugar mill were assessed through exergy analysis except for the cogeneration system. Literature survey has revealed that most exergy destruction in a sugar mill takes place in the cogeneration unit and should be the most important unit to consider for improving the exergy efficiency of the sugar mill. This chapter focused on the cogeneration system of the sugar mill to locate the magnitude and causes of its inefficiency (objective 1) and the selection of suitable energy-efficient technologies for integration to improve the energy efficiency of the process (objective 2) while considering other irreversibility sources revealed in Chapter 3. The economic feasibility of integrating the selected technologies was also evaluated (objective 3).

The results showed that the boiler was responsible for 96 % of the total cogeneration exergy destruction mainly due to combustion irreversibility. While higher-pressure boiler could improve the efficiency, they are expensive and lower-cost investments that are complimentary should be done first. Moreover, most sugar mills follow an incremental approach to improving energy efficiency. Therefore, the organic Rankine cycle (ORC) and absorption heat pump (AHP) were selected for waste heat recovery due to the contribution of waste heat to the boiler inefficiency and with considerations

of improvement needs revealed by the exergy analysis of the main sugar mill process (Chapter 3), as well as literature reports.

The AHP integration reduced the irreversibility of the air heat of the drying unit and the deaerator of the cogeneration unit and simultaneously provided alternative cooling to support the cooling tower while the ORC integration improved the cogeneration energy efficiency by producing bagasse-free electricity sale, both recovering exergy from waste heat. Even though minimal exergy improvements were achieved by ORC and AHP, both technologies were economically feasible and could be explored for improving sugar mill efficiency.

Based on the results of this chapter, the possibility of recovering waste heat using absorption chiller to provide chilled water was considered in the designing and costing of integrated biorefineries for the sugar mill in Chapter 5 and Chapter 6. ORC may be considered for more electricity-intensive processes.

Declaration by the candidate

With regards to Chapter 4, pages 114 - 157, the nature and scope of my contribution were as follows:

Nature of contribution	Extent of contribution (%)
Project and scope definition, Aspen simulation work, analysis of data, interpretation of results and writing of manuscript	80

The following co-authors have contributed to Chapter 4, pages 114 - 157:

Name	e-mail address	Nature of contribution	Extent of contribution (%)
J.F. Görgens	jgorgens@sun.ac.za	Provided writing assistance through review and proof reading of manuscript and general discussion	5
M.A. Mandegari	mandegari@sun.ac.za	Reviewed Aspen simulations, and provided writing assistance through suggestions, review and proof reading of manuscript and general discussions	15

Signature of candidate:

Date:

Declaration by co-authors:

The undersigned hereby confirm that

1. the declaration above accurately reflects the nature and extent of the contributions of the candidate and the co-authors to Chapter 4, pages 114 - 157,
2. no other authors contributed to Chapter 4, pages 114 - 157 besides those specified above, and
3. potential conflicts of interest have been revealed to all interested parties and that the necessary arrangements have been made to use the material in Chapter 4, pages 114 - 157, of this dissertation.

Signature	Institutional affiliation	Date
	Stellenbosch University	
	Stellenbosch University	

Assessment of the thermodynamic performance improvement of a typical sugar mill through the integration of waste-heat recovery technologies

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Abstract

The cogeneration system is the most inefficient unit of a typical sugar mill. This paper evaluates the improvements in the exergy performance of a sugar-mill cogeneration system, through the integration of the absorption heat pump (AHP) and the organic Rankine cycle (ORC) technologies for waste-heat recovery. The cogeneration system was simulated in Aspen Plus® process simulator and its thermodynamic performance determined through exergy analysis. In addition to providing an alternative cooling for the mill, the AHP integration minimized the irreversibility of the deaerator, sugar-drying unit and the overall system by 51 %, 76 % and 0.14 %, respectively, saving 0.83 % of total bagasse combusted. The ORC increased the cogeneration system's exergy efficiency by 1.7 % by generating "green" electricity up to 10 % of the cogeneration capacity. The AHP and ORC were economically feasible with payback periods of 8.3 years and 6.3 years, respectively. This paper provides basis for waste-heat recovery based exergy improvement studies in other process industries.

Keywords: Absorption heat pump; Aspen Plus® simulation; Cogeneration system; Economic feasibility analysis; Exergy analysis; Organic Rankine cycle.

4.1. Introduction

Sugar is one of the most important agricultural products and yields an annual production of 170 million tonnes (USDA, 2018) from sugarcane and sugar beet, in over 130 countries. About 80 % of global sugar production is from sugarcane, which means that more than 28 million hectares of land are used for sugarcane production across the world ("FAOSTAT," n.d.; International Sugar Organization, 2016). The sugar production process is energy self-sufficient, that is, energy supply is from burning the sugarcane residue (bagasse) in a cogeneration system (Aghbashlo, Mandegari, et al., 2018b).

The cogeneration systems were traditionally designed to dispose of all available bagasse by burning it in inefficient boilers (Pellegrini et al., 2010). However, bagasse as a pre-treated accessible lignocellulosic residue is currently gaining attention for valorisation in biorefineries (M. A. Mandegari et al., 2017). Moreover, some sugar mills in South Africa now burn coal to supplement bagasse fuel due to facility ageing, steam demand of various auxiliary processes (including process transients) and the demand on bagasse for by-products (M. J. Reid, 2006). For instance, South African sugar mills consume on average 11 t of coal per 1000 t of cane processed (Hess et al., 2016), which is economically and environmentally undesirable. In order to reduce coal consumption and to make bagasse available for high-value by-products, it is essential to improve the energy efficiency of the sugar mill.

Exergy analysis is an advanced thermodynamic method that reveals the locations, causes and magnitudes of process inefficiencies in order to improve their performance, and is supplementary to the conventional energy analysis (Ghannadzadeh & Sadeqzadeh, 2017). Exergy studies of sugar mills showed that the cogeneration system, mainly the boiler, accounted for more than 70 % of the sugar industry's thermodynamic inefficiencies (Hector, Carolina Mesa, & Sergio, 2013). Therefore, improvements in the cogeneration system are most significant for increasing the energy performance of a sugar mill.

Researchers showed that increasing steam temperature in high-pressure boilers increased the boiler exergetic efficiency (Kamate & Gangavati, 2009b). However, using advanced cogeneration systems, including condensing-extraction steam turbine (CEST) and biomass integrated gasification combined cycle (BIGCC) in combination with high-pressure boilers for surplus electricity production, improved the overall cogeneration efficiency (Pellegrini et al., 2010). Ensinas et al (A. V. Ensinas et al., 2009) conducted exergy analysis of a sugar and ethanol plant including a cogeneration system and found that increasing the inlet steam temperature from 300 °C to 520 °C, increasing pressure from 22 bar to 90 bar and using CEST to increase surplus electricity production improved the exergy efficiency from 18 % to 23 %.

However, high levels of a steam economy would be required to implement these technologies (A. V. Ensinas et al., 2009). Moreover, even though the CEST is the preferred cogeneration system in modern sugar mills, it has a higher operating cost (Deshmukh et al., 2013; Kamate & Gangavati, 2009b), and is therefore only desirable when electricity generation for sale is a priority. On the other hand, even though the BIGCC promises up to three times the power generated by the CEST, there is no commercial BIGCC in the sugar industry. This is due to its high installation cost and its inability to meet the process steam demands of a typical sugar mill (Deshmukh et al., 2013). Therefore, this

study explored the possibility of improving the cogeneration exergy performance through waste-heat recovery.

Efforts have been made towards improving the sugar industry's energy efficiency through energy monitoring (Mkwananzi, Mandegari, & Görgens, 2019) and the concept of Pinch Analysis to utilise waste heat, especially by means of economizers and air heaters in the cogeneration system (Rein, 2007). However, opportunities to recover low-grade waste heat using commercial technologies such as absorption heat pumps (AHPs), absorption chillers and organic Rankine cycles (ORCs) (Oluleye, Jobson, Smith, & Perry, 2016) still exist for increasing energy efficiency. Balaji & Ramkumar (2012) studied waste-heat recovery from the steam turbine exhaust in a sugar mill to provide cooling by using absorption chillers and reported nearly 75 t/y fuel savings using this application. Similarly, Chouhan (2014) proposed waste-heat recovery (from flue gases, boiler blowdown and evaporator condensates) using absorption chillers to provide an alternate cooling in the sugar industry, resulting in annual bagasse savings of 1870 t. However, these studies do not account for the heat released by the absorber and condenser of the absorption chiller to the environment through a cooling tower. There cooling tower may be inefficient due to unfavourable ambient conditions. Conversely, Zhang et al. (2011) introduced a combined cold, heat and power plant where an electric heat pump recovered heat from the cooling water of a steam-driven LiBr chiller to pre-heat the demineralised water of the boiler. No studies were found in the literature on the integration of a waste-heat driven absorption heat pump for simultaneous heating and cooling in the sugar industry. Moreso, a recent review of waste heat recovery by ORC reveals that recovering industrial process waste heat by ORC has not received adequate attention (Mahmoudi, Fazli, & Morad, 2018). Even though various studies have evaluated the exergy performance AHP and ORC systems (Ahmed, Esmaeil, Irfan, & Al-Mufadi, 2018; Colorado, Demesa, Huicochea, & Hernández, 2016), no studies have assessed the improvement on the exergy performance of the existing process into which they are integrated, particularly the sugar mill. Ferrer et al. (Ferrer, Mezquita, Aguilera, & Monfort, 2019) investigated the most critical parameters affecting kiln energy efficiency through exergy analysis. The authors proposed that waste heat recovery could significantly increase kiln energy performance but no evaluation of the improvements was reported.

This study assesses the exergy improvement of a sugar mill cogeneration system through the integration of waste-heat driven AHP for simultaneous cooling and heating, and ORC for "green" electricity production. In addition, the economic feasibility of these integrations was evaluated. Finally, the study addressed cooling capacity problem of the sugar industry on hot and humid days.

The first step in the study procedure develops an Aspen Plus[®] simulation and evaluating the thermodynamic performance of the cogeneration system of a typical sugar mill, corresponding to and

in continuation of the sugar mill exergy study reported previously (Chapter 3). The second step integrates ORC and AHP into the sugar plant, based on the exergy results, and evaluates the improvement in the exergy performance. The final step assesses the profitability of implementing these integrations in the sugar mill, to assist decision-making. The findings of these assessments could serve as a guide for exergy improvement options in various process industries.

4.2. Description, simulation and exergy analysis of the cogeneration system

4.2.1. Process description

The cogeneration system is the powerhouse of the sugar mill and consists mainly of the feed water, the boiler and the turbine sub-systems. Figure 4-1 represents; (a) the process flow of the cogeneration system and (b) the corresponding Aspen simulation flowsheet. In the feed water unit, hot condensates from the sugar production process are stored at atmospheric pressure and at a temperature of about 90 °C in a storage tank (according to the sugar industry practice), while boiler make-up water is pumped to the deaerator to remove air and other non-condensable gases by venting them (ABMA, 2011). The treated feed water is then sent to the boiler to produce high-pressure (HP) steam using hot flue gasses produced in a furnace by combusting bagasse (Chapter 3). Before exiting the boiler, the flue gas is used to preheat the combustion air followed by the feed water in an economizer and released to the atmosphere at 190 °C (Wienese, 2001). A back-pressure turbine expands some of the HP steam at an exhaust steam pressure of about 2.5 bar to generate electricity for the process (A. V. Ensinas et al., 2009), and in some instances, surplus electricity for sale. The remaining HP steam drives turbines for knives, shredders and mills in the extraction unit of the mill. The addition of condensates to the exhaust steam de-superheats it to the required steam conditions for the main process.

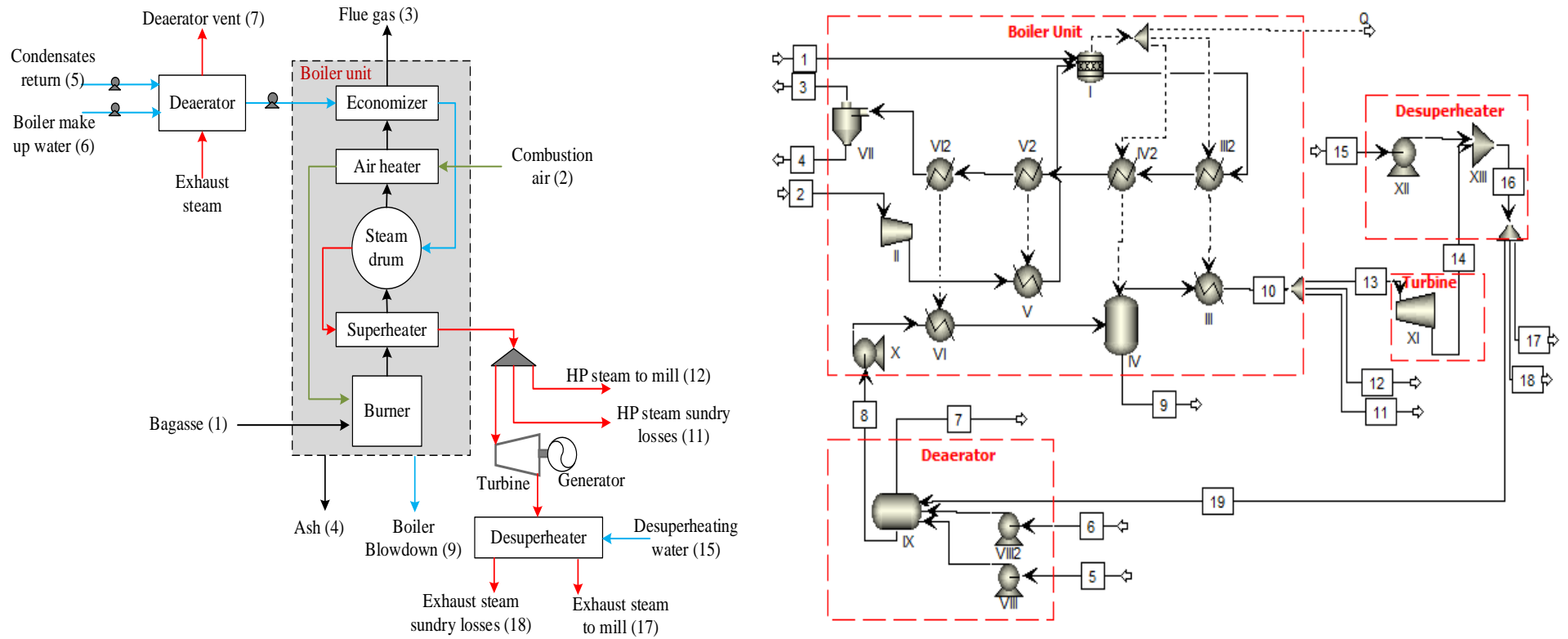


Figure 4-1: A typical Sugar mill Cogeneration system (a) Process flow diagram (b) Aspen simulation flowsheet

4.2.2. Aspen Plus® simulation

In order to obtain the mass and energy balances and to determine the thermophysical properties of streams that form the basis for exergy analysis, the cogeneration system of a typical sugar mill that processes 250 ton of cane per hour was simulated in Aspen Plus® under steady-state conditions, as depicted in Figure 4-1 (b). Aspen Plus® henceforth referred to as Aspen, is a well-known chemical process simulation software with optimization, sensitivity analysis and economic tools for the complete evaluation of the process performance (Fontalvo Alzate, 2014).

The simulation was developed using the UNIQUAC (UNIversal QUAsi Chemical) property method recommended for calculating the thermodynamic properties of non-ideal systems (Aspentech, 2013). The studies of Palacios-Bereche & Nebra (2009) and Nsaful et al. (2013) showed that Aspen is suitable for the simulation of cogeneration systems with accurate results. Table A.4-1 and Table A4-.2 in the Appendix present the data for the simulation of the cogeneration system and the bagasse composition used for the simulation, respectively.

In addition, steam losses were assumed as 2.5 % and 6.5 % of the total HP and exhaust steam, respectively (B. M. J. Reid & Rein, 1983). All radiation and convective losses from the furnace and heat exchangers to the atmosphere were assumed as 1.5 % of the gross calorific value (GCV) of bagasse (Rein, 2007) and taken from the furnace as a single heat stream. The combustion reactions in equations (A.4-1) to (A.4-7) of the Appendix were simulated with 0.99 conversion on each bagasse component to account for unburnt carbon losses of 1.5% estimated for bagasse boilers and 25 % excess air to ensure complete combustion and no carbon monoxide production (Wienese, 2001). Equations (A.4-8) to (A.4-10) of the Appendix give the calculations for the required amount of air for combustion. Equation (4-1) shows the calculation of bagasse gross calorific value (GCV) (Rein, 2007; Wienese, 2001).

$$GCV = [19605 - 196.05 M - 196.05 A - 31.14 B] \text{ kJ/kg} \quad (4-1)$$

The boiler efficiency, defined as the ratio of the rate of energy output to the rate of energy input (Rein, 2007), was calculated using equation (4-2) (Wienese, 2001).

$$\eta = \frac{(\dot{m}_{st} \times h_{st}) - (\dot{m}_{FW} \times h_{FW})}{\dot{m}_{Fuel} \times GCV_{Fuel}} \quad (4-2)$$

4.2.3. Formulations for exergy analysis

The total exergy flow of streams was evaluated in Excel using the appropriate equations described in the literature (Enrique Querol et al., 2013; Utlu, Sogut, Hepbasli, & Oktay, 2006), (Chapter 3). The physical exergy of streams was evaluated through the Aspen simulation based on the reference

temperature and pressure of 25 °C and 1.01325 bar, respectively (Kotas, 1995), whereas the chemical exergy was determined from the specific chemical exergy of substances involved in the process (reported in Table A.4-3 of the Appendix) and their respective mole ratios. The exergy performance indicators, irreversibility (\dot{I}) and rational exergy efficiency (ψ) were calculated using equations (4-3) and (4-4), respectively (Utlu et al., 2006).

$$\dot{I} = \sum \dot{E}x_{in} - \sum \dot{E}x_{product} - \sum \dot{E}x_w \quad (4-3)$$

$$\psi = \frac{\dot{E}x_{desired\ product}}{\dot{E}x_{used}} \quad (4-4)$$

The expressions for the irreversibility and rational efficiency for this application are described by equations (4-5) and (4-6), respectively with reference to the stream numbers in Figure 4-1.

$$\dot{I} = (\dot{E}x_1 + \dot{E}x_2 + \dot{E}x_5 + \dot{E}x_6 + \dot{E}x_{15} + \dot{W}_{in}) - (\dot{E}x_{12} + \dot{E}x_{17} + \dot{W}_{out}) - (\dot{E}x_3^w + \dot{E}x_4^w + \dot{E}x_7^w + \dot{E}x_9^w + \dot{E}x_{11}^w + \dot{E}x_{18}^w) \quad (4-5)$$

$$\psi = \frac{(\dot{E}x_{12} + \dot{E}x_{17} + \dot{W}_{out}) - (\dot{E}x_5 + \dot{E}x_6 + \dot{E}x_{15})}{(\dot{E}x_1 + \dot{E}x_2 + \dot{W}_{in}) - (\dot{E}x_3 + \dot{E}x_4)} \quad (4-6)$$

The rational exergy efficiency represents the percentage of the fuel exergy expended in the cogeneration system to produce steam and electricity for the sugar mill.

4.3. Integration of AHP and ORC into a typical sugar plant

The integration of AHP was based on the exergy results of the cogeneration system in this study, as well as the results and recommendations of the exergy study of the corresponding sugar mill reported in (Chapter 3). Therefore, this study considers AHP for simultaneous heating of sugar drying air and cold streams within the cogeneration system and cooling of process warm water from the sugar mill. The absorber and condenser release heat for the useful pre-heating, while the evaporator takes up heat from the process warm water and by that, cooling it without the cooling tower (Chouhan, 2014). The operation of the AHP was dependent on the waste heat available in the sugar mill and the cogeneration system, as well as the heating and cooling requirements of the mill. In order to compare the exergy improvements and cost-effectiveness of the AHP, the study also considers ORC integration as an alternative waste-heat recovery technology for power generation.

4.3.1. Process simulation of the integrated AHP and ORC

Absorption heat pump consists of an evaporator, an absorber, a generator and a condenser, with a refrigerant-absorbent working fluid to provide heating, cooling or both. Water-lithium bromide and ammonia-water solutions are the most commonly used working pair. Water-lithium bromide AHP is

considered in this study because it gives higher performance than the ammonia-water solution and has no working fluid toxicity issues, unlike ammonia which is slightly toxic (Kurem & Horuz, 2001; Somers et al., 2011). Figure 4-2 (a) illustrates the operating principle of the absorption heat pump as described in Somers et al. (2011). Points 1 – 6 in Figure 4-2 (a) represent the solution cycle while points 7 – 10 represent the refrigerant cycle.

The simulation of the LiBr/water AHP shown in Figure 4-2 (b) was based on the assumptions used by Somers et al (2011) in developing Aspen simulation of an absorption chiller, using the operating parameters presented in Table A.4-4 of the Appendix. The thermodynamic property method used for the simulation was the unsymmetric electrolyte NRTL (Non-Random-Two-Liquid) model with the Redlich-Kwong equation of state and Henry's law (ENRTL-RK), which is applicable for liquid electrolyte solutions containing water (Aspentech, 2013). Furthermore, NBS steam tables (STEAMNBS) and Peng-Robinson equation of state (PENG-ROB) property methods were used for blocks with pure refrigerant water and air streams, respectively.

The temperature and concentration of the strong solution were chosen from the thermodynamic design data for water-lithium bromide absorption systems for simultaneous cooling and heating (Eisa, Devotta, & Holland, 1986), based on the temperatures of the available waste-heat source. These were used to derive the pump exit pressure from water-lithium bromide solution equilibrium chart (Oluleye et al., 2016). The waste heat available also determines the separation ratio of refrigerant in the generator, based on the strong solution concentration.

The solution heat exchanger was modelled with two Aspen heater blocks and a heat stream, specifying hot stream outlet temperature, T_5 determined by equation (4-7), using an assumed value of 0.8 for ϵ (Lansing, 1976) and known values of T_4 and T_2 (temperatures of the solutions exiting the generator and the solution pump, respectively) (Eisa et al., 1986).

$$\epsilon = \frac{T_4 - T_5}{T_4 - T_2} \quad (4-7)$$

The coefficient of performance of the heat pump as expressed in equation (4-8), is defined by the ratio of the heat recovered in the absorber and the condenser to the waste heat flow to the generator and the energy flow in the pumps (Oluleye et al., 2016).

$$COP = \frac{\dot{Q}_{absorber} + \dot{Q}_{condenser}}{\dot{Q}_{waste\ heat} + \dot{W}_{pump}} \quad (4-8)$$

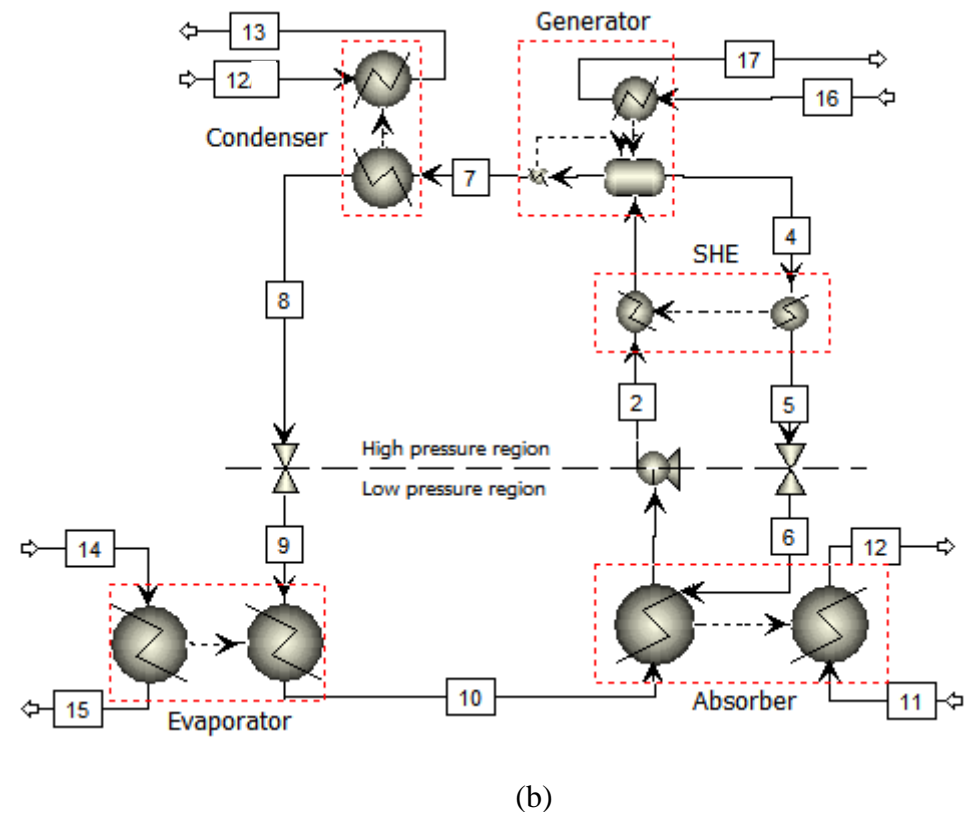
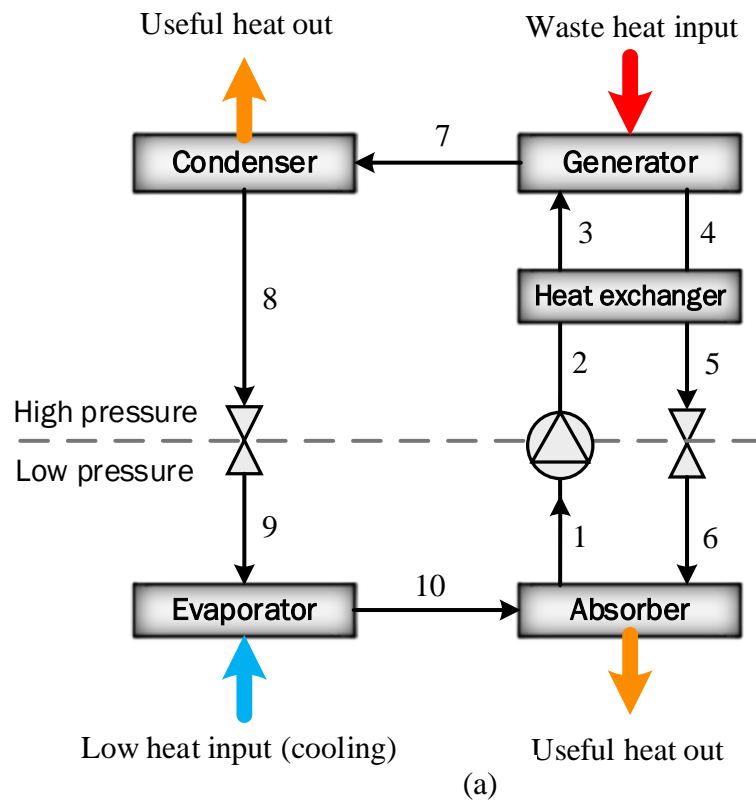


Figure 4-2: Absorption heat pump: (a) process flow diagram (b) Aspen simulation

Organic Rankine cycle (ORC) is a commercial thermodynamic cycle that uses organic fluids with high molecular weights and low boiling points as working fluids, instead of water, to convert low-grade heat to electricity (C. W. Chan et al., 2013). Figure 4-3 depicts the working principle of a simple ORC. A more advanced configuration, the regenerative ORC uses the working fluid at the turbine exit to preheat the fluid entering the evaporator for improved heat recovery. However, for low to medium temperature heat sources, the efficiency improvement when using a recuperator is low (Uusitalo, Honkatukia, Turunen-Saaresti, & Grönman, 2018) and does not warrant the additional cost involved.

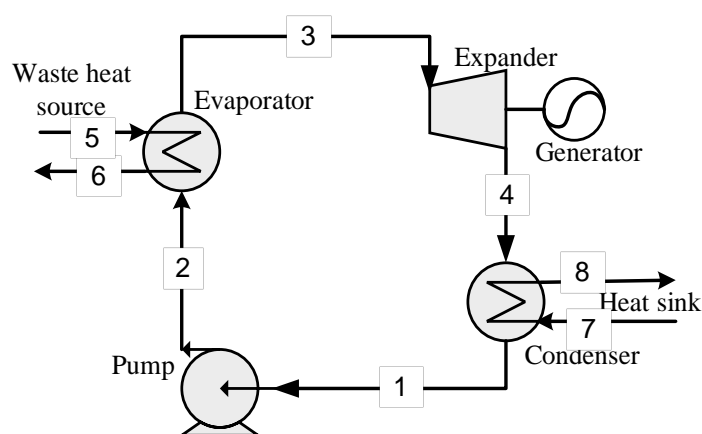


Figure 4-3: Organic Rankine cycle for low-temperature waste-heat recovery

The working fluid selected for this work is n-pentane, which is among the fluids reported suitable for low to medium heat source applications, due to its relatively low critical temperature $196.6\text{ }^{\circ}\text{C}$ to ensure maximum waste heat to electricity efficiency at a relatively low critical pressure of 33.4 bar (Uusitalo et al., 2018). In addition, it has zero ozone depletion potential and low global warming potential, which guarantees its environmental safety. Besides, n-pentane is easily available as it is used in commercial ORCs for waste-heat recovery (Quoilin, Broek, Declaye, Dewallef, & Lemort, 2013).

The ORC was simulated in Aspen using the Peng-Robinson equation of state property method recommended for hydrocarbon applications (Aspentech, 2013). The UNIQUAC property method was used for the simulation of the hot side of the evaporator and the cold side of the condenser. The assumptions adopted for the simulation were reported in the literature (Xi, Li, He, & Tao, 2015).

The mass flows of working fluid and cooling water were set to ensure there is no internal temperature cross in the evaporator and condenser respectively. Similarly, the operating pressure of 18 bar, which is below the pressure at the maximum entropy point on the n-pentane P-H diagram

(Perry & Green, 1965), was chosen to ensure no temperature cross. The thermal efficiency of the ORC technology is the ratio of net power produced to the waste heat recovered as expressed by (4-9).

$$\eta_{ORC} = \frac{\dot{W}_{expander} - \dot{W}_{pump}}{\dot{Q}_{waste\ heat}} \quad (4-9)$$

4.3.2. Economic analysis of AHP and ORC

The technical and economic feasibility of heat recovery technologies would determine their adoption in the sugarcane mill (Lemmens, 2016). A project cost analysis for the AHP and ORC for the sugar industry which was not reported previously in literature was therefore completed, based on the expected annual cash flows of investments in these technologies, calculated from the capital investments (CAPEX), operating costs (OPEX) and revenue (Lemmens, 2016). The operating cost, which involves annual expenses, is generally of far less importance than the capital investment for waste-heat recovery technologies (Lemmens, 2016). Therefore, this work assumed zero operating costs for ORC and AHP, since these technologies are retrofits of an existing plant to run on waste heat. Table A.4-5 in the Appendix presents the method of determining the total capital investment (TCI) and the factors used to calculate components of TCI based on the purchased equipment costs (PEC). The TCI comprises the fixed capital cost and working capital, which was assumed to be 5 % of the fixed capital cost (Nsaful et al., 2013). The sale of electricity produced from waste-heat recovery (ORC) and the sales of bagasse saved as well as cooling water provided by AHP application were the basis for calculating the revenue for the cash flow analysis.

The purchased equipment costs for the ORC and AHP were determined using Aspen Plus® economic analyser considering the individual units that comprised these technologies due to the challenge in obtaining the cost of ORC and AHP installed specifically for the sugar industry (see Table A.4-6 and Table A.4-7 of the Appendix, respectively). The estimated costs for the technologies were then compared with cost estimates from the literature, most of which are used for preliminary and research purposes (Lemmens, 2016). Cash flow analysis was carried out in an Excel spreadsheet to determine the economic indicators, i.e. internal rate of return (IRR) and the payback period. In addition, a sensitivity analysis was conducted to show the impact that the uncertainties in the estimated input parameters would have on the economic performance of the technologies. Table A.4-8 of the Appendix reports the economic assumptions and input parameters employed for the economic assessment.

4.4. Results and discussion

The mass and energy balances necessary for the exergy analysis of the cogeneration system under study were obtained from the Aspen simulation. Table 4-1 presents the Aspen simulation results and the stream exergies evaluated in Excel.

Table 4-1: Simulation results and mass and energy balances of the cogeneration system

Stream number	Description	\dot{m} [kg/s]	T [°C]	P [bar]	\dot{H}^a [kW]	$\dot{E}x^{ph}$ [kW]	$\dot{E}x$ [kW]
Inlet streams							
1	Bagasse burnt in boiler ^b	14.05	25	1.01	-152463	29	133100
2	Air ^b	48.89	25	1.01	-8531	0	2368
5	Boiler feed water	27.97	90	1.01	-435949	800	2197
6	Boiler make-up water	2.95	25	1.01	-46800	0	147
15	Desuperheating water	0.83	90	1.01	-12937	24	65
\dot{W}_{in}	Power input	-	-	-	242	242	242
Total		94.69	-	-	-656437	-	138121
Outlet stream							
3	Flue gas released to atmosphere ^b	62.63	190	1.01	-246005	4670	11945
4	Ash ^b	0.32	-	-	-3207	11	1742
7	Deaerator vent	0.03	105	1.21	-398	14	16
9	Boiler blowdown	0.94	233	29.60	-13933	280	327

Table 4-1: Simulation results and mass and energy balances of the cogeneration system (continued 1)

Stream number	Description	\dot{m} [kg/s]	T [°C]	P [bar]	\dot{H}^a [kW]	$\dot{E}_{x^{ph}}$ [kW]	\dot{E}_x [kW]
Outlet streams (continued)							
11	HP steam sundry losses	0.77	360	28.60	-9835	886	924
12	HP steam to mill turbines	5.43	360	28.60	-69357	6264	6535
17	Exhaust steam to the main process	22.54	130	2.62	-298101	14326	15452
18	Exhaust steam sundry losses	2.04	130	2.62	-26980	1297	1399
\dot{Q}_{loss}	Heat loss	-	-	-	1931	-	-
\dot{W}_{out}	Power produced	-	-	-	9199	9199	9199
\dot{W}_{loss}	Power loss	-	-	-	188	-	-
Total		94.70	-	-	-656498	-	47539

Table 4-1: Simulation results and mass and energy balances of the cogeneration system (continued 2)

Stream number	Description	\dot{m} [kg/s]	T [°C]	P [bar]	\dot{H}^a [kW]	$\dot{E}x^{ph}$ [kW]	$\dot{E}x$ [kW]
Intermediate streams							
8	Boiler feed water from deaerator	32.19	105	1.21	-499529	1383	2991
10	HP steam from superheater	31.25	360	28.60	-399153	36035	37596
13	HP steam to turbo alternator	25.05	360	28.60	-319961	28886	30137
14	Exhaust steam from turbo alternator	25.05	171	2.62	-329348	16503	17754
16	De-superheated steam	25.88	130	2.62	-342274	16452	17745
19	Steam to deaerator	1.30	130	2.62	-17193	829	894
20	Flue gas from burner	62.95	950	1.01	-185819	44283	53290

^a Aspen Plus reports negative enthalpies because it uses 298 K and 1 atm for the constituent elements as an ideal gas as the reference state. To get the positive enthalpy in the steam tables reference state, 15970.31 kJ/kg is added to the specific enthalpy value to account for the heat of formation of water.

^b The moisture contents of the non-water streams 1, 2 and 3 are 50.8, 13.0 and 18.0 w%, respectively. The ash stream 4 was assumed to have zero moisture.

The system produced net electricity of 36 kWh/t of cane, 4 kWh more than the maximum value of 32 kWh/t of cane reported in the literature for a typical BPST sugar mill cogeneration system (Seebaluck et al., 2008). The lower power consumption obtained could be due to the omission of some auxiliary electrical components from the simulation for lack of appropriate data. The boiler efficiency was 65 % for a furnace and minimum flue gas temperatures of 950 °C and 190 °C, respectively, HP steam conditions of 360 °C and 28.60 bar, and bagasse moisture of 50.84. Rein (2007) reported a boiler efficiency of 67 % for 51 % bagasse moisture and 190 °C flue gas temperature, while Wienese (2001) reported 62.9 % for a steam temperature of 400 °C. In summary, the simulation results of the cogeneration system in this study generally compared well with similar sugar mill cogeneration systems reported in the literature (A. V. Ensinas et al., 2009; Rein, 2007; Wienese, 2001).

4.4.1. Exergy analysis of the cogeneration system

The Grassmann diagram illustrated in Figure 4-4 shows the detailed account of exergy flow through the units of the cogeneration system. The total irreversibility of the system was 90582 kW to which the boiler unit contributed 87066 kW (96 %). The other units (deaerator, turbine and desuperheater) altogether are responsible for the remaining 4 % of the total irreversibility. This finding agrees with the work of Kamate & Gangavati (2009b) who also found that the boiler is the principal exergy destructive unit of a sugar mill cogeneration system.

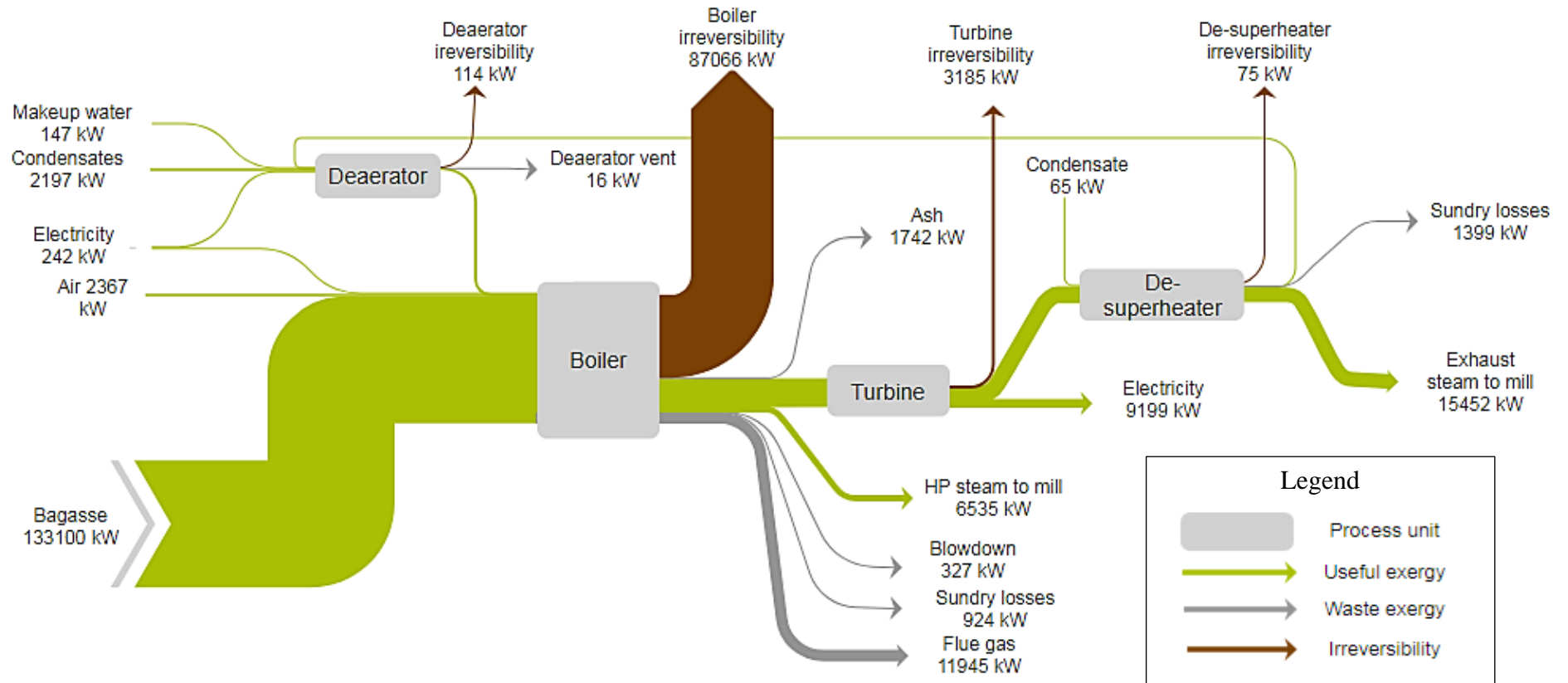


Figure 4-4: Grassmann diagram of the cogeneration system

The rational exergy efficiency 23.6 % compares well with the results reported by Kamate & Gangavati (2009b) for typical sugar mill BPST cogeneration plants as Table 4-2 indicates. It can be seen (Table 4-2) that producing steam at higher temperatures and pressures improves the exergy efficiency of the cogeneration system (Kamate & Gangavati, 2009b) because the resource (input) exergy is expended on producing higher quality (exergy) product, i.e., high temperature and pressure steam. However, where there is no incentive for surplus electricity production from the high temperature and pressure steam, the process steam conditions and power requirements limit this avenue for exergy improvement. Moreover, further extensions of exergy analysis such as exergoeconomic analysis would be required to ascertain the economic impact of replacing the current boiler with a higher pressure/temperature one, relative to the thermodynamic improvement.

Table 4-2: Comparison of the rational exergy efficiency with literature results

Description	Kamate & Gangavati (2009b)			This work
	Plant 1	Plant 2	Plant 3	
Steam temperature [°C]	340	388	423	360
Steam pressure [bar]	21.0	31.0	41.0	28.6
Steam to bagasse ratio	2.11	2.25	2.36	2.22
Rational exergy efficiency [%]	20.6	23.9	27.6	23.6

4.4.2. Waste-heat sources and recovery potentials

Exergy lost through waste streams from the cogeneration system accounts for 4.2 % of the input exergy, which waste-heat recovery could minimize. However, only the physical exergy, which depends on temperature (an important factor in waste-heat recovery) and pressure of the waste streams, is available for recovery. The chemical exergy possessed by the chemical components of the waste streams cannot be recovered through heat recovery, because it is only available at the environmental state (Kotas, 1995). Figure 4-5 presents the various waste streams in the cogeneration system and their physical and

chemical exergy contents. The flue gas has the largest exergy content of 11.95 MW, because of its high flow rate and temperature (refer to Table 4-1), while the deaerator vent had the least exergy content of 0.02 MW. The physical exergy contents of ash and deaerator vent were comparatively negligible (less than 0.02 MW) and are not suitable sources for waste-heat recovery. Similarly, sundry losses are not a consolidated stream heat recovery, leaving flue gas and boiler blowdown as the only streams with waste-heat recovery potential. The previous sugar mill exergy study showed that flash drum vapour (from the clarification unit of the main sugar process) had relatively sufficient physical exergy (0.4 MW) for recovery (Chapter 3). Therefore, this study considered the flue gas and boiler blowdown as well as the flash drum vapour from the main sugar plant as suitable waste-heat sources for recovery.

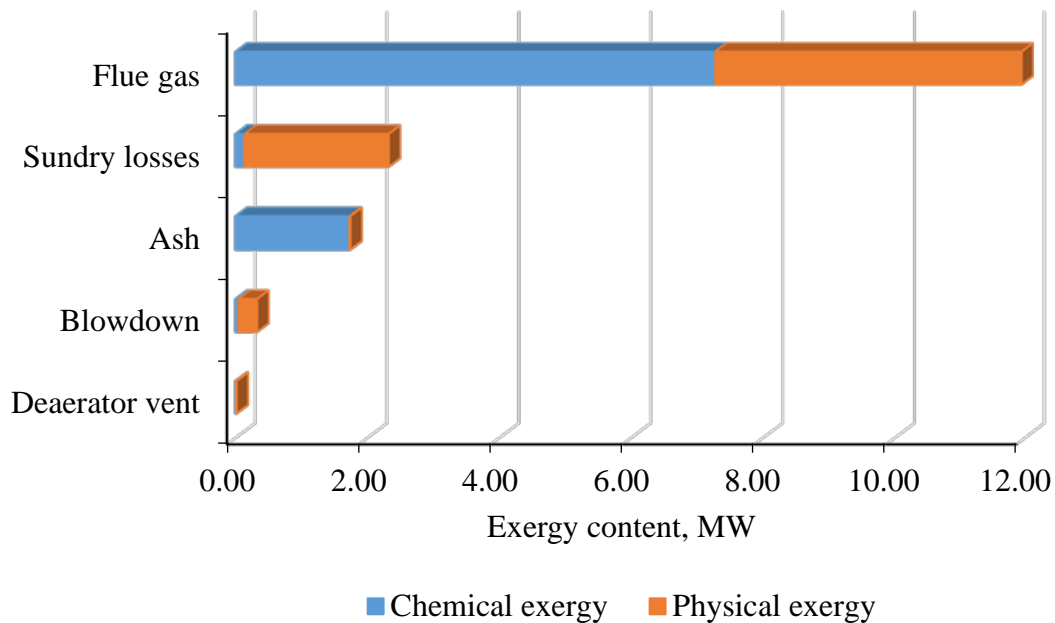


Figure 4-5: Physical and chemical exergies of waste streams in sugar mill cogeneration system

4.4.3. Integration and simulation of waste-heat recovery technologies

There are several ways to improve the exergy efficiency of a system including, reducing resource exergy input, minimizing exergy loss and increasing product exergy. The following sections present the results of heat recovery from the waste-heat sources selected in Section 4.4.2 to minimize exergy loss in these waste streams and

simultaneously (a) minimize irreversibility using AHP and (b) increase product exergy using ORC.

4.4.3.1. Absorption heat pump (AHP)

The AHP integration in this study accomplishes three outcomes simultaneously towards exergy improvement of the sugar plant, viz., (i) recover waste heat in the generator, (ii) pre-heat (referred to as temperature upgrade) cold streams used as heat sinks for the absorber and condenser in succession and (iii) and cool process warm water being used as heat source for the evaporator. The previous chapter (Chapter 3) and this identified the sugar drying air and boiler make-up water, respectively for preheating leading to exergy improvement. The available heat sinks identified; boiler make-up water (1) and sugar drying air (2) and the waste-heat sources selected; flash drum vapour (A), boiler blowdown (B) and flue gas (C) led to the simulation of six scenarios of AHP integrations as represented in Figure 4-6. The label of each AHP scenario bears the letter assigned to the waste-heat source (A, B, and C) followed by either the number “1” or “2” indicating the heat sink used. The simulation results include the exit temperature of the waste-heat source, the upgrade temperature of the heat sink and the amount of process warm-water cooled.

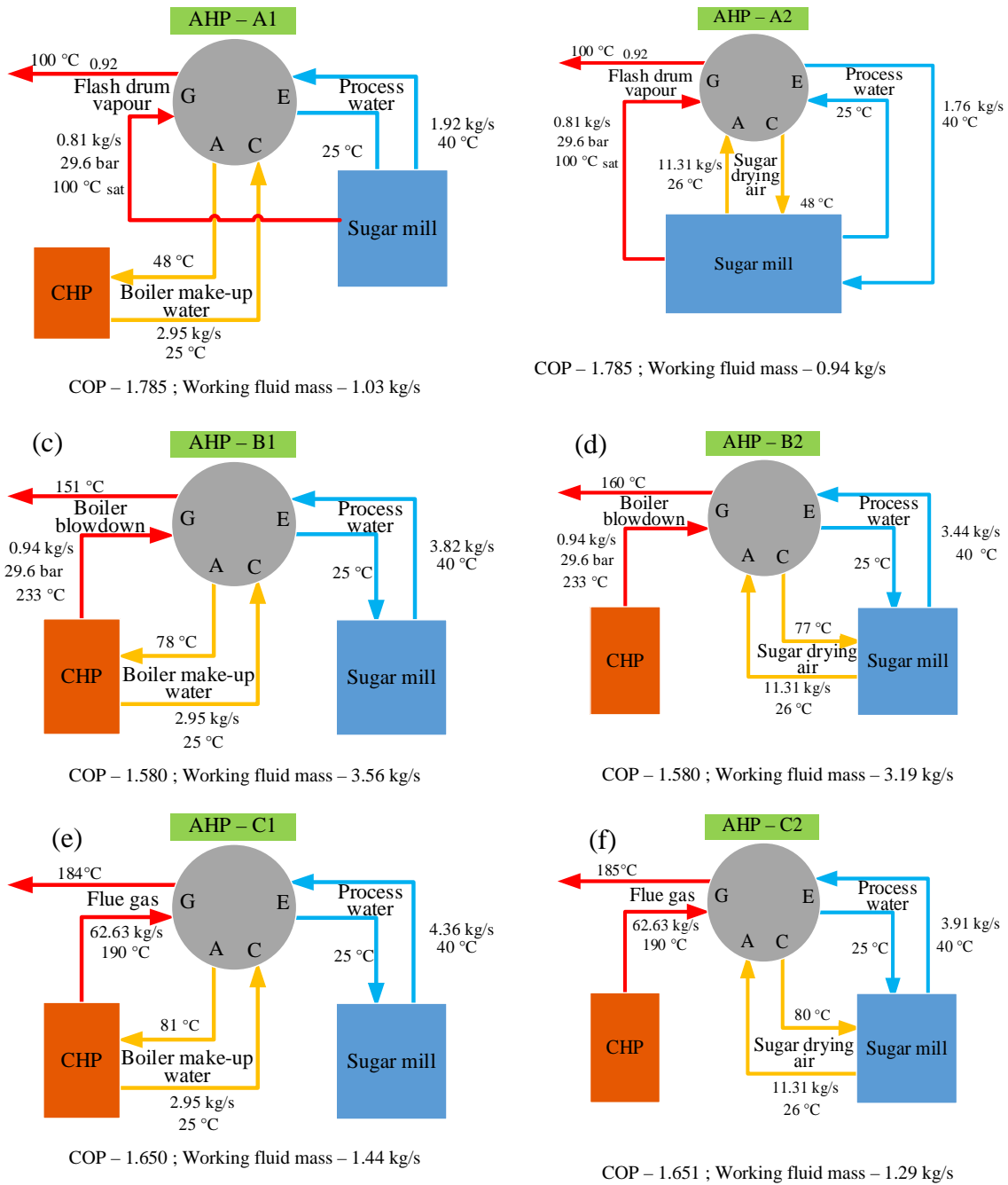


Figure 4-6: Simulation results of AHP integration scenarios (AHP components G – generator; E - evaporator; A – absorber; C – condenser)

The flue gas and boiler blowdown scenarios C-1 and B-1 upgraded the make-up water temperature to 81 °C and 78 °C, and provided 4.36 kg/s and 3.82 kg/s cooling water (about 0.5 % of total cooling water requirement), respectively. These values are almost double what the flash drum vapour scenario A-1 provided due to its relatively low heat quality.

However, scenario B-1 had the lowest coefficient of performance (COP) of 1.58, respectively and the highest working fluid flow of 3.56 kg/s, which could indicate bigger equipment size and consequently, high equipment cost. This could be due to a lower heat transfer rate in the boiler blowdown than in the flash vapour and flue gas, requiring a higher working fluid flow rate to retrieve maximum amount of heat from the blowdown. Therefore, only the flue gas scenario seemed technically attractive by all the criteria described.

Scenarios A-2, B-2 and C-2 with sugar drying air as heat sink showed a similar trend. Generally, the boiler make-up water provided slightly higher heat upgrade and cooling capacity for each heat source than for sugar drying air. Conversely, the COPs for each heat source remained almost the same, regardless of the type of heat sink used. This is because COP of an AHP depends on the heat recovered from the absorber and condenser per heat input (Oluleye et al., 2016), rather than the rate of heat transfer to the heat sinks. Table A.4-9 and Table A.4-10 of the Appendix presents further results from the Aspen simulations of the AHP scenarios.

Temperature upgrade of boiler make-up water reduces the mixing irreversibility of the deaerator and the cogeneration system as a whole, whereas pre-heating sugar drying air reduces the indirect heat transfer irreversibility in the air heater of the drying unit. Consequently, the heating steam requirements for the deaerator and the air heater are minimised resulting in bagasse savings for further valorisation. The AHP integrations also provide an alternative cooling, reducing the cooling tower load and hence, minimizing its irreversibility (Chapter 3). Figure 4-7 presents the exergy improvements and bagasse savings for all scenarios of AHP integrated sugar mill expressed in percentages. Here again, scenario C-1 achieved the highest reduction of 0.14 % in the irreversibility of the overall plant and consequently saved 0.42 ton/h bagasse, i.e., 0.83 % savings on the bagasse burnt to supply energy to the mill.

It is important to note that even though the percentage improvements relative to the entire plant were minimal due to the constraints of the available heat sinks on the extent of heat recovery, the impact on the individual units was significant. The irreversibilities of the deaerator and the sugar dryer were reduced by 51 % and 76 % in scenarios AHP-C1 and AHP-C2, respectively. Therefore, this study could be explored for sugar mills and other

process plants with significant low-temperature heating capacity in order to recover maximum waste heat from the flue gas, cooling it to a temperature as low 90 °C (Rein, 2007). Economic analysis was conducted for the best scenario (AHP-C1) to determine the profitability of integrating AHP in the sugar mill.

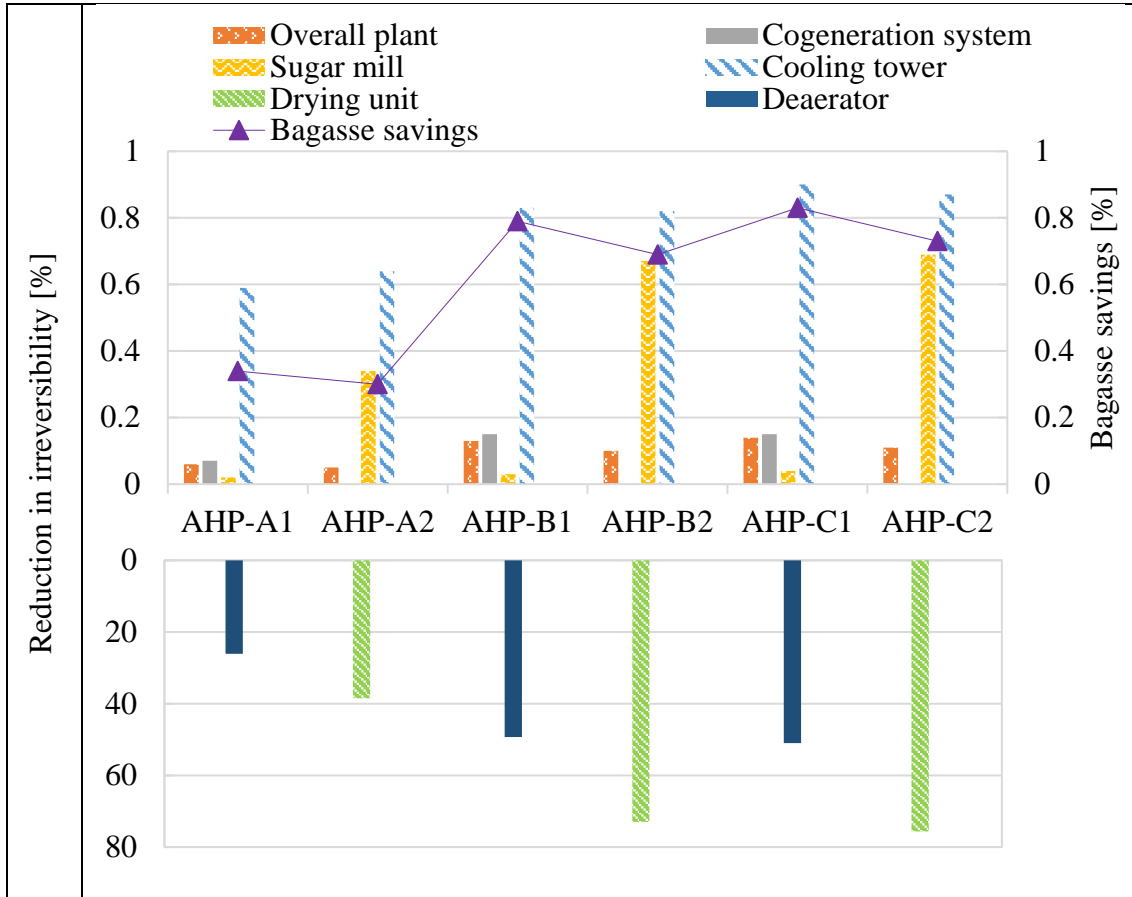


Figure 4-7: Exergy improvement and bagasse savings in AHP-integrated systems (See Table A.4-11 in the Appendix for the data table of this graph)

4.4.3.2. Organic Rankine cycle (ORC)

Two scenarios of ORC integration were simulated for the cogeneration system using the boiler blowdown (B) and the flue gas (C) as waste-heat sources for the ORC, in addition to one for the main sugar process using flash drum vapour (A) as a heat source, making a total of three simulations. Table 4-3 presents the simulation results including the waste heat recovered, the exit temperature of the waste stream and the electricity produced, as well as the evaluated thermal efficiency of each scenario. Of the three scenarios, the C-ORC produced the highest renewable electricity of 932 kW, releasing the flue gas to the environment at 108.5 °C with the process constraints selected. Even though both C-ORC

and B-ORC have the same thermal efficiency of 15.43 % because of their comparable quality of heat sources, the latter produced almost seven times less power than the former due to the low flow rate of boiler blowdown compared to flue gas. On the other hand, the A-ORC had a low thermal efficiency of 10.13 % because of a lower exergy heat source converted to electricity. Besides, the power capacity of the A-ORC (76 kW) was among the smallest commercial ORC capacities reported (Oluleye et al., 2016; Quoilin et al., 2013) and as a result, expected to have a high cost per unit kW (Tartière & Astolfi, 2017). Moreover, it was noted in the literature that 100 °C is the lowest applicable limit for ORC, as such, the cycle operating at this temperature cannot work efficiently (Frate, Ferrari, & Desideri, 2019). Therefore, the A-ORC scenario was not attractive technically and economically, hence no further analysis was conducted on it.

Table 4-3: Results of ORC integrations into a sugar plant

Parameter	A-ORC	B-ORC	C-ORC
Temperature of heat source (6) ^a [°C]	100.0	233	190
Exit temperature of heat source (6) ^a [°C]	100.0	62.1	108.5
Waste heat recovered [kW]	750	831	6042
Net power produced [kW]	76	128	932
Thermal efficiency of ORC [%]	10.13	15.43	15.43

A-ORC, B-ORC and C-ORC represent the organic Rankine cycle powered by flash drum vapour, boiler blowdown and flue gases, respectively.

^a Refers to the state point in Figure 4-3.

The exergy analysis of the ORC- integrated systems showed that irreversibilities of B-ORC (92.40 MW) and C-ORC (93.02 MW) integrated systems were higher than that of the base system (90.58 MW). This resulted from additional irreversibilities generated in the components of the ORC system. On the other hand, the additional electricity produced from the waste heat increased the desired output exergy of the integrated system, while the input resource exergy remained the same. Hence, the systems retrofitted with the B-ORC and the C-ORC had slightly higher exergetic efficiencies of 23.7 % and 24.0 %, respectively, compared to 23.6 % for the base cogeneration system.

4.4.4. Economic feasibility and sensibility analyses of ORC and AHP

The economic feasibility of each of the heat recovery technologies was determined based on the economic evaluation of the best scenarios of AHP and ORC integration in terms

of their waste-heat recovery potential and improvements to the system exergy performance. Table 4-4 presents the economic analysis outcome of the AHP fuelled by flue gas (C-1) and the two attractive ORC scenarios, C-ORC and B-ORC. With an annual net cash flow of US\$ 0.15 million obtained from bagasse savings and free cooling water provided, the AHP technology is profitable with an IRR of 14.1 % and a payback period of 8.3 years.

Both ORC scenarios are profitable with IRRs of 18.7 % (C-ORC) and 11.5 % (B-ORC), relative to a standard minimum IRR of 10 %. The capital cost of the C-ORC (US\$ 2 570/kW) compares well with US\$ 1 500 - US\$ 3 500/kW reported as the general capital cost of ORC (Reddy, Naidu, & Rangaiah, 2013), while that of the B-ORC (US\$ 4 770/kW) is rather high due to the economy of scale. The payback periods of 6.3 years (for C-ORC) and 9.7 years (for B-ORC) are typical of heat recovery projects which are characterized by long payback periods (Thekdi & Nimbalkar, 2014).

Table 4-4: Economic results of attractive AHP and ORC scenarios

	Units	AHP-C1	B-ORC	C-ORC
Waste thermal power to ORC	kW	-	831	6042
Net ORC electricity produced	kW	-	128	932
Net yearly electricity production	MWh/yr	-	641	4662
Net yearly bagasse saved	t/yr	2098	-	-
Yearly cooling rate provided	GJ/yr	4920	-	-
Total capital investment (TCI)	US\$	450 000	611 000	2 386 000
Capital cost	US\$/kW	-	4 770	2 570
Net cash flow	US\$/yr	42 100	46 000	312 500
Internal rate of return (IRR) @25 year operation	%	14.1	11.5	18.7
Payback period	year	8.3	9.7	6.3

Figure 4-8 shows the sensitivity of IRR to ± 25 % changes in fixed capital investment (embedded in the total capital investment), the income tax rate and the prices of electricity (for ORC scenarios) and bagasse (for AHP scenario), as well as ± 5 year change in the

plant life. As seen from Figure 4-8 (a) AHP C-1, the total capital cost and bagasse price have a strong impact on the IRR, however, should the actual AHP cost from the technology vendors be 25 % more than the estimated cost, the technology would still be economically feasible. On the other hand, the technology will be more profitable in the future considering that bagasse is expected to gain higher-value (as high as US\$ 90.00/ton) as a biorefinery feedstock (Petersen et al., 2018), than the current minimum based on its heating value. Possible variations in the income tax rate and plant life based on depreciation do not have a significant influence on the economic performance of the AHP technology.

Similar to the AHP sensitivity results, the TCI has a significant effect on the ORC scenarios (Figure 4-8 (b) C-ORC and (c) B-ORC) and cost estimates as close as possible to the actual technology is of importance. In the same vein, the economic performance of the ORC strongly depended on the price of electricity generated and as a result, the technology will be more attractive for investment when renewable electricity sales price increase. The B-ORC would be economically unprofitable with 25 % unfavourable variation in the TCI and electricity price and would experience higher impact than the C-ORC if the plant equipment depreciated faster (resulting in shorter plant life) than estimated, due to its low profitability.

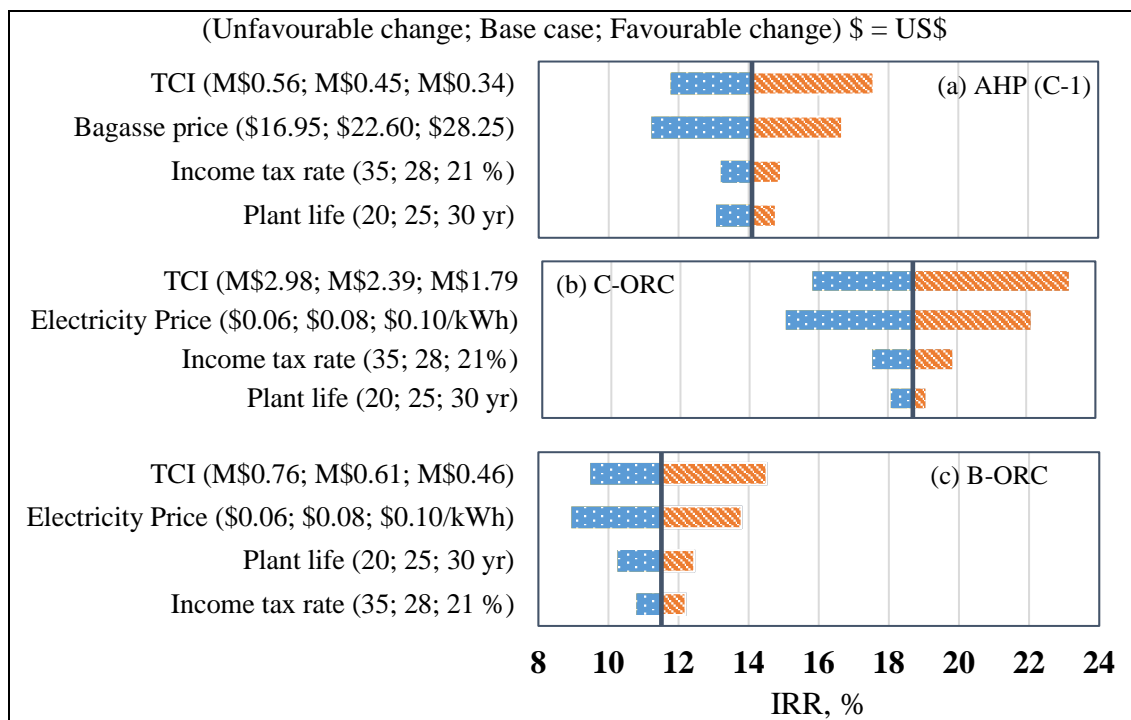


Figure 4-8: Economic sensitivity analysis of AHP and ORC technologies for sugar industry application

Conclusions

The exergy analysis of a sugar-mill cogeneration system and improvements on the exergy performance through the integration of absorption heat pump (AHP) and organic Rankine cycle (ORC) were presented. The boiler efficiency, exergetic efficiency and irreversibility of the base cogeneration system were 65 %, 23.6 % and 6447 kJ/kg of bagasse combusted, respectively. The AHP integration reduced the irreversibility of the overall system, the deaerator and the sugar-drying units by 0.14 %, 51 % and 76 %, respectively, while providing free cooling water for the mill. The ORC improved the cogeneration exergy efficiency by 1.7%, providing additional “green” electricity for the mill. The AHP and ORC integrations were economically feasible with capital investments of US\$ 0.45 million and US\$ 2.39 million, IRR of 14.1 % and 18.7 %, and payback periods of 8.3 years and 6.3 years, respectively. The availability of heat sinks for the AHP is important for maximizing heat recovery and increasing the exergy improvement.

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References

- ABMA. (2011). *Deaerator white paper for use with industrial/commercial and institutional boilers*. Vienna, Virginia: American Boiler Manufacturers Association.
- Aghbashlo, M., Mandegari, M., Tabatabaei, M., Farzad, S., Mojarab Soufiyan, M., & Görgens, J. F. (2018). Exergy analysis of a lignocellulosic-based biorefinery annexed to a sugarcane mill for simultaneous lactic acid and electricity production. *Energy*, *149*, 623–638. <http://doi.org/10.1016/j.energy.2018.02.063>
- Ahmed, A., Esmail, K. K., Irfan, M. A., & Al-Mufadi, F. A. (2018). Design methodology of organic Rankine cycle for waste heat recovery in cement plants. *Applied Thermal Engineering*, *129*, 421–430. <http://doi.org/10.1016/j.applthermaleng.2017.10.019>
- Aspentech. (2013). Aspen Physical Property System. Burlington, MA: Aspen Technology, Inc. Retrieved from <http://www.aspentech.com>
- Balaji, K., & Ramkumar, R. (2012). Study of Waste Heat Recovery From Steam Turbine Exhaust for Vapour Absorption System in Sugar Industry. *Procedia Engineering*, *38*, 1352–1356. <http://doi.org/10.1016/j.proeng.2012.06.167>
- Chan, C. W., Ling-Chin, J., & Roskilly, A. P. (2013). A review of chemical heat pumps, thermodynamic cycles and thermal energy storage technologies for low grade heat utilisation. *Applied Thermal Engineering*, *50*(1), 1257–1273. <http://doi.org/10.1016/j.applthermaleng.2012.06.041>
- Chouhan, P. (2014). Performance enhancement of sugar mill by alternate cooling system for condenser. *IOSR Journal of Mechanical and Civil Engineering (IOSR-JMCE)*,

11(5), 18–25. Retrieved from <http://www.iosrjournals.org/iosr-jmce/papers/vol11-issue5/Version-1/C011511825.pdf>

- Colorado, D., Demesa, N., Huicochea, A., & Hernández, J. A. (2016). Irreversibility analysis of the absorption heat transformer coupled to a single effect evaporation process. *Applied Thermal Engineering*, *92*, 71–80. <http://doi.org/10.1016/j.applthermaleng.2015.09.076>
- Deshmukh, R., Jacobson, A., Chamberlin, C., & Kammen, D. (2013). Thermal gasification or direct combustion? Comparison of advanced cogeneration systems in the sugarcane industry. *Biomass and Bioenergy*, *55*, 163–174. <http://doi.org/10.1016/j.biombioe.2013.01.033>
- Eisa, M. A. R., Devotta, S., & Holland, F. A. (1986). Thermodynamic design data for absorption heat pump systems operating on water-lithium bromide: Part III-Simultaneous cooling and heating. *Applied Energy*, *25*, 83–96. [http://doi.org/10.1016/0890-4332\(95\)90067-5](http://doi.org/10.1016/0890-4332(95)90067-5)
- Ensinas, A. V., Modesto, M., Nebra, S. A., & Serra, L. (2009). Reduction of irreversibility generation in sugar and ethanol production from sugarcane. *Energy*, *34*(5), 680–688. <http://doi.org/10.1016/j.energy.2008.06.001>
- FAOSTAT. (n.d.). Retrieved July 27, 2018, from <http://www.fao.org/faostat/en/#data>
- Ferrer, S., Mezquita, A., Aguilera, V. M., & Monfort, E. (2019). Beyond the energy balance: Exergy analysis of an industrial roller kiln firing porcelain tiles. *Applied Thermal Engineering*, *150*, 1002–1015. <http://doi.org/10.1016/j.applthermaleng.2019.01.052>
- Fontalvo Alzate, J. (2014). Using user models in Matlab® within the Aspen Plus® interface with an Excel® link. *Ingeniería e Investigación*, *34*(2), 39–43. <http://doi.org/10.15446/ing.investig.v34n2.41621>
- Frate, G. F., Ferrari, L., & Desideri, U. (2019). Analysis of suitability ranges of high temperature heat pump working fluids. *Applied Thermal Engineering*, *150*, 628–640. <http://doi.org/10.1016/j.applthermaleng.2019.01.034>
- Ghannadzadeh, A., & Sadeqzadeh, M. (2017). Exergy aided pinch analysis to enhance

- energy integration towards environmental sustainability in a chlorine-caustic soda production process. *Applied Thermal Engineering*, 125, 1518–1529. <http://doi.org/10.1016/j.applthermaleng.2017.07.052>
- Hector, I. V. A., Carolina Mesa, M., & Sergio, A. G. A. (2013). Energy and Exergy Analysis of the Combined Production Process of Sugar and Ethanol from Sugarcane (A Colombian Case Study). In *Proceedings of the 26th International Conference on Efficiency, Cost, Optimization, Simulation and Environmental Impact of Energy Systems* (pp. 87–96). Retrieved from <http://www.scopus.com/inward/record.url?eid=2-s2.0-77952919514&partnerID=40&md5=5b02e7a81b7bb5d72d770abaa34c2a55>
- Hess, S., Beukes, H., Smith, G., & Dinter, F. (2016). Initial study on solar process heat for South African sugar mills. In *Proc S Afr Sug Technol Ass* (pp. 324–349). Durban, South Africa.
- International Sugar Organization. (2016). About sugar. Retrieved August 28, 2017, from <http://www.isosugar.org>
- Kamate, S. C., & Gangavati, P. B. (2009). Exergy analysis of cogeneration power plants in sugar industries. *Applied Thermal Engineering*, 29(5–6), 1187–1194. <http://doi.org/10.1016/j.applthermaleng.2008.06.016>
- Kotas, T. J. (1995). *The Exergy Method of Thermal Plant Analysis* (Reprint Ed). Florida, USA: Krieger Publishing Company, Malabar. <http://doi.org/10.1016/B978-0-408-01350-5.50003-9>
- Kurem, E., & Horuz, I. (2001). A comparison between ammonia-water and water-lithium bromide solutions in absorption heat transformers. *International Communications in Heat and Mass Transfer*, 28(3), 427–438. [http://doi.org/10.1016/S0735-1933\(01\)00247-0](http://doi.org/10.1016/S0735-1933(01)00247-0)
- Lansing, F. L. (1976). Computer modeling of a single-stage lithium bromide/water absorption refrigeration unit. *The Interplanetary Network Progress Report*. Retrieved from http://ipnpr.jpl.nasa.gov/progress_report2/42-32/32AA.PDF
- Lemmens, S. (2016). Cost engineering techniques and their applicability for cost

- estimation of organic rankine cycle systems. *Energies*, 9(485), 1–18. <http://doi.org/10.3390/en9070485>
- Mahmoudi, A., Fazli, M., & Morad, M. R. (2018). A recent review of waste heat recovery by Organic Rankine Cycle. *Applied Thermal Engineering*, 143, 660–675. <http://doi.org/10.1016/j.applthermaleng.2018.07.136>
- Mandegari, M. A., Farzad, S., & Görgens, J. F. (2017). Recent trends on techno-economic assessment (TEA) of sugarcane biorefineries. *Biofuel Research Journal*, 4(3), 704–712. <http://doi.org/10.18331/BRJ2017.4.3.7>
- Mkwananzi, T., Mandegari, M., & Görgens, J. F. (2019). Disturbance modelling through steady-state value deviations: The determination of suitable energy indicators and parameters for energy consumption monitoring in a typical sugar mill. *Energy*, 176, 211–223. <http://doi.org/10.1016/j.energy.2019.03.191>
- Nsaful, F., Görgens, J. F., & Knoetze, J. H. (2013). Comparison of combustion and pyrolysis for energy generation in a sugarcane mill. *Energy Conversion and Management*, 74, 524–534. <http://doi.org/10.1016/j.enconman.2013.07.024>
- Oluleye, G., Jobson, M., Smith, R., & Perry, S. J. (2016). Evaluating the potential of process sites for waste heat recovery. *Applied Energy*, 161, 627–646. <http://doi.org/10.1016/j.apenergy.2015.07.011>
- Palacios-Bereche, R., & Nebra, S. A. (2009). Thermodynamic modeling of a cogeneration system for a sugarcane mill using Aspen-Plus, difficulties and challenges. In *20th International Congress of Mechanical Engineering* (pp. 1–9). Gramado, RS, Brazil.
- Pellegrini, L. F., de Oliveira Jr, S., & Burbano, J. C. (2010). Supercritical steam cycles and biomass integrated gasification combined cycles for sugarcane mills. *Energy*, 35(2), 1172–1180. <http://doi.org/10.1016/j.energy.2009.06.011>
- Perry, R., & Green, D. (1965). Thermodynamic Properties. In *Perry's Chemical Engineer's Handbook* (7th Editio, pp. 1869–1875). New York.
- Petersen, A. M., Van der Westhuizen, W. A., Mandegari, M. A., & Görgens, J. F. (2018). Economic analysis of bioethanol and electricity production from sugarcane in South Africa. *Biofuels, Bioproducts and Biorefining*, 12(2), 224–238.

<http://doi.org/10.1002/bbb.1833>

- Querol, E., Gonzalez-Regueral, B., & Perez-Benedito, J. L. (2013). Exergy Concept and Determination. In *Practical Approach to Exergy and Thermo-economic Analyses of Industrial Processes* (pp. 9–28). SpringerBriefs in Energy. http://doi.org/10.1007/978-1-4471-4622-3_2
- Quoilin, S., Broek, M. Van Den, Declaye, S., Dewallef, P., & Lemort, V. (2013). Techno-economic survey of organic rankine cycle (ORC) systems. *Renewable and Sustainable Energy Reviews*, 22, 168–186. <http://doi.org/10.1016/j.rser.2013.01.028>
- Reddy, C. C. S., Naidu, S. V., & Rangaiah, G. P. (2013). Waste Heat Recovery Methods And Technologies. *Chemical Engineering (United States)*, 120(1), 28–38.
- Reid, B. M. J., & Rein, P. W. (1983). Steam balance for the new Felixton II mill. In *Proceedings of The South African Sugar Technologists' Association* (pp. 85–91).
- Reid, M. J. (2006). Why do we continue to burn so much coal? In *Proceedings of The South African Sugar Technologists' Association* (Vol. 80, pp. 353–363).
- Rein, P. (2007). *Cane Sugar Engineering*. (A. Bartens, Ed.). Berlin.
- Seebaluck, V., Mohee, R., Sobhanbabu, P. R. K., Leal, M. R. L. V., & Johnson, F. X. (2008). *Bioenergy for sustainable development and global competitiveness: The case of sugar cane in Southern Africa*. Stockholm, Sweden.
- Somers, C., Mortazavi, A., Hwang, Y., Radermacher, R., Rodgers, P., & Al-Hashimi, S. (2011). Modeling water/lithium bromide absorption chillers in ASPEN Plus. *Applied Energy*, 88(11), 4197–4205. <http://doi.org/10.1016/j.apenergy.2011.05.018>
- Tartière, T., & Astolfi, M. (2017). A World Overview of the Organic Rankine Cycle Market. *Energy Procedia*, 129, 2–9. <http://doi.org/10.1016/j.egypro.2017.09.159>
- Thekdi, A., & Nimbalkar, S. U. (2014). *Industrial waste heat recovery - potential applications, available technologies and crosscutting R&D opportunities*. Tennessee. Retrieved from <http://www.osti.gov/scitech/biblio/1185778-industrial-waste-heat-recovery-potential-applications-available-technologies-crosscutting-opportunities>

- USDA. (2018). Sugar: World Markets and Trade. Retrieved June 5, 2018, from <https://public.govdelivery.com/accounts/USDAFAS/subscriber/new>.
- Utlu, Z., Sogut, Z., Hepbasli, A., & Oktay, Z. (2006). Energy and exergy analyses of a raw mill in a cement production. *Applied Thermal Engineering*, 26(17–18), 2479–2489. <http://doi.org/10.1016/j.applthermaleng.2005.11.016>
- Uusitalo, A., Honkatukia, J., Turunen-Saaresti, T., & Grönman, A. (2018). Thermodynamic evaluation on the effect of working fluid type and fluids critical properties on design and performance of Organic Rankine Cycles. *Journal of Cleaner Production*, 188, 253–263. <http://doi.org/10.1016/j.jclepro.2018.03.228>
- Wienese, A. (2001). Boilers, boiler fuel and boiler efficiency. In *Proc S Afr Sug Technol Ass* (pp. 275–281). Durban, South Africa.
- Xi, H., Li, M., He, Y., & Tao, W. (2015). A graphical criterion for working fluid selection and thermodynamic system comparison in waste heat recovery. *Applied Thermal Engineering*, 89, 772–782. <http://doi.org/10.1016/j.applthermaleng.2015.06.050>
- Zhang, C., Yang, M., Lu, M., Shan, Y., & Zhu, J. (2011). Experimental research on LiBr refrigeration – Heat pump system applied in CCHP system. *Applied Thermal Engineering*, 31(17–18), 3706–3712. <http://doi.org/10.1016/j.applthermaleng.2011.02.004>

Appendix

Table A.4-1: Data for simulating the cogeneration system of a typical sugar mill

Parameter	Unit	Value	Reference
Steam to cane ratio	[%]	45	[a]
Feed water pump discharge pressure	[bar]	36	(Rein, 2007)
Deaerator steam vented	[kg/s]	0.03	(b)
Boiler blowdown (% of steam rate)	[%]	3	(a)
Feed water temperature difference in deaerator	[°C]	90 - 105	(a)
Feed water pump isentropic efficiency	[%]	75	(Rein, 2007)
Turbine isentropic efficiency	[%]	70	(b)
Turbine mechanical efficiency	[%]	98	(Nsaful et al., 2013; Rein, 2007)
Bagasse moisture	[%]	50.84	(Dogbe et al., 2018)
Excess air	[%]	25	(Wienese, 2001)
Furnace temperature	[°C]	950	(Kamate & Gangavati, 2009b; Wienese, 2001)

[a] Data provided by the South African Sugar Milling Research Institute NPC (SMRI)

[b] CleaverBrooks. (1998). Deaerator and Surge controls. U.S.A. Retrieved from <http://www.cleaverbrooks.com>

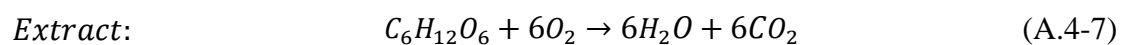
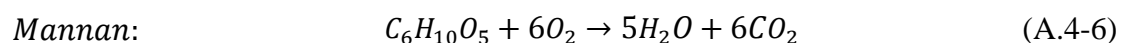
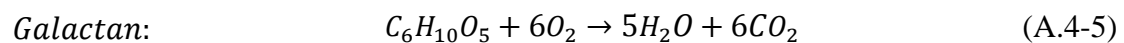
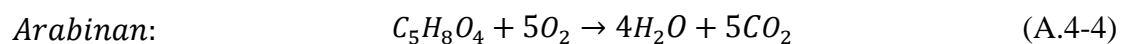
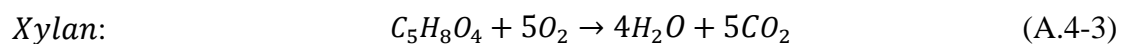
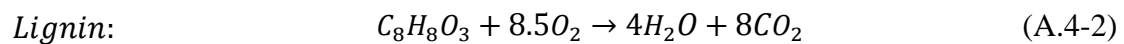
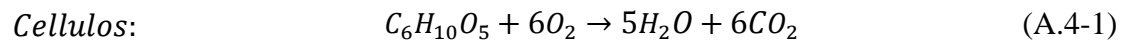
[c] Data provided by RCL FOODS, South Africa

Table A4-2: Bagasse Composition for boiler simulation

Bagasse component (dry basis)	Component name in Aspen	Type	Formula	Composition (% by mass) (Nsaful et al., 2013)
CELLULOSE	DILACTIC-ACID ^a	CISOLID	C ₆ H ₁₀ O ₅	40.69
LIGNIN	VANILLIN	CISOLID	C ₈ H ₈ O ₃	25.53
XYLAN	GLUTARIC-ACID ^a	CISOLID	C ₅ H ₈ O ₄	19.97
ARABINAN	GLUTARIC-ACID ^a	CISOLID	C ₅ H ₈ O ₄	1.65
GALACTAN	DILACTIC-ACID ^a	CISOLID	C ₆ H ₁₀ O ₅	0.75
MANNAN	DILACTIC-ACID ^a	CISOLID	C ₆ H ₁₀ O ₅	0.15
EXTRACT	D-FRUCTOSE	MIXED	C ₆ H ₁₂ O ₆	7.51
ASH	CALCIUM-OXIDE	CISOLID	CaO	3.75
Moisture	WATER	MIXED	H ₂ O	50.84

^a The assumed component names are based on NREL (National Renewable Energy Laboratory) technical report (Humbird et al., 2011) [Humbird, D., Davis, R., Tao, L., Kinchin, C., Hsu, D., & Aden, A. (2011). *Process Design and Economics for Biochemical Conversion of Lignocellulosic Biomass to Ethanol. Technical Report NREL/TP-5100-47764*. Golden, Colorado.]

The combustion reactions in equations (A.4-1) to (A.4-7) led to the evaluation of the stoichiometric air required for bagasse combustion.



Equation (4-8) expresses the stoichiometry amount of oxygen needed for combustion of bagasse.

$$\dot{m}_{oxygen} = MW_{oxygen} \left(\sum_{component} \left(\frac{a}{b} \times \dot{n} \right) \right) \quad (\text{A.4-8})$$

Equation (A.4-9) defines the theoretical air.

$$\dot{m}_{air} = \frac{\dot{m}_{oxygen}}{0.228} \quad (\text{A.4-9})$$

Equation (A.4-10) evaluates the mass flow of combustion air considering 25 % excess air used.

$$\dot{m}_{combustion.air} = 1.25 \times \dot{m}_{air} \quad (\text{A.4-10})$$

Table A.4-3: Specific chemical exergy of chemical components in the cogeneration system

Component	Specific chemical exergy [kJ/mol]	Component	Specific chemical exergy [kJ/mol]
Arabinan (C ₅ H ₈ O ₄)	2281.65	Extract (C ₆ H ₁₂ O ₆)	2779.83
Calcium oxide (CaO)	110.16	Mannan (C ₆ H ₁₀ O ₅)	2745.07
Carbon dioxide (CO ₂)	19.87	Nitrogen (N ₂)	0.72
Cellulose (C ₆ H ₁₀ O ₅)	2745.07	Oxygen (O ₂)	3.97
Galactan (C ₆ H ₁₀ O ₅)	2745.07	Water (H ₂ O)	0.9
Lignin (C ₈ H ₈ O ₃)	4284.69	Xylan (C ₅ H ₈ O ₄)	2281.65

References:

1. Sohel, M. I., & Jack, M. W. (2011). Thermodynamic analysis of lignocellulosic biofuel production via a biochemical process : Guiding technology selection and research focus. *Bioresource Technology*, 102(3), 2617–2622. <http://doi.org/10.1016/j.biortech.2010.10.032>
2. Wall, G. (2009). *Exergetics*. Bucaramanga: Exergy, Ecology, Democracy.

Table A.4-4: Operating parameters for the simulation of absorption heat pump

Parameter	Waste-heat source		
	Flash drum vapour	Boiler blowdown	Flue gas
Absorber temperature [°C]	50	60	60
Generator temperature [°C]	90	140	170
Condenser temperature [°C]	50	85	100
Evaporator temperature [°C]	20	20	20
LiBr mass fraction in weak solution, x_1	0.5410	0.5915	0.5915
LiBr mass fraction in strong solution, x_2	0.5696	0.6104	0.6495
Evaporator (lower) pressure [kPa]	2.339	2.339	2.339
Condenser (higher) pressure [kPa]	12.351	57.867	101.418

Table A.4-5: Method of determining total capital investment for ORC and AHP

Components of total capital investment (TCI)	Calculation factor
Fixed-capital investment (FCI)	
<i>(a) Direct fixed - capital investment (DFCI)</i>	
Purchased equipment cost (PEC)	Derived from Aspen
Purchased equipment installation	47 % of PEC
Piping	66 % of PEC
Instrumentation and controls	18 % of PEC
Electrical equipment and materials	11 % of PEC
Service facilities	70 % of PEC
Civil, structural and architectural work	44 % of PEC
<i>(b) Indirect fixed - capital investment (IFCI)</i>	
Project contingencies	42 % of PEC
Construction costs including contractor's profit	62 % of PEC
Engineering and supervision	33 % of PEC
Other costs	
Working capital (Start-up costs)	5 % of FCI

Table A.4-6: ORC major equipment costing retrieved from Aspen Economic Analyser

Equipment	Cost [US\$]	
	WBB ORC	Flue gas ORC
Pump	55000	42800
Evaporator	75700	13700
Turbine	164300	45700
Condenser	211100	22400
Total purchased equipment cost (PEC)	506100	124600

Table A.4-7: AHP major equipment costing retrieved from Aspen Economic Analyser

Equipment	Cost [US\$]
Evaporator	14900
Absorber	19800
Solution pump	3800
Solution heat exchanger	10900
Generator	31290
Condenser	11000
Total purchased equipment cost (PEC)	91690

Table A.4-8: Input parameters for economic analysis of ORC and AHP technologies

Economic Parameter	Value	Reference
Bagasse price	US\$ 22.60 /ton ^a	[a]
Electricity price	US\$ 0.08 /kWh	[b]
Cooling water price	US\$ 0.22 /GJ	[Aspen simulation]
Yearly inflation rate	6.87 %	[c]
Income tax rate	28 %/year	[b]
Economic life of plant	25 years	[b]
Operating hours per year	5000 ^b	[d], [e]
Minimum IRR for real term	10 %	[b]

[a] Currently, bagasse is burnt as a fuel, therefore, its minimum value is based on the heating value. Therefore, the price of bagasse was estimated based on the unit price of coal and the heating value of bagasse with respect to coal. Coal was delivered at 1100 ZAR/t (Hess et al., 2016) in 2015 which is equivalent to US\$ 90.50 /ton in 2016 based on 12.3 % price increase per annum (Hess et al., 2016) and an

exchange rate of 13.65 ZAR/1USD as at 31 December 2016. The equivalent ratio of coal to bagasse consumed per unit steam produced is 1:4 units (Smith, Davis, Madho, & Achary, 2015)

[b] Mandegari, M., Farzad, S., & Görgens, J. F. (2018). A new insight into sugarcane biorefineries with fossil fuel co-combustion: Techno-economic analysis and life cycle assessment. *Energy Conversion and Management*, *165*, 76–91. <http://doi.org/10.1016/j.enconman.2018.03.057>.

[c] Inflation.eu. (2016). World inflation data. Retrieved December 19, 2016, from <http://www.inflation.eu/inflation-rates/south-africa/current-cpi-inflation-south-africa.aspx>

[d] Based on an average South African sugar industry milling season of 255 days and time efficiency of 81.44 % [e], the effective annual operating time for the economic analysis was calculated to be approximately 5000 hours.

[e] Smith, G., Davis, S., Madho, S., & Achary, M. (2015). Ninetieth Annual Review of the Milling Season in Southern Africa (2014/2015). In *Proceedings South African Sugar Technology Association* (pp. 23–54). Retrieved from <http://www.sasta.co.za/wp-content/uploads/2011/12/2011-Annual-Review-Factory.pdf>.

Table A.4-9: Mass flow of working fluid in the absorption heat pump scenarios from the Aspen simulations

State point	Description	Mass flow rate [kg/s]					
		A-1	A-2	B-1	B-2	C-1	C-2
1, 2, 3	LiBr weak solution	1.028	0.944	3.556	3.194	1.444	1.294
4, 5, 6	LiBr strong solution	0.976	0.897	3.446	3.096	1.315	1.179
7, 8, 9, 10	Refrigerant flow	0.052	0.047	0.110	0.099	0.129	0.116

Table A.4-10: Aspen simulation results the absorption heat pump scenarios

State point	Temperature			Pressure			LiBr concentration		
	[°C]			[kPa]			[%]		
	A-1/2	B-1/2	C-1/2	A-1/2	B-1/2	C-1/2	A-1/2	B-1/2	C-1/2
1	48.9	58.0	58.0	2.34	2.339	2.339	54.10	59.15	59.15
2	48.9	58.0	58.1	12.35	57.87	101.42	54.10	59.15	59.15
3	78.5	121.5	139.0	12.35	57.87	101.42	54.10	59.15	59.15
4	88.9	140.0	170.0	12.35	57.87	101.42	56.96	61.04	64.95
5	56.9	74.4	80.4	12.35	57.87	101.42	56.96	61.04	64.95
6	54.1	62.9	71.7	2.34	2.34	2.34	56.96	61.04	64.95
7	83.0	135.0	153.0	12.35	57.87	101.42	0.00	0.00	0.00
8	50.0	85.0	100.0	12.35	57.87	101.42	0.00	0.00	0.00
9	20.0	20.0	20.0	2.34	2.34	2.34	0.00	0.00	0.00
10	20.0	20.0	20.0	2.34	2.34	2.34	0.00	0.00	0.00

Table A.4-11: Bagasse savings and irreversibility reductions in a sugar mill by the integration of AHP

Improvements gained	AHP integration scenarios					
	AHP-A1	AHP-A2	AHP-B1	AHP-B2	AHP-C1	AHP-C2
Bagasse savings [%]	0.34	0.30	0.79	0.69	0.83	0.73
<i>Total reduction in irreversibility [%]</i>						
Overall plant	0.06	0.05	0.13	0.1	0.14	0.11
Cogeneration system	0.07	-	0.15	-	0.15	-
Sugar mill	0.02	0.34	0.03	0.67	0.04	0.69
<i>Irreversibility reduction in individual units [%]</i>						
Cooling tower	0.59	0.64	0.83	0.82	0.9	0.87
Drying unit	-	38.49	-	72.99	-	75.64
Deaerator	26.04	-	49.28	-	50.99	-

Chapter 5

Towards revitalization of the sugarcane industry through biorefineries valorising A-molasses

Objective of dissertation in this chapter and summary of findings

Chapter 3 of this thesis investigated the locations, magnitudes and causes of energy efficiency in the sugar mill being studied and revealed that the three-stage (A, B and C) sugar crystallization process, especially the B- and C- crystallizations were the most inefficient among the production processes. A single-stage crystallization process was recommended to reduce the process complexity and hence improve the sugar mill exergy efficiency. However, a single-stage sugar crystallization results in lower recovery of crystalline sucrose for sale and a sucrose-rich by-product, A-molasses, which could be valorised into higher-value products through an integrated biorefinery concept.

Therefore, based on these results of the crystallization unit exergy performance of Chapter 3, an integrated biorefinery concept was developed for the valorisation of A-molasses considered as a better design option for enhancing the energy efficiency and the economic sustainability and competitiveness of the sugar industry (the end goal of this project) than maximally producing only crystalline sucrose in an energy inefficient way. This consideration was motivated by dropping world sugar market prices caused by subsidised overproduction by major sugar-producing countries. Moreover, diverting A-molasses to a biorefinery rather than for pure sugar eliminates the cost of further processing to pure sugar in the B- and C-crystallisations, especially due to process complexity, low energy efficiency and exergy destruction in these stages. Furthermore, A-molasses is a purer sugar stream than B- and C-molasses and may be used as a cheaper feedstock than pure sugar for biorefinery products with less purification requirement.

Therefore, the key contribution of this chapter was the concept of valorising A-molasses in an integrated biorefinery based on the single sugar crystallization idea established through the exergy analysis results of the sugar mill and the motivation of falling global sugar market prices. This Chapter evaluated the economic feasibility and profitability of valorising all A-molasses available for the production of succinic acid (SA), short-chain fructooligosaccharides (scFOS) in powder and syrup forms in single- and multi-product biorefinery scenarios, because of their importance and promising markets (Objective 4). Additional scenarios were considered to co-utilise A-molasses with the scFOS process by-product rich in fermentable sugars to produce additional SA in the multi-product scenarios. Seven biorefinery scenarios were developed in all and simulated in Aspen Plus for their energy and mass balances necessary for the economic evaluation.

The economic analysis was conducted considering the effect of production capacity (relative to market sizes) on the market prices of the products. Moreover, due to the energy saved (from B- and C- pans and drying unit) by producing less crystalline sucrose that was available to the biorefinery, it was assumed that the existing cogeneration unit of the sugar mill supplied steam and electricity to the biorefinery by addition available surplus bagasse to produce 10 % more steam. Moreover, based on the findings of Chapter 4 on the exergy benefits and economic feasibility of absorption heat pump technology, absorption chillers were considered for chilled water supply to the biorefinery. It should be noted that the economic assumptions made for the integration though optimistic and could be revised, were uniform bases for comparing all the scenarios developed.

All the developed biorefineries were economically viable with very high internal rates of return (IRRs) due to integration benefits, particularly the sharing of sugar mill cogeneration system eliminating the need for a new boiler in the biorefineries. In essence, the insight given by exergy analysis in Chapter 3 led to the design of profitable biorefineries to enhance the economic sustainability and competitiveness of the sugar mill. Moreover, the scenario that co-produced SA and scFOS syrup using appropriate A-molasses split based on their different market sizes presented the lowest minimum selling prices of the products and was considered the most desirable scenario for easy market entry of the products.

However, because all the scenarios were economically viable, they are suitable for exergoeconomic (exergy-based costing) analysis to determine the preferred one holistically for integration into the sugar mill. Therefore, the exergy concept can be applied to not only improve the efficiency of sugar mills but also as a design tool in adding biorefineries to sugar mills.

Declaration by the candidate

With regards to Chapter 5, pages 162 -195, the nature and scope of my contribution were as follows:

Nature of contribution	Extent of contribution (%)
Project and scope definition, Aspen simulation work, analysis of data, interpretation of results and writing of manuscript	80

The following co-authors have contributed to Chapter 5, pages 162 -195:

Name	e-mail address	Nature of contribution	Extent of contribution (%)
J.F. Görgens	jgorgens@sun.ac.za	Provided writing assistance through review and proof reading of manuscript and general discussion	8
M.A. Mandegari	mandegari@sun.ac.za	Reviewed Aspen simulations, and provided writing assistance through suggestions, review and proof reading of manuscript and general discussions	12

Signature of candidate:

Date:

Declaration by co-authors:

The undersigned hereby confirm that

1. the declaration above accurately reflects the nature and extent of the contributions of the candidate and the co-authors to Chapter 5, pages 162 - 195,
2. no other authors contributed to Chapter 5, pages 162 -195 besides those specified above, and
3. potential conflicts of interest have been revealed to all interested parties and that the necessary arrangements have been made to use the material in Chapter 5, pages 162 -195, of this dissertation.

Signature	Institutional affiliation	Date
	Stellenbosch University	
	Stellenbosch University	

Towards revitalization of the sugarcane industry through biorefineries valorising A-molasses

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Abstract

The survival of the South African sugar industry is uncertain considering the dropping global market price of sugar and rising energy prices. Exergy analysis of a typical South African sugar mill revealed that the B- and C- crystallizations of the current three-stage sugar crystallization process were the most inefficient due to the process complexities. A single-stage sugar crystallization, which produces less crystalline sugar and sucrose-rich A-molasses was recommended to improve the exergy efficiency of the sugar mill and valorisation of the A-molasses was proposed towards extending the economic life of the sugar industry. This work presents a techno-economic assessment of integrated biorefinery annexed an existing sugar mill producing succinic acid (SA) and short-chain fructooligosaccharides (scFOS) from A-molasses at target markets of 440 kt/y and 160 kt/y, respectively. SA and scFOS were selected based on their importance, promising market growth and favourable economics from previous studies. Considering the set of economic assumptions applied, all scenarios presented gave IRRs far above the minimum required of 9.7 %, due to the benefits of integration. Co-production of SA and scFOS syrup using a 70 – 30 % split of the A-molasses at production rates of no more than 10 % of global market sizes (40.5 kt/y and 13.5 kt/y) and market prices of 2500 US\$/t and 471 US\$/t, respectively, provided the lowest minimum selling prices of 1222 US\$/t and 35 US\$/t. By this result, this scenario indicates suitability for easy market entry of the products. The proposed integration can improve energy efficiency and also reduce export-sugar economic losses of the sugar mill.

Keywords: A-molasses; Economic analysis; Integrated biorefinery; Short-chain fructooligosaccharides; Succinic acid; Sugar mill.

5.1. Introduction

The South African sugar industry is in decline due to multiple challenges including drought in cane producing areas, plunging sugar prices in the international market leading to cheap imports and unattractive export prices, and increasing energy prices (Somayeh Farzad, Mandegari, Guo, et al., 2017). To alleviate the effects of these challenges, sugar industries worldwide are shifting from sugar only production towards diversifying their products portfolio, by co-producing valuable products in a biorefinery concept (Martínez-Guido et al., 2016; Renó et al., 2014). This shift is also motivated by a global trend toward replacing fossil-based products bio-based ones, due to the associated environmental benefits and limited reserves of fossil resources (Aghbashlo, Mandegari, et al., 2018b).

Common feedstock studied for sugarcane biorefineries include sugarcane plant residues (trash) and sugar process by-products such as final (C) molasses and bagasse (Martínez-Guido et al., 2016). These are used as cheap feedstocks instead of processed sugar to reduce production cost (Ali Mandegari et al., 2017; Nieder-Heitmann et al., 2019). However, lignocellulose biomass such as bagasse and leaves contain structural carbohydrates, which require expensive pre-treatment and hydrolysis steps to convert them to fermentable sugars for biorefineries, which may supersede the perceived low cost of the lignocelluloses (Nieder-Heitmann et al., 2019). On the other hand, C-molasses contain high amounts of non-sucrose impurities (Starzak & Davis, 2016) and may not be suitable for sucrose-based non-fermentation products like fructooligosaccharides (Eggleston & Lima, 2015). The same could be said of SA production considering the expected purity for platform chemicals. Moreover, Brazilian and other sugarcane biorefineries have demonstrated that using clarified juice and intermediate (A and B) molasses for co-production of sugar and ethanol (M. O. de S. Dias et al., 2015a; Krajnc & Glavič, 2009) have superior economics compared to using C-molasses (Krajnc & Glavič, 2009).

A-molasses is a purer sugar stream than B- and C-molasses and may easily replace pure sugars (especially sucrose) as a cheaper biorefinery feedstock (Krajnc & Glavič, 2009; Starzak & Davis, 2016), while providing biorefinery products with less purification

requirements. The crystallization unit of a typical South African is designed in three (A, B and C) stages to recover the maximum amount of sucrose in the saleable crystal sugar stream (Rein, 2007). Diverting the A-molasses from the A-stage to a biorefinery will therefore decrease the amount of crystalline sucrose produced, similar to diverting the clarified juice, but will also save the amount of energy expended on the last two crystallization stages (B and C) and eliminate the associated exergy destruction. A biorefinery concept would focus on maximising the overall value extracted from sugarcane, rather than maximizing sugar recovery, which is achieved through a combination of high-value products and crystalline sucrose production. Moreover, diverting A-molasses to a biorefinery rather than pure sugar eliminates the cost of further processing to pure sugar in the B- and C-crystallisations, especially due to process complexity, low energy efficiency and exergy destruction in these stages (Chapter 3).

The biorefinery concept presents the possibility of using a common feedstock for multiple products including biochemicals, bioenergy, biofuels and food products (Soltanian et al., 2019). Two biorefinery products that have received research attention with the use of molasses as cheap feedstock are succinic acid (SA) (S. Chan et al., 2012; Y. P. Liu et al., 2008; Ma et al., 2014) and fructooligosaccharides (Shin, 2004; Ur Rehman et al., 2016), because of their promising market growth. Bio-based SA, produced from microbial fermentation of plant-derived sugars (Biddu et al., 2016) is a platform chemical widely used for high-value niche applications in the food and beverage industries, as well as for high volume products such as plasticizers, polyurethanes, resins, coatings and more than thirty commercially important products (Vaswani, 2010). The potential for SA as a platform chemical to replace maleic anhydride for the BDO production is the main driver for the growing global SA market, estimated at 40 kt/y in 2014 and expected to 600 kt/y by 2020 (Klein et al., 2017).

Furthermore, demand for short-chain fructooligosaccharides (scFOS) as functional foods is increasing due to their properties as prebiotics and low calorie sweeteners with health benefits (Dominguez et al., 2014a). These health benefits include prevention of colon cancer, reduction of cholesterol levels in the body (thus, preventing obesity and safe for diabetic patients) and improved mineral absorption in the gut (Dominguez et al., 2014a). ScFOS are fructose oligomers consisting of a terminal glucose group linked to 2 – 4 fructose units by β (1 \rightarrow 2) – glycosidic bonds, called 1-kestose (GF2), nystose (GF3) and

fructofuranosyl nystose (GF4), respectively (Ur Rehman et al., 2016). They are produced by the degradation of inulin or transformation of sucrose by intracellular enzymes from bacteria, fungi or yeast sources (Bedzo, Mandegari, et al., 2019). The global production output scFOS reported in 2015 was 134 kt/y and projected to almost double by 2024 (Grand View Research Inc., 2016).

Techno-economic analyses of producing SA from crystalline sucrose (Efe et al., 2013) and lignocellulose (Nieder-Heitmann et al., 2019) raw materials have been reported. Efe et al. (2013) found that sucrose raw material had the highest contribution (29 %) to the yearly operational cost and suggested the need for alternative raw materials in the future with suitable downstream processes to yield the required market purity. On the other hand, the use of lignocelluloses as cheap raw material present a challenge of high pre-treatment and hydrolysis costs (Nieder-Heitmann et al., 2019) and low product concentrations typical of xylose-based feedstocks (Jiang et al., 2017; Pateraki et al., 2016). Similarly, techno-economic studies of scFOS production from pure crystalline sucrose as stand-alone plants were reported in the literature, but with a focus on the types of enzyme systems and fermentations used rather than the raw material (Bedzo, Mandegari, et al., 2019; Mussatto et al., 2015).

This chapter focuses on exploring the economic opportunities of modifying the existing sugar mill presented in Chapter 3 to a single-stage crystallization as an avenue to improve the exergy efficiency of the mill through the valorisation of all of the resulting A-molasses to (co-)produce SA and/or scFOS in an integrated biorefinery concept. The goals are to determine the economic feasibility of the developed integrated biorefinery and to find suitable product design and process configuration options, with a specific focus on benefits of integration into a sugar mill, rather than stand-alone production. To achieve these goals, Aspen plus simulations and techno-economic assessments of seven scenarios of the integrated biorefinery are presented. These consist of single-product scenarios of SA, scFOS powder and scFOS syrup productions from all A-molasses produced, as well as multi-product scenarios using 70 % of A-molasses for SA and 30 % for scFOS powder or syrup production based on the current market sizes and perceived market access.

5.2. Materials and methods

The methodology used in this work involved simulation in Aspen Plus® of the integrated biorefinery processes as a retrofit to an existing sugar mill in different configurations. The simulation generated mass and energy balances necessary for sizing and costing of equipment and the operation costing of the process. The costing data then became the basis for the economic evaluation of all scenarios simulated. Finally, due to fluctuations in the economic parameters used leading to uncertainty of the results, a sensitivity analysis was carried out to establish the robustness of the economic performance of the scenarios.

5.2.1. Reconfiguration of the sugar mill for the integration of biorefinery scenarios

The production processes in the existing sugar mill involve milling of sugarcane and extraction to separate the juice from bagasse, clarification to purify the juice, evaporation to concentrate the juice, a three-stage crystallization to separate and recover crystalline sucrose from non-sucrose, and drying. Chapter 3 presented a detailed process description and Aspen simulation of the existing sugar mill, also reported elsewhere (Guest et al., 2019). The goal of the three-stage crystallization in sugar mills is to maximize the recovery of crystalline sucrose, as a single product, from the final molasses (Rein, 2007). The first (A) stage of sugar crystallization in a typical sugar mill produces A-massecurite (a mixture of crystals and mother liquor), which is separated into the mother liquor called A-molasses and the raw sugar crystals in a centrifuge. In the current process the A-molasses is processed further in two more (B and C) stages to crystallize out more sugar of low quality, which is recycled to the A-pan in the form of magma and re-melt mixture, leaving a sucrose depleted C-molasses (Mkwanzani et al., 2019). The C-molasses composing of 78.1 % water, 5.5 % sucrose and 16.4 % non-sucrose (i.e., glucose and fructose) is sold as a by-product (Chapter 3).

Based on the exergy analysis results in Chapter 3, using only the A-crystallization was proposed to improve the exergy efficiency of the sugar mill. This proposal requires a reconfiguration of the crystallization unit of the existing sugar mill to a single-stage sugar boiling and subsequent crystallization process, which starts on fine seed slurry instead of

magma in the usual three-stage system. The seed slurry, which is negligible in quantity relative to the syrup, was assumed to have a negligible effect on the mass balance of the new system (Ziegler, 1978). Upon the reconfiguration, the sugar mill produced 23.1 t/h (i.e., approx. 28 % reduction from the initial production of 32.0 t/h (Chapter 3) in addition to 21.2 t/h of sucrose-rich A-molasses with 78 % sugars (69.8 % sucrose, 15.1 % glucose and 15.1 % fructose on total solids basis) and 22 % water. The valorisation of this simulated A-molasses as feedstock is the basis for biorefineries development and simulation in the present chapter. Figure 5-1 illustrates a simplified configuration of the proposed integrated biorefinery annexed an existing South African sugar mill. The description of the process and simulation of the products (SA and scFOS) developed from the A-molasses feedstock are presented in the subsequent sections.

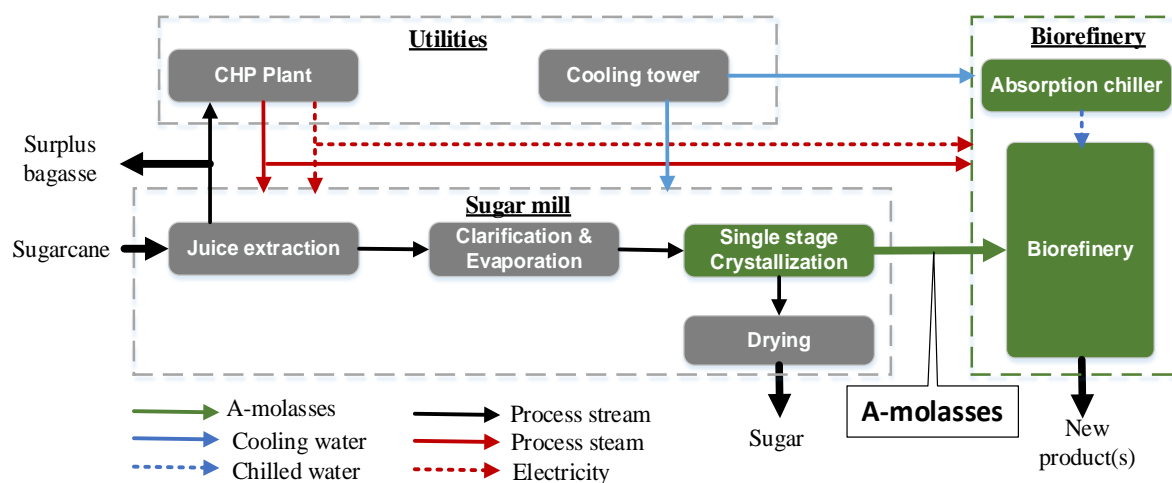


Figure 5-1: Process configuration of proposed integrated sugarcane biorefinery

5.2.2. Process description and simulation of the biorefinery products

The processes described in this study were simulated in Aspen Plus[®], which has all required thermodynamic database to define the processes and generate mass and energy balances necessary to assess them. The data required for the simulation were mainly adapted from various literature sources for cost-effective production based on reasonable assumptions.

5.2.2.1. Succinic acid (SA)

Succinic acid (SA) can be synthesized from biomass sugars by microorganisms through anaerobic fermentation (Jiang et al., 2017; Morales et al., 2016). The processes include seed train for microorganism growth, fermentation of feedstock, and recovery and purification of product from the fermentation broth. Literature data were adapted for the simulation of the fermentation (S. Chan et al., 2012) and downstream purification (Morales et al., 2016) processes, giving SA purity of ≥ 99.8 wt. %.

The A-molasses feedstock is diluted with water to sugars concentration of 150g/L required for genetically engineered *E. coli* in batch fermentation (S. Chan et al., 2012). *E. coli* is a promising bio-SA-producing strain and its advantages over the natural SA producer *A. succinogenes* include easy genetic engineering, fast growth rate and has no need for complex nutrients (McKinlay et al., 2007; Morales et al., 2016; Vaswani, 2010), yet with comparable yields, titres and productivity of SA (S. Chan et al., 2012). These properties are required for flexibility of feedstock use and reduce the complexity of fermentation and cost of downstream processes (Vaswani, 2010). Moreover, genetically modified *E. coli* has been successfully used for commercial production of SA from commodity sugars (Bidy et al., 2016; Jansen & van Gulik, 2014; Vaswani, 2010). The production process includes seed train, fermentation and downstream processing (DSP) sections. Figure 5-2 shows the Aspen simulation flowsheet of the various sections of the SA production process

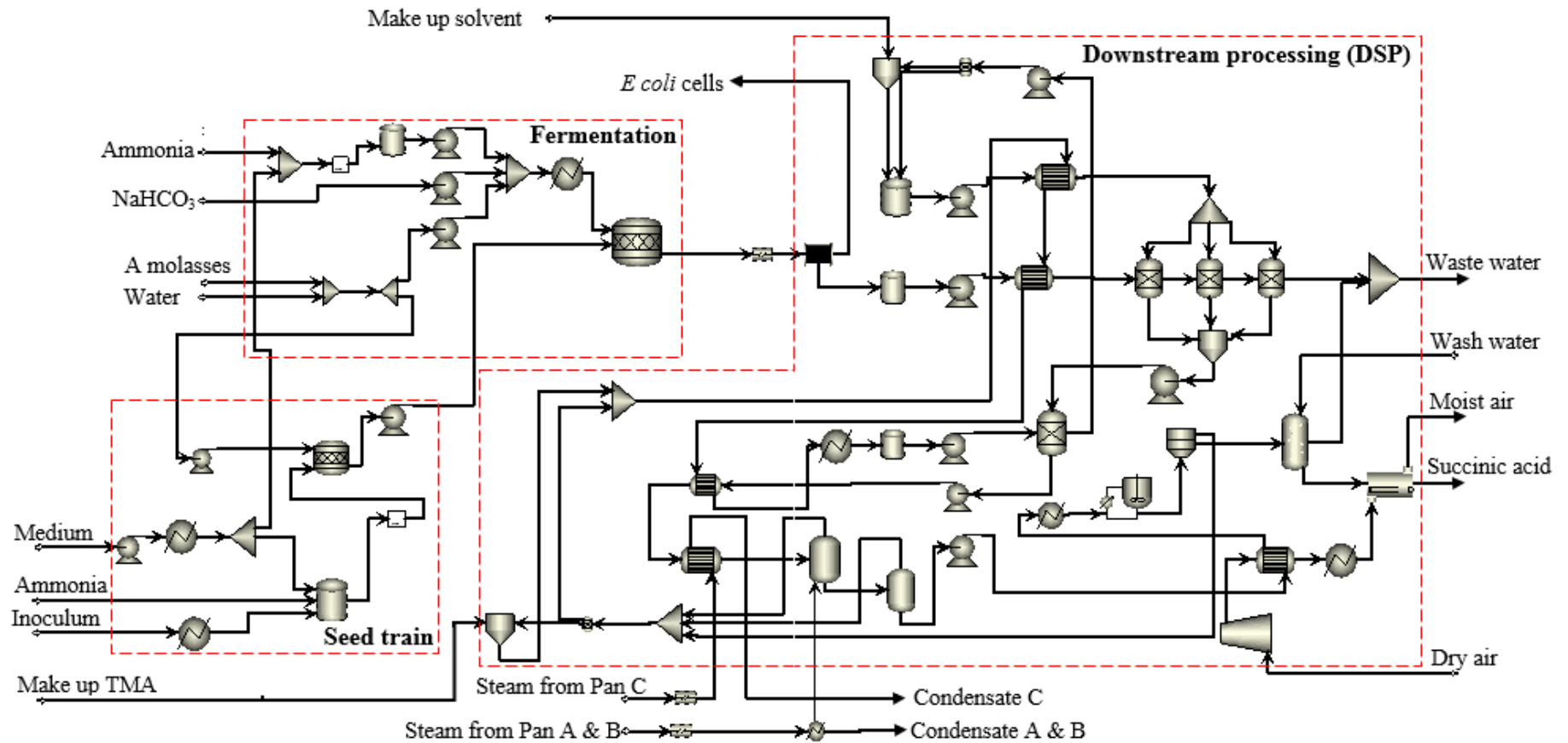
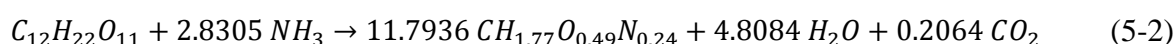
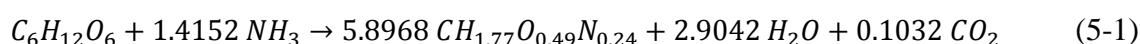


Figure 5-2: Aspen simulation flowsheet of the succinic acid (SA) production process including Seed train, Fermentation and DSP unit

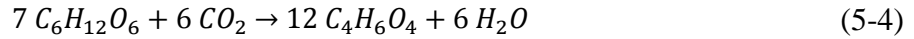
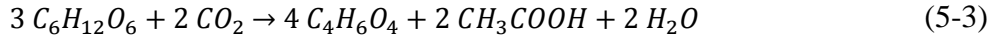
Seed train

A portion of the fermentable feedstock, about 10% (v/v), was diverted to facilitate cell growth with some succinate production in a two-stage seed train (Nieder-Heitmann et al., 2019; Vaswani, 2010) due to non-trivial issues with large scale cell recycle (Jansen & van Gulik, 2014). A simple mineral salt medium (AM1) sterilized at 120 °C (S. Chan et al., 2012) supplied the nutrients required for the micro-organism to grow. Equations (5-1) and (5-2) show the simplified growth reaction of *E. coli* on glucose (or fructose) and sucrose, respectively, modelled in RStoic reactor at 37 °C



Fermentation

The remaining 90 % of the diluted feedstock, together with the micro-organism inoculum, nutrient medium and a carbonate salt are fed into the fermenter. The carbonate salt, NaHCO₃ was used as a cheaper source of required CO₂ for the fermentation and to control the pH at 7.0 for the succinate production. (S. Chan et al., 2012; Jiang et al., 2017; Thakker et al., 2012). Research is on-going to develop suitable strains that can produce the free acid at lower pH (≤ 3) conditions in a cost-effective way (Jansen & van Gulik, 2014). However, it is assumed in this study that *E. coli* strain produced the free SA instead of a succinate salt to simplify the process and for ease of purification (Morales et al., 2016). Most *E. coli* strains lack the genes that utilise sucrose and therefore require metabolic evolution for that purpose. Chan et al. (2012) reported a metabolically engineered *E. coli* strain KJ122-KJSUC-24T that hydrolysed all sucrose in the molasses into glucose and fructose and subsequently converts them to SA (C₄H₆O₄) at a yield of 0.96 g per g of sugars consumed with titre and productivity of 55.8 g/L and 0.77 g/L/h, respectively, which are comparable to productions using *A. succinogenes*. The batch fermentation at a temperature of 37 °C, pH of 7.0 and an optimal molasses concentration and incubation time of 150 g/L and 72 h, respectively (S. Chan et al., 2012) was modelled in RStoic reactor as shown by equations (5-3) and (5-4). The fractional conversions of glucose and fructose in each reaction were calculated based on the expected composition of the fermentation broth (S. Chan et al., 2012).



Downstream recovery and purification (DSP)

E. coli cells and solid impurities in the fermentation broth were removed by filtration and treated as waste (Nieder-Heitmann et al., 2019). Based on the assumption of free acid production, the next step was the effective extraction of the SA into the organic phase using tri-*n*-octylamine (TOA) diluted with 1-octanol as the extractant in three successive stages of reactive extraction process, leaving an aqueous waste (Morales et al., 2016). Subsequently, a mixture of trimethyl-amine and water was used in a back extraction step to recover the product from the organic phase (the extractant) (Ahn et al., 2017; Morales et al., 2016). The purification process includes a distillation step to eliminate volatile acetic acid from the product and to concentrate the product, feeding back the trimethyl-amine-water mixture into the back extraction column to reduce cost (Ahn et al., 2017). This is followed by crystallization and washing to purify the product. The separation and purification processes selected were adequate to achieve a final purity and recovery of 99.9 % and 99.7 %, respectively for high-value applications (Morales et al., 2016). Finally, the purified SA crystals are dried for packaging.

5.2.2.2. Fructooligosaccharides (scFOS)

ScFOS are produced from the transformation of sucrose by a fungal enzyme with transfructosilase activity, called β -D-fructofuranosidase (FFase) (Bedzo, Trollope, et al., 2019). The process involves an enzyme production step and scFOS production cum purification step, where the scFOS product is separated from the mono- and disaccharides. Figure 5-3 shows the Aspen flowsheet of the scFOS plant. The process description and conditions for the simulation are presented in the following sections.

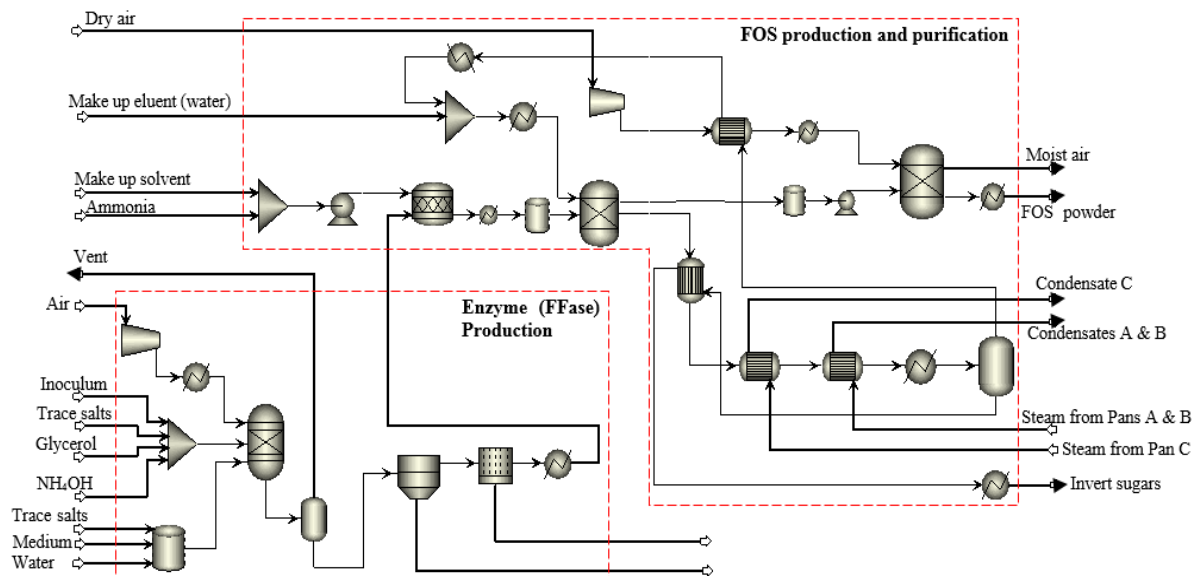


Figure 5-3: Aspen simulation flowsheet of Production of β -D-fructofuranosidase (FFase) and scFOS

Enzyme production

Enzyme production in the same facility with scFOS was noted to enhance the economic performance of the plant (Bedzo, Mandegari, et al., 2019; Katarína Vaňková et al., 2008). The inoculum of the yeast strain (*Pichia pastoris*) grown in the preparation laboratory was transferred into a sterilised fermenter to produce the enzyme, β -D-fructofuranosidase (FFase), feeding on the carbon source (glycerol) and oxygen supplied by air. The production medium and trace salts maintained the growth of the cells until fermentation started, while ammonium hydroxide supplied nitrogen for the yeast and maintained the pH at 5 for maximum growth (Bedzo, Mandegari, et al., 2019). The enzyme, which was secreted together with other proteins from the yeast cells into the supernatant, was separated from the dead cells in a centrifuge (Katarína Vaňková et al., 2008). An ultrafiltration step follows to undesired constituents of the supernatant. Prior to the separation steps, the fermentation mixture was flashed to remove gaseous effluents. The purified enzyme was stored at 4 °C to maintain its maximum activity for the scFOS production process (Bedzo, Mandegari, et al., 2019). The enzyme production was simulated using the RYield Aspen model at the optimal growth and protein expression temperature of 30 °C, specifying the desired enzyme yield (Bedzo, Mandegari, et al., 2019).

ScFOS production and purification

ScFOS production by the transfructosylation of sucrose was simulated using RYield reactor model in Aspen. The A-molasses was diluted to a sucrose content of 600 g/L (Bedzo,

Mandegari, et al., 2019) and reacted with the FFase from the enzyme production stage at a temperature of 62 °C. The sucrose conversion was 92 % with 58 % scFOS production, 29 % glucose and 5 % fructose according to the experimental data presented by Ur Rehman et al. (2016) using molasses. Bedzo et al. (2019) reported similar results using sucrose solution while much lower yield of 52 % was reported using sugar mill A-molasses (Coetzee, 2019). It is assumed that no pre-treatment was required for A-molasses, which is relatively pure compared to the usual final molasses (C-molasses) available as sugar mill by-product. The scFOS distribution adopted for the simulation was 38.8 % GF₂, 47.8 % GF₃ and 13.4 % GF₄ for residence time of 8 h (Bedzo, Trollope, et al., 2019; Soni & Tsai, 2016).

The reaction mixture after enzymatic transformation of sucrose contained 35.3 % glucose, 18.6 % fructose, 5.6 % sucrose, 15.7 % GF₂, 19.4 % GF₃ and 5.4 % GF₄ (Coetzee, 2019). The scFOS (GF₂ + GF₃ + GF₄) was separated from the sugar impurities using simulated moving bed chromatography (SMBC) to achieve 90 % purity and a yield of 97 % (K. Vaňková & Polakovič, 2012). The extract composed mainly of glucose, fructose and unreacted sucrose is concentrated (from about 34 ° to 60 ° Brix) for sale as a by-product, while the raffinate, which contains the product, was either spray-dried to produce scFOS powder or concentrated to 75 % Brix as scFOS syrup (Soni & Tsai, 2016). The required quantity of air for the spray dryer was determined with the help of design specification in Aspen, given the air inlet temperature of 180 °C and the desired outlet temperature of 90 °C (Bedzo, Mandegari, et al., 2019). It is worth noting that a high concentration of raffinate (255 g/L) attained by the SMBC process, which is twice higher than conventional chromatography (K. Vaňková & Polakovič, 2012) is advantageous for the exergy and economics of the spray dryer and consequently, the entire process.

5.2.3. Process utilities integration

The integrated processes for scFOS and/or SA production were designed to supply both the sugar mill and the new products plants (referred to here as biorefineries) with electricity and hot utilities from a common source, i.e. the existing cogeneration system. This was possible considering the unused steam available from the B and C crystallization stages and the drying unit, due to the reduction in sugar production when A-molasses is diverted to these products. Moreover, additional steam can be made available to the biorefinery by reducing sundry steam losses (Chapter 4) to the minimum possible attained through steady state best practices (Hoekstra, 1995; B. M. J. Reid & Rein, 1983). In addition, excess steam use can be minimized

through reducing imbibition water use in the extraction process (Chapter 3) and A-masseccite recycling (Mkwanzani et al., 2019). Furthermore, the sugar mill being analysed in this study produced 77.4 t/h bagasse (Chapter 3) and burned 50.6 t/h of it in the existing boiler, leaving a surplus of 26.8 t/h (Chapter 4) that can be added to the boiler to accommodate the steam requirement of the biorefinery, considering that it is possible for the current sugar mill boiler to produce up to 10 % more steam than its current capacity (Dr W Lawlor 2019, RCL FOODS, pers. comm., 18 March). Similarly, the sugar mill cooling tower was expected to take on 10 % extra capacity to provide the cooling water for the biorefinery. In terms of electricity, the typical sugar mill under study currently produces 12.5 % surplus electricity of the sugar mill requirement (32 kWh/t of cane) (Chapter 4) that can be made available to the biorefinery instead of selling it to the grid since currently, there is no economic motivation to do so. Therefore, the only costs involved in providing utilities for the biorefinery were payment for the surplus bagasse and equipment cost of an absorption chiller package powered by available waste heat from the cogeneration system (Birru, Erlich, Beyene, & Martin, 2016; Dogbe, Mandegari, & Görgens, 2019) as proposed in Chapter 4 to provide the required chilled water, since the sugar mill has no need for chilled water. Export (surplus) water from the sugar mill (Chapter 3) was also channelled to the biorefinery to supplement water use. These integrations offered economic benefits to the developed biorefinery scenarios compared to stand-alone biorefineries.

5.2.4. Biorefinery scenarios development

Annexed to the existing sugar mill producing crystalline sugar, a new product (SA and/or scFOS) biorefinery was integrated to utilise A-molasses as feedstock. Seven (7) scenarios of this integrated biorefinery were simulated in Aspen Plus[®] considering different product combinations and feedstock (A-molasses) share, each scenario making use of all available A-molasses for SA and/or scFOS production, and included the same quantity of crystalline sucrose as a product. Table 5-1 summarises the developed scenarios. Scenarios SA-S, S-F_P and S-F_S are single-product scenarios producing SA scFOS powder and scFOS syrup from all A-molasses. In addition to the products, scenarios S-F_P and S-F_S each produce invert sugar syrup as a by-product. The difference between the process configurations of the two scenarios is that S-F_P has a spray-drying unit to produce scFOS powder in place of an evaporation unit in S-F_P to concentrate the product to syrup. Scenario SA70-S-F_P is a multi-product scenario producing both SA from 70 % of the A-molasses and scFOS powder as well as invert sugar by-product from 30 % A-molasses in the same biorefinery. Scenario SA70-S-F_S has the same

configuration as SA70-S-F_P but produces scFOS syrup instead of powder. The split percentages of the A-molasses for each product was determined based on estimated 2018 global market sizes of 440 kt/y for SA (Klein et al., 2017) and 160 kt/y for scFOS (Grand View Research Inc., 2016), as well as the market growth rates. These feedstock allocations yielded 9 % and 6 % of the SA and the scFOS market sizes, respectively so as not to flood the market. The consequence of flooding the market with new productions is discussed in the next section. The last set of scenarios SA70I-S-F_P and SA70I-S-F_S correspond to the configuration of the previous multi-product set (SA70I-S-F_P and SA70I-S-F_S) in terms of A-molasses share between the two products. However, invert sugar from the scFOS plant section was not concentrated for sale as by-product, but instead it was combined with the 70 % A-molasses as feedstock to produce additional SA. SA70I-S-F_P produced scFOS powder, while SA70I-S-F_S produces scFOS syrup, each in addition to SA crystals.

Table 5-1: A summary of the simulated biorefinery scenarios

	SA-S	SA70-S-F _P	SA70-S-F _S	SA70I-S-F _P	SA70I-S-F _S	S-F _P	S-F _S
Feedstock							
A-molasses for SA, %	100	70	70	70	70	-	-
A-molasses for scFOS, %	-	30	30	30	30	100	100
Invert sugar for SA	-	-	-	√	√	-	-
Product for sale							
Succinic acid (SA)	√	-	-	-	-	-	-
scFOS powder	-	√	-	√	-	√	-
scFOS syrup	-	-	√	-	√	-	√
Invert sugar (by-product)	-	√	√	-	-	√	√

Notes: 1. The “-“ indicates “not applicable” while “√” shows “applicable”

2. SA: succinic acid, F_P: scFOS powder, F_S: scFOS syrup, S: sugar, I: invert sugar, 70: 70% A-molasses

5.2.5. Techno-economic assessment (TEA)

Techno-economic assessment (TEA) was conducted to inform process/product selection and investment decision-making towards implementation (M. A. Mandegari et al., 2017). The simulation of the biorefinery scenarios in Aspen Plus[®] generated the mass and energy balance and costing data of some common process units like pumps, heat exchangers and columns for the economic assessment. The costs of all other process units were taken from the literature, adjusted for capacity and time of the study (M. Mandegari et al., 2018). The calculation of the total capital investment (TCI), operation costs (OPEX) as well as the cash flow analysis were

accomplished in Microsoft Excel spreadsheet (Humbird et al., 2011). Table 5-2 presents the set of economic parameters employed for the techno-economic analyses.

Table 5-2: Economic parameters for the techno-economic analysis

Parameter	Value	Parameter (prices)	Value
Annual Operational hours	5000 ^a	Sugar (US\$/t)	347 ^c
Project life time (years)	20	C-molasses (US\$/t)	192 ^c
Project year	2018	Bagasse (US\$/t)	26 ^a
Depreciation method	Straight line	Succinic acid (US\$/t)	2500 ^d
Salvage value	0 ^b	scFOS powder (US\$/t)	3200 ^e
Discount rate (%)	9.7 ^h	scFOS syrup (US\$/t)	471 ^f
Income tax rate (%)	28	Invert sugar (US\$/t)	304 ^g

^a From (Dogbe et al., 2019) ^b From (Ali Mandegari et al., 2017) ^c From (OECD/FAO, 2018)

^d From (Klein et al., 2017) ^e From (Bedzo, Mandegari, et al., 2019) ^f From (RCL Foods, South Africa)

^g Invert sugar price is about 25% more than sugar price (<http://www.sugar-and-sweetener-guide.com/inverted-sugar.html>). This price was calculated for syrup with about 70% solids

^h The minimum acceptable IRR for real term analysis is 9.7 % (Ali Mandegari et al., 2017). However, a minimum IRR of 20 % was targeted in the calculation of minimum selling prices (MSP) in this work to make the project attractive to private investors.

The profitability analysis of the biorefinery scenarios was based on the marginal change in revenue of the existing sugar mill, together with the additional capital and operational costs of the biorefinery, as these are brownfields extensions to existing sugar mills, rather than greenfield construction of the combined facility. Most of the existing sugar mills in the South African context are already fully amortised. However, sugar mill cost values from elsewhere were used to validate that the brownfields and greenfields approaches yielded the same results of economic performance. Revenues were determined on the assumption that a specific market can absorb up to 10 % extra production of the market size without any change in the selling price of that product (Nieder-Heitmann et al., 2019). For more than 10 % production of the market size, the price of the product was assumed to reduce by the percentage of the market size produced, since it is not clear by how much a bio-based product price reduces when there is an oversupply.

Parameters commonly used to indicate economic performance include the internal rate of return (IRR), the net present value (NPV), minimum selling price (MSP) and the payback period, all of which are presented in this study to indicate the profitability of the biorefineries (Ziyai et al., 2019). The NPV and IRR were determined based on the minimum required IRR (discount rate) of 9.7 %. IRR is the rate of return evaluated at the product market price when

the NPV is zero. The plant is considered profitable when the IRR is more than discount rate (Bedzo, Mandegari, et al., 2019). On the other hand, an MSP below the prevailing market price of a product indicate good economic performance of the biorefinery. Therefore, the MSP was evaluated at a high IRR of 20 %, considering a safe margin for private investors in South Africa. The sensitivity of the MSP to the changes in major economic parameters was also presented to ascertain any risk associated with the project.

5.3. Results and discussion

Seven scenarios of integrated biorefinery were simulated annexed an existing sugar mill using all of the available A-molasses from the mill. The configuration include single-product SA, scFOS powder and scFOS syrup as well as multi-product scenarios combining SA with either scFOS power or syrup in 70 – 30 % A-molasses ratio. The mass and energy balances as well as the techno-economic results of the simulated scenarios are presented and discussed in this section.

5.3.1. Mass and energy balances

The feedstock distributions and the production rates of the studied scenarios are summarised in Table 5-3. Scenarios SA70-S-F_P and SA70-S-F_S produced 9.0 t/h scFOS powder and 13.5 t/h scFOS syrup, respectively from 30 % A-molasses, each in addition to 13.5 t/h saleable invert sugar by-product (composing of 58.7 % glucose, 30.6 % fructose, 7.2 % sucrose and 3.5 % scFOS) and 40.5 t/h SA from the remaining 70 % A-molasses. The corresponding multi-product scenarios SA70I-S-F_P and SA70I-S-F_S utilised the invert sugar as co-feedstock with the 70 % A-molasses to increase SA production to 51.0 % (i.e., 26 % increase). Considering the sugar mill operation time of 5000 h/y (Chapter 4) adopted for this study, the production rates of SA in each scenario compared to the global market size (440 kt/y) are 13 % (SA-S), 12 % (SA70I-S-F_P and SA70I-S-F_S) and 9 % (SA70I-S-F_P and SA70I-S-F_S). The scFOS (powder and syrup) production capacity was 19 % of the market size (160 kt/y) for scenarios both S-F_P and S-F_S and 6 % each for all four multi-product scenarios. For production rates of more than 10 % of a product's market size, the market price reduction mentioned in Section 5.2.5 was applied in determining their economics of the scenario.

Table 5-3: Feedstock distribution and production rates of the studied scenarios

Scenario	Unit	SA-S	SA70-S-F _P	SA70-S-F _S	SA70I-S-F _P	SA70I-S-F _S	S-F _P	S-F _S
<i>Feedstock distribution</i>								
A-molasses for SA	kt/y	106.0	74.0	74.0	74.0	74.0	-	-
A-molasses for scFOS	kt/y	-	32.0	32.0	32.0	32.0	106.0	106.0
Invert sugar for SA	kt/y	-	-	-	8.6	8.6	-	-
<i>Product rate</i>								
Succinic acid (SA)	kt/y	58.0	40.5	40.5	51.0	51.0	-	-
scFOS powder	kt/y	-	9.0	-	9.0	-	31.0	-
scFOS syrup	kt/y	-	-	13.5	-	13.5	-	45.0
Invert sugar	kt/y	-	21.5	21.5	-	-	70.5	70.5

The overall energy demand of the studied scenarios including the surplus bagasse from the baseline sugar mill (Chapter 4) added to the existing boiler to supply heating utilities to the biorefinery is depicted in Figure 5-4. Electricity is principally consumed by air compressors and is highest in scenario S-F_P (2.73 MW in Figure 5-4) because of the voluminous amount of air required for spray drying. Scenario SA-S has the highest cooling demand of 36.1 MW (predominantly, chilled water), followed by the multi-product scenarios in the order of increasing SA production. This is attributed to the fermenters and crystallizers, which are the major heat sinks in the SA production process (Efe et al., 2013). The single-product scFOS powder scenario S-F_P has the lowest cooling demands of 21.35 MW, mainly for cooling of reactors and products. This is slightly lower than scenario S-F_S (21.54 MW) due to the difference in scFOS product rates (31.0 kt/y powder vs. 45.0 kt/y syrup as seen Table 5-3). The cooling water rates of all scenarios were found to be more 10 % of the current sugar mill cooling tower capacity of 2 811 t/h, except scenario SA-S (9.8 %). Therefore, a new cooling tower of the required cooling water capacity was included in the equipment costing of each scenario for the economic assessment.

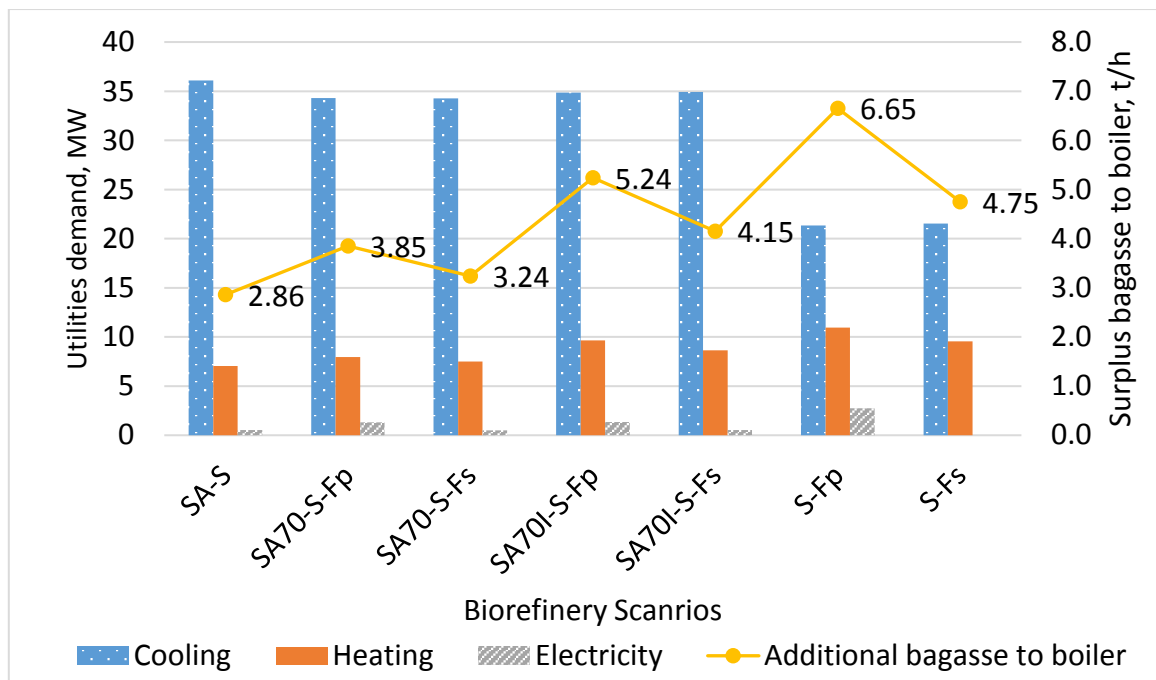


Figure 5-4: Utility requirements of the A-molasses biorefinery scenarios

The heating utility (LP and HP steam) demands shown in Figure 5-4 for each scenario is 4.7 MW less than the actual demand because of the 4.7 MW heating utilities from the crystallization and drying units of the sugar mill available through the integration. Scenario S-F_P shows the highest heating demand intensity of 10.9 MW mainly attributed to the spray dryer in the scFOS powder process as the dominant heating utility user (Bedzo, Mandegari, et al., 2019). On the other hand, the high heating demand of scenario S-F_S (9.55 MW) in the absence of the spray dryer is due to heating required to concentrate to scFOS in an evaporator (also energy-intensive) to syrup of 75 % solids for sale (Soni & Tsai, 2016). The same variations in the heating demand of scFOS syrup and powder productions show in the multi-product scenarios as well.

The sugar mill considered in this study has 26.8 t/h surplus bagasse (Chapter 3) which is sold as animal feed or for paper and furfural production (M. J. Reid, 2006). Part of this surplus bagasse is added to the 50.6 t/h already burnt in the existing boiler (Chapter 4), producing more steam to accommodate the heating demand of the biorefinery. Figure 5-4 shows that the surplus bagasse required for scenarios SA70I-S-F_P (5.24 t/h) and S-F_P (6.65 t/h) are 10.3 % and 13.0 % of the current boiler capacity, respectively while all other scenarios require less than 10 %. Based on the economic performance of these scenarios compared to others, it may be necessary to reduce the plant capacity to get heating utility rate below 10 % of the current boiler, rather than to purchase a new boiler. The effect of purchasing a new boiler on the economics of the

plant is demonstrated through a sensitivity analysis using changes in the fixed capital investment (Section 5.3.2.4).

5.3.2. Techno-economic assessment

This section evaluates the economics of the simulated biorefinery scenarios considering the integration design and the different feedstock and product splits. The techno-economic assessments of the integrated biorefinery scenarios were conducted based on the marginal cost of the sugar mill added to the production costs as expenses for each biorefinery scenario and not on the full cost of the sugar mill. The components of the marginal cost include the reduction in crystalline sucrose and C-molasses sale of the sugar mill, as well as the cost of surplus bagasse added to the existing boiler. A sensitivity analysis is also presented to establish the profitability of the scenarios under different economic conditions.

5.3.2.1. Capital investment and Production costs

The production cost of each scenario consists of the capital investment to add the new products plant to the existing sugarcane mill, and the yearly operational costs thereof, including all marginal cost effects due to the integration. Table 5-4 summarises the main cost components of each scenario including the contributions of each plant area to the total installed equipment cost (TEPC), the total capital investment (TCI) and the operation cost (OPEX). The TEPC consists of the process battery limit (sum of plant area installed costs) and the utilities components including the costs of a new cooling tower and an absorption chiller for the biorefinery, based on cooling requirements of each scenario. Because of the integration into the sugar mill described in Section 5.2.3, no boiler cost was included in the total equipment costing.

Table 5-4: Total capital investment, distributed installed equipment cost and operation cost of biorefinery scenarios

	SA-S	SA70-S-F _P	SA70-S-F _S	SA70I-S-F _P	SA70I-S-F _S	S-F _P	S-F _S
Unit	MUS\$						
<i>Plant area installed cost,</i>							
FFase production	-	1.0	1.0	1.0	1.0	2.2	2.2
scFOS production	-	13.5	4.0	13.0	3.4	27.4	7.6
<i>E. coli</i> seed train	3.7	2.9	2.9	3.6	3.6	-	-
SA fermentation	19.7	15.3	15.3	19.1	19.1	-	-
Downstream processes	12.9	10.3	10.3	11.8	11.8	-	-
<i>Inside battery limit (IBL)</i>	36.3	43.0	33.5	48.6	39.0	29.6	9.8
Utility components	3.6	3.4	3.4	3.3	3.3	2.4	2.4
Total equipment cost (TEPC)	39.9	46.4	36.9	51.8	42.2	32.0	12.3
Fixed capital investment (FCI)	76.8	89.6	70.9	100.2	81.5	61.7	23.1
Working capital (5 % FCI)	3.9	4.5	3.6	5.0	4.0	3.1	1.2
Total capital investment (TCI)	80.7	94.1	74.5	105.2	85.5	64.8	24.3
Biorefinery operating cost	15.2	11.6	11.1	13.8	13.3	2.4	1.5
Marginal cost of sugar mill	25.1	25.2	25.2	25.4	25.3	25.6	25.4
Total operating cost (OPEX)	40.3	36.8	36.3	39.2	38.6	28.0	26.9

The multi-product scenario SA70I-S-F_P shows the highest TEPC and consequently the highest TCI of 51.8 MUS\$ and 105.2 MUS\$, respectively. Generally, the single-product scenarios SA-S, S-F_P and S-F_S have relatively higher TEPC and consequently, higher TCI of 80.7 MUS\$, 64.8 MUS\$ and 24.3 MUS\$, respectively compared to the multi-product scenarios (TCI of 94.1 MUS\$, 105.2 MUS\$ and 85.5 MUS\$ for scenarios SA70-S-F_P, SA70I-S-F_P and SA70I-S-F_S, respectively). This is due to the effect of economy of scale considering 100 % feedstock used for each product in the single-product plants compared lower capacities (30 % scFOS and 70 % SA) in the multi-product scenarios. Nonetheless, the risk of investment for product diversification might be less than single product due to reduced exposure to price fluctuations in one particular market.

It can be observed from Table 5-4 that the scFOS production is less capital intensive than the SA process (TCI of 64.8 MUS\$ for scenario SA-S vs. 80.7 MUS\$ for scenario S-F_P). FFase (enzyme) production contributes less than 10 % of the total equipment cost of the scFOS powder (2.2 MUS\$ vs. 27.4 MUS\$, 1.0 MUS\$ vs 13.5 MUS\$ and 1.0 MUS\$ vs 13.0 MUS\$ for scenarios S-F_P, SA70-S-F_P and SA70I-S-F_P, respectively) irrespective of the production capacities. This cost distribution justifies the preference of the on-site scFOS enzyme

production to outsourcing it, which may be expensive (Bedzo, Mandegari, et al., 2019; Katarína Vaňková et al., 2008). The spray dryer emerges the most expensive equipment in the scFOS powder plant because eliminating it reduces the EPC of the scFOS syrup scenarios S-F_S, SA70-S-F_S and SA70I-S-F_S by 66.8 %, 65.5 % and 68.2 %, respectively compared to their corresponding scFOS powder scenarios. These results agree with Bedzo et al. (2019) whose results show a reduction of scFOS powder production TEPC by 68.2 % to produce scFOS syrup for a free enzyme system.

In succinic acid scenario SA-S, the fermentation section contributes 49.4 % to the TEPC of the plant probably due to the requirement for stand-by fermenters (Humbird et al., 2011). However, this contribution is less than 63.8 % reported by Efe et al. (2013) on SA production from sucrose. This reported high fermenter contribution is due to low SA production titre (13.7 g/L, compared to 55.8 g/L used in this study) resulting from low pH (4) fermentation strategy. Nieder-Heitmann et al. (2019) also reported a higher TEPC of 51.3 MUS\$ (seed train and cellulose plant), fermentation and downstream processing plant areas) for 13.5 t/h SA production from sugarcane lignocellulosic residues compared to 36.3 MUS\$ for SA production rate of 11.6 t/h (58.0 kt/y) in scenario SA-S. However, considering that less dilute sugar stream (< 100g/L) was used and consequently larger equipment compared to 150 g/L in this study, the costing presented here show more robustness. The effect of changes in the capital costing on profitability is detailed in the sensitivity analysis.

The OPEX of each scenario shown in Table 5-4 include the cost of producing the new product and the marginal changes in the production costs of the sugar mill, charged as an expense to the biorefinery. The reduction in crystalline sucrose sale and all C- molasses sale was paid as raw material (A-molasses) cost, which is the same for all scenarios (24.8 MUS\$), while the cost of changes to the utilities of the sugar mill was paid in the form of the cost of surplus bagasse added to the existing boiler. The single product SA scenario SA-S recorded the highest OPEX of 40.3 MUS\$, while the scFOS syrup scenario S-F_S has the lowest (26.9 MUS\$). This is attributed to the costs of other significant components such as CO₂ source (NaHCO₃) for the succinate fermentation (114 US\$/t SA) and wastewater disposal (70 US\$/t SA) compared to the raw material cost of 428 US\$/t SA, whereas scFOS production barely had any wastewater for disposal. However, the OPEX of scFOS powder scenarios S-F_P (28.0 MUS\$) is slightly higher than the corresponding scFOS syrup scenarios S-F_S (28.0 MUS\$) due to the higher energy demand of drying scFOS into powder than concentrate it into syrup.

5.3.2.2. Effect of the integration on the capital and production costing

The integration advances minimum heating utilities (surplus bagasse) costs for the biorefinery because surpluses (4.7 MW) in the sugar mill due to reduced crystalline sucrose production were available at no cost. Other 3.2 MW heating utility is available in the boiler considering the lower limit of steam sundries and losses reported for a typical South African mill (B. M. J. Reid & Rein, 1983). Moreover, the biorefinery uses the surplus electricity in the existing boiler (Dogbe et al., 2019) because there is no market for it at present in the South African context. Through these benefits accrued by the integration, the average utilities cost of the integrated scenarios is only 2 % of the total OPEX whereas Efe et al. (2013) reported utilities (cooling water, boiler fuel and electricity) cost of 22 % of the OPEX for the best case SA scenario. Another principal advantage of the integration is the exclusion of the cost of a new cogeneration system, which is reported to be about 33 % of the TEPC (Nieder-Heitmann et al., 2019). The effects of higher TEPC and OPEX in the form of FCI and sugar price, respectively are included in the sensitivity analysis section to ascertain the economic uncertainty of the integration.

5.3.2.3. Profitability of new products plant

Four profitability criteria that define the economic viability of the biorefineries; NPV, MSP, IRR and payback period were evaluated as presented in Table 5-5 with annual revenues from each scenarios. All the parameters were determined at a minimum required IRR of 9.7 % except the MSP evaluated at 20 % IRR to attract private investors. For the multi-product scenarios, the MSP of SA, scFOS powder and scFOS syrup were determined by applying the MSP obtained from the single-product scenarios SA-S, S-FP and S-FS, respectively (Mandegari et al., 2018). The MSP is used to compare the results of this study with findings from the literature because it is an intensive parameter (like the IRR and payback period) but also independent of product prices, which vary widely in the literature. Moreover, an IRR of more 9.7 % and the lower MSPs products compared to the market prices clearly indicates that the developed biorefineries are economically viable.

Table 5-5: Annual revenues and profitability parameters of the biorefinery scenarios

	SA-S	SA70-S-F _P	SA70-S-F _S	SA70I-S-F _P	SA70I-S-F _S	S-F _P	S-F _S
Annual revenue, MUS\$ ^a							
scFOS powder	-	29.41	-	29.41	-	79.56	-
scFOS syrup	-	-	6.32	-	6.32	-	17.11
Invert sugar	-	6.45	6.45	-	-	21.43	21.43
Succinic acid (SA)	125.67	101.29	101.3	112.2	112.2	-	-
Total revenue	125.67	137.15	114.07	141.61	118.52	100.99	38.54
Profitability							
NPV, MUS\$	422	500	384	502	386	367	39
Payback, years	1.7	1.9	1.7	1.8	1.9	1.8	4.1
IRR, %	57.1	58.7	56.6	54.3	51.6	61.1	24.1
MSP of scFOS powder, US\$/t	-	730.1	-	1432	-	1133	-
MSP of scFOS syrup, US\$/t	-	-	35	-	507	-	417
MSP of SA, US\$/t	1348	1257	1222	1409	1375	-	-

^a Determined considering a target 10 % production capacity of the global market size. Scenario SA-S, SA70I-S-F_P and SA70I-S-F_S produced 13 %, 12 % and 12 %, respectively of the SA market at reduced prices of 2175 US\$/t, 2200US\$/t and 2200US\$/t, respectively. The scFOS production capacity of scenarios S-F_P and S-F_S was 19 % each of the scFOS market size, attracting 19 % reduced prices of 2592 US\$/t and 382 US\$/t, respectively.

Even though all the studied scenarios prove economically feasible, scenario SA70-S-F_S is judged the best economically having the lowest MSPs of 1222 US\$/t for SA and 35 US\$/t for scFOS syrup compared to their market prices of 2500 US\$/t and 471 US\$/t, respectively, which allow for easy market entry. On the other hand, valorising invert sugar in addition to 70 % A-molasses in scenarios SA70I-S-F_P for SA production shows the worst economic performance with the highest MSP of 1409 US\$/t for SA and 1432 US\$/t for scFOS powder, respectively. Production rate of SA is a principal reason for these outcomes. While scenario SA70-S-F_S produces 40.5 kt/y of SA, which is within 10 % of the market size (440 kt/y), SA70I-S-F_P yields 58 kt/y (i.e., 12 % of the market size), attracting an assumed 12 % reduction in SA market price due to overproduction. Therefore, the best economic model under the economic assumptions applied in this study is the multi-product design with the sale of invert sugar as a by-product while conversion it into additional SA makes it worse than the single-product design (see Table 5-5).

The MSP of scFOS powder in scenario S-F_P, 1133 US\$/t (Table 5-5) is lower than the 2610 US\$/t evaluated at 9.7 % IRR reported for a stand-alone scFOS production from sucrose

(Bedzo, Mandegari, et al., 2019). The principal reason for this large difference in MSP is attributed to economy of scale effect (31 kt/y in this study versus 2 kt/y production). In addition, the savings on TCI and OPEX (discussed in Section 5.3.2.2) of scenario S-F_P due to shared cogeneration system with the sugar mill and the use of A-molasses instead of crystalline sucrose results in a better economic performance of the integrated scenario than the stand-alone scFOS plant reported (Bedzo, Mandegari, et al., 2019).

Moreover, the MSP of SA in scenario SA-S (1348 US\$/t) is less than the 2260 US\$/t (2.26 US\$/kg) at 10 % IRR reported by Efe et al. (2013), despite the use of an IRR of 20% in the present study. One reason for the low MSP in this study is the economy of scale effect on the production of 58 kt/y of SA in this study versus 30 kt/y production in Efe et al. (2013). The fermentation approach used by Efe et al. (2013) leading to low SA titres (13.7 g/L vs. 55.8 g/L for this study) also contributed to the high MSP (2260 US\$/t) such that, a four-fold increase in the titre to 55 g/L reduced the MSP to 1480 US\$/t (Efe et al., 2013). However, this (1480 US\$/t) is still higher than the MSP of the integrated SA scenario (1348 US\$/t), especially considering that it was evaluated at IRR of 9.7 % versus the 20 % used in this work. Again, the integrated SA production from A-molasses in the present study gave a higher IRR (57.1 %) than production from lignocellulose in a stand-alone plant (Nieder-Heitmann et al., 2019). This is due to the absence of the costs of cogeneration (CHP) plant as well as pre-treatment, detoxification and enzymatic hydrolysis, which all together constituted 60 % of the TEPC for SA production from lignocelluloses (Nieder-Heitmann et al., 2019).

It is clear from the results that the integration of scFOS and SA biorefineries into a sugar mill using A-molasses feedstock result in low capital investment and operation cost and consequently gives a better economics of lower product MSPs and higher IRRs compared to the stand-alone plants reported in the literature. Whereas the use of A-molasses instead of cheaper lignocellulosic residues for SA production eliminates the pre-treatment, detoxification and enzymatic hydrolysis processes reducing the capital investment (by 28%), it avoids the energy-intensive further processing into crystalline sucrose and hence, presents a lower OPEX than utilising crystalline sucrose for the production of both scFOS and SA. Moreover, the integration with an existing sugar mill encourages efficient energy use in the mill, which contribute to the objectives of Chapters 3 and 4 and consequently, enables the sharing of the existing CHP plant with the sugar mill to avoid the associated cost. Therefore, the integration played a major role in the better economic performance of the biorefineries in this study compared to previous studies.

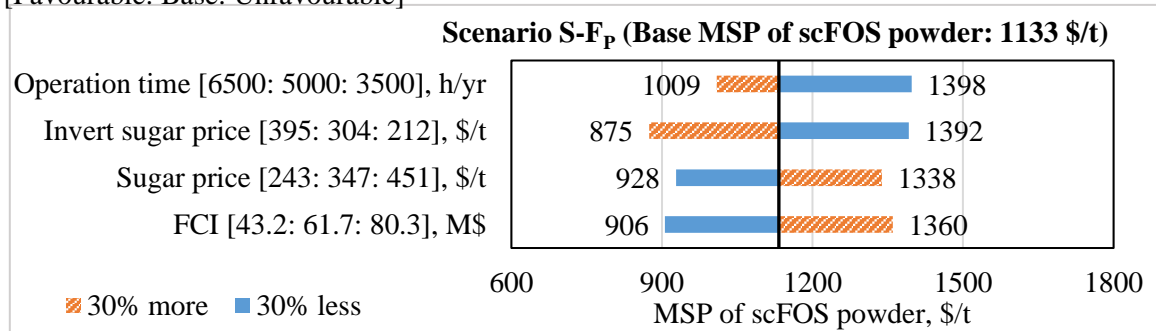
5.3.2.4. Sensitivity analysis

The economic sensitivity analysis is presented to assess the impact of uncertainties associated with the economic evaluation and to establish its robustness for investment decision making. Figure 5-5 presents the sensitivity of product MSPs to $\pm 30\%$ changes in operational time, crystalline sucrose (sugar) prices, FCI and IRR for the single-product scenarios (a) S-F_P, (b) SA-S and (c) SA70-S-F_S (the most profitable scenario of this study).

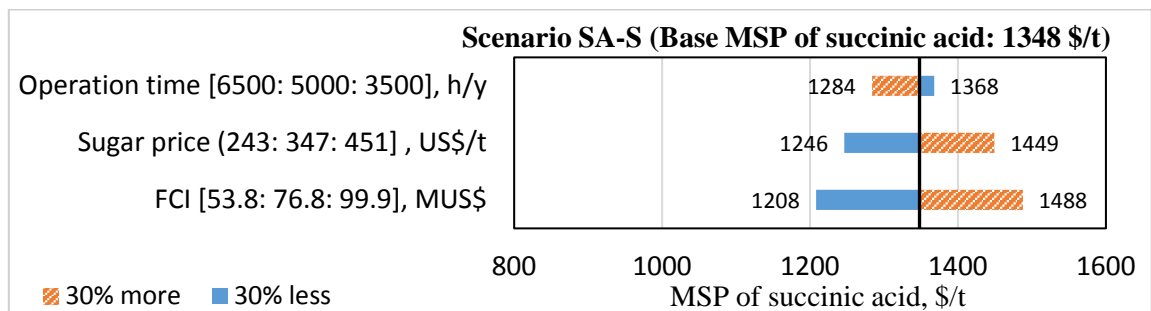
The results shows that a 30 % increase in the FCI increases the MSPs of scFOS and SA by 10 % and 20 %, respectively (Figure 5-5 (a) and (b)). The increase in FCI is possible with the inclusion of a separate boiler for biorefinery and flexible equipment sizing as a strategy for dealing with fickle market of the biorefinery products, especially in the multi-product scenario. Sugar price also shows a strong influence on the product MSP, consequently, high sugar prices (or low market prices for scFOS and SA) could warrant switching A-molasses back to crystalline sucrose production, since that equipment is installed and available. However, world sugar price is projected to increase by only 9 % in 10 years (OECD/FAO, 2018). On the other hand, the current market prices reported in the literature for SA and scFOS are far higher and are unlikely to go below the MSPs obtained in this study within the project lifetime.

Increases in the operational time of the sugar mill due to improved time efficiency also show a favourable change in the product MSPs (Figure 5-5 (a) and (b)), except in scenario SA70-S-F_S (Figure 5-5 (c)), where SA overproduction led to higher MSPs than the base (94 vs. 35 US\$/t for scFOS syrup and 1227 vs. 1222 US\$/t for SA). Lastly, invert sugar price shows the strongest influence on the MSP of scFOS powder in scenario S-F_P (Figure 5-5 (a)) because of its relatively high production rate compared to the product (70.5 kt/y vs. 31.0 kt/y). However, this study provides an alternative economic model of valorizing the invert sugar to succinic acid should the market price become unfavourable.

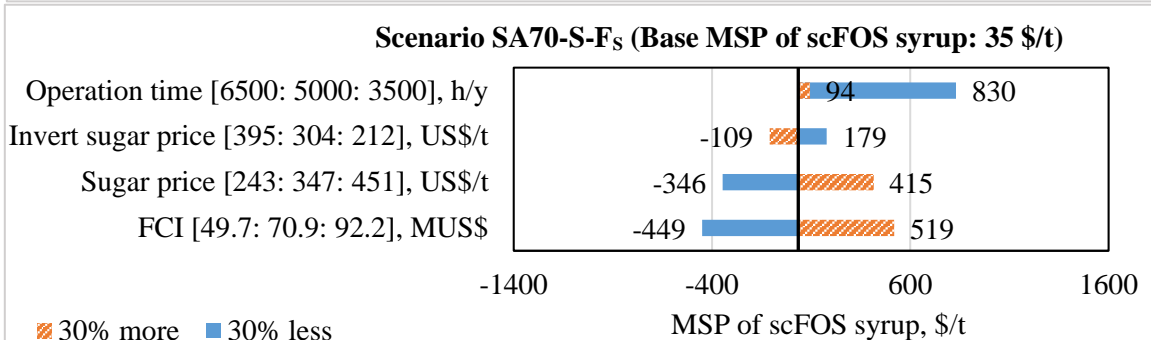
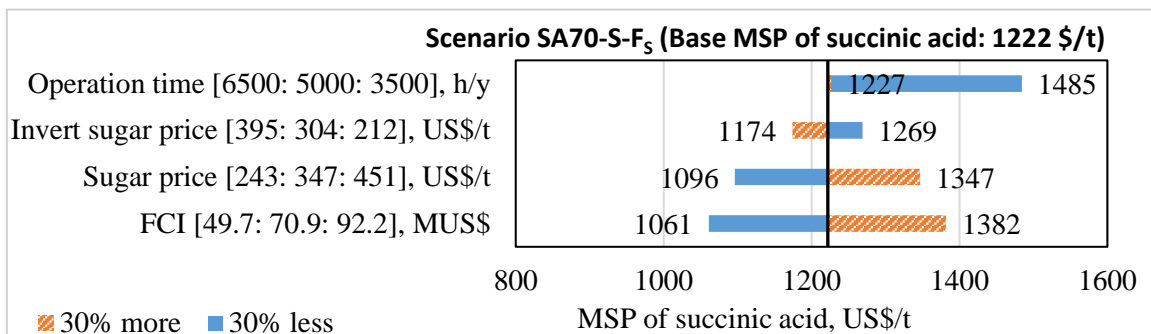
[Favourable: Base: Unfavourable]



(a)



(b)



(c)

Figure 5-5: Sensitivity of product minimum selling prices (MSPs) to ± 30% changes in major economic parameters. (a) MSP of scFOS powder in scenario S-F_P (b) MSP of SA in scenario SA-S (c) MSP of SA and scFOS syrup in scenario SA70-S-F.

5.4. Conclusion

The valorisation of A-molasses for the production of succinic acid (SA) and short-chain fructooligosaccharides (scFOS), in an integrated biorefinery with an existing sugar mill, has been investigated. The design and integration of the biorefinery were based on the outcome of exergy analysis of the sugar mill to evaluate its inefficiencies. The design considerations involved the biorefinery sharing the existing cogeneration system with the sugar mill in each scenario. This was possible due to the operation of a single-stage crystallization system instead of three stages to save energy, which had a significant influence on the profitability of the biorefinery.

The techno-economic results showed that the integration of scFOS and SA biorefineries into a sugar mill using A-molasses feedstock result gave lower capital investment and operation costs and consequently higher profitability compared to stand-alone biorefineries reported in the literature that used crystalline sucrose and lignocellulose as feedstocks. While the use of A-molasses eliminated further processing to crystalline sucrose that waste exergy or expensive pre-treatment of lignocelluloses, the integration allowed the sharing of resources that significantly influenced the profitability of the integrated biorefineries. Specifically, co-production of SA and scFOS syrup considering the products market sizes to avoid oversupplying one particular market, while also providing diversification of revenue, and easy market access was the preferred economic model yielding the lowest MSPs.

Further studies require a correlation between oversupply of bio-products and associated market price drop to determine accurately the profitability of large biorefinery plant sizes. Moreover, exergoeconomic analysis that considers the cost of system inefficiencies in addition to the investment-related cost provided by the simple economic analysis, would provide better design cost information for the optimization of the design and the selection of the preferred biorefinery for implementation.

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References

- Aghbashlo, M., Mandegari, M., Tabatabaei, M., Farzad, S., Mojarab Soufiyan, M., & Görgens, J. F. (2018). Exergy analysis of a lignocellulosic-based biorefinery annexed to a sugarcane mill for simultaneous lactic acid and electricity production. *Energy*, *149*, 623–638. <http://doi.org/10.1016/j.energy.2018.02.063>
- Ahn, J. H., Jang, Y. S., & Lee, S. Y. (2017). Succinic Acid. In C. Wittmann & J. C. Liao (Eds.), *Industrial Biotechnology: Products and Processes* (pp. 505–544). Weinheim, Germany: Wiley & VCH Verlag GmbH & Co. KGaA. <http://doi.org/10.1002/9783527807833>
- Ali Mandegari, M., Farzad, S., & Görgens, J. F. (2017). Economic and environmental assessment of cellulosic ethanol production scenarios annexed to a typical sugar mill. *Bioresource Technology*, *224*, 314–326. <http://doi.org/10.1016/j.biortech.2016.10.074>
- Bedzo, O. K. K., Mandegari, M., & Görgens, J. F. (2019). Comparison of immobilized and free enzyme systems in industrial production of short-chain fructooligosaccharides from sucrose using a techno-economic approach. *Biofuels, Bioproducts and Biorefining*, *bbb.2025*. <http://doi.org/10.1002/bbb.2025>
- Bedzo, O. K. K., Trollope, K., Gottumukkala, L. D., Coetzee, G., & Görgens, J. F. (2019). Amberlite IRA 900 versus calcium alginate in immobilization of a novel, engineered β -fructofuranosidase for short-chain fructooligosaccharide synthesis from sucrose. *Biotechnology Progress*, *35*(3), e2797. <http://doi.org/10.1002/btpr.2797>
- Biddy, M. J., Scarlata, C., Biddy, M. J., & Scarlata, C. (2016). *Chemicals from Biomass : A Market Assessment of Bioproducts with Near-Term Potential*. National Renewable Energy Laboratory (NREL), Denver.
- Birru, E., Erlich, C., Beyene, G. B., & Martin, A. (2016). Upgrading of a traditional sugar cane mill to a modern mill and assessing the potential of energy saving during steady state and transient conditions—part II: models for a modified cogeneration unit. *Biomass Conversion and Biorefinery*, *6*(2), 233–245. <http://doi.org/10.1007/s13399-015-0180-8>
- Chan, S., Kanchanatawee, S., & Jantama, K. (2012). Production of succinic acid from sucrose and sugarcane molasses by metabolically engineered *Escherichia coli*. *Bioresource Technology*, *103*(1), 329–336. <http://doi.org/10.1016/j.biortech.2011.09.096>

- Coetzee, G. (2019). *Optimization of Aspergillus fijiensis β -fructofuranosidase expression and production using Pichia pastoris, for the production of fructooligosaccharides from sucrose*. Stellenbosch University.
- Dias, M. O. de S., Maciel Filho, R., Mantelatto, P. E., Cavalett, O., Rossell, C. E. V., Bonomi, A., & Leal, M. R. L. V. (2015). Sugarcane processing for ethanol and sugar in Brazil. *Environmental Development*, 15, 35–51. <http://doi.org/10.1016/j.envdev.2015.03.004>
- Dogbe, E. S., Mandegari, M., & Görgens, J. F. (2019). Assessment of the thermodynamic performance improvement of a typical sugar mill through the integration of waste-heat recovery technologies. *Applied Thermal Engineering*, 158, 113768. <http://doi.org/10.1016/j.applthermaleng.2019.113768>
- Dominguez, A. L., Rodrigues, L. R., Lima, N. M., & Teixeira, J. A. (2014). An Overview of the Recent Developments on Fructooligosaccharide Production and Applications. *Food and Bioprocess Technology*, 7(2), 324–337. <http://doi.org/10.1007/s11947-013-1221-6>
- Efe, Ç., van der Wielen, L. A. M., & Straathof, A. J. J. (2013). Techno-economic analysis of succinic acid production using adsorption from fermentation medium. *Biomass and Bioenergy*, 56, 479–492. <http://doi.org/10.1016/j.biombioe.2013.06.002>
- Eggleston, G., & Lima, I. (2015). Sustainability Issues and Opportunities in the Sugar and Sugar-Bioproduct Industries. *Sustainability*, 7(9), 12209–12235. <http://doi.org/10.3390/su70912209>
- Farzad, S., Mandegari, M. A., Guo, M., Haigh, K. F., Shah, N., & Görgens, J. F. (2017). Multi-product biorefineries from lignocelluloses: a pathway to revitalisation of the sugar industry? *Biotechnology for Biofuels*, 10. <http://doi.org/10.1186/s13068-017-0761-9>
- Grand View Research Inc. (2016). *Fructooligosaccharides (FOS) Market Analysis By Source (Inulin, Sucrose) By Application (Food & Beverages, Infant Formula, Dietary Supplements, Animal Feed, Pharmaceuticals) And Segment Forecasts To 2024*. US. Retrieved from <https://www.grandviewresearch.com/industry-analysis/fructooligosaccharides-market>
- Guest, K., Stark, A., & Starzak, M. (2019). Development and verification of an ASPEN Plus ® model of a sugarcane biorefinery. *Proc S Afr Sug Technol Ass*, 92(August), 254–273.
- Hoekstra, R. G. (1995). Energy consequences of diffusion versus milling. In *Proc S Afr Sug*

Technol Ass (pp. 205–207).

Humbird, D., Davis, R., Tao, L., Kinchin, C., Hsu, D., & Aden, A. (2011). *Process Design and Economics for Biochemical Conversion of Lignocellulosic Biomass to Ethanol. Technical Report NREL/TP-5100-47764*. Golden, Colorado.

Jansen, M. La, & van Gulik, W. M. (2014). Towards large scale fermentative production of succinic acid. *Current Opinion in Biotechnology*, 30, 190–197. <http://doi.org/10.1016/j.copbio.2014.07.003>

Jiang, M., Ma, J., Wu, M., Liu, R., Liang, L., Xin, F., ... Dong, W. (2017). Progress of succinic acid production from renewable resources: Metabolic and fermentative strategies. *Bioresource Technology*, 245, 1710–1717. <http://doi.org/10.1016/j.biortech.2017.05.209>

Klein, B. C., Silva, J. F. L., Junqueira, T. L., Rabelo, S. C., Arruda, P. V., Ienczak, J. L., ... Bonomi, A. (2017). Process development and techno-economic analysis of bio-based succinic acid derived from pentoses integrated to a sugarcane biorefinery. *Biofuels, Bioproducts and Biorefining*, 11(6), 1051–1064. <http://doi.org/10.1002/bbb.1813>

Krajnc, D., & Glavič, P. (2009). Assessment of different strategies for the co-production of bioethanol and beet sugar. *Chemical Engineering Research and Design*, 87(9), 1217–1231. <http://doi.org/10.1016/j.cherd.2009.06.014>

Liu, Y. P., Zheng, P., Sun, Z. H., Ni, Y., Dong, J. J., & Zhu, L. L. (2008). Economical succinic acid production from cane molasses by *Actinobacillus succinogenes*. *Bioresource Technology*, 99(6), 1736–1742. <http://doi.org/10.1016/j.biortech.2007.03.044>

Ma, J., Li, F., Liu, R., Liang, L., Ji, Y., Wei, C., ... Ouyang, P. (2014). Succinic acid production from sucrose and molasses by metabolically engineered *E. coli* using a cell surface display system. *Biochemical Engineering Journal*, 91, 240–249. <http://doi.org/10.1016/j.bej.2014.08.014>

Mandegari, M. A., Farzad, S., & Görgens, J. F. (2017). Recent trends on techno-economic assessment (TEA) of sugarcane biorefineries. *Biofuel Research Journal*, 4(3), 704–712. <http://doi.org/10.18331/BRJ2017.4.3.7>

Mandegari, M., Farzad, S., & Görgens, J. F. (2018). A new insight into sugarcane biorefineries with fossil fuel co-combustion: Techno-economic analysis and life cycle assessment. *Energy Conversion and Management*, 165, 76–91.

<http://doi.org/10.1016/j.enconman.2018.03.057>

- Martínez-Guido, S. I., Betzabe González-Campos, J., Ponce-Ortega, J. M., Nápoles-Rivera, F., & El-Halwagi, M. M. (2016). Optimal reconfiguration of a sugar cane industry to yield an integrated biorefinery. *Clean Technologies and Environmental Policy*, *18*(2), 553–562. <http://doi.org/10.1007/s10098-015-1039-1>
- McKinlay, J. B., Vieille, C., & Zeikus, J. G. (2007). Prospects for a bio-based succinate industry. *Applied Microbiology and Biotechnology*, *76*(4), 727–740. <http://doi.org/10.1007/s00253-007-1057-y>
- Mkwananzi, T., Mandegari, M., & Görgens, J. F. (2019). Disturbance modelling through steady-state value deviations: The determination of suitable energy indicators and parameters for energy consumption monitoring in a typical sugar mill. *Energy*, *176*, 211–223. <http://doi.org/10.1016/j.energy.2019.03.191>
- Morales, M., Ataman, M., Badr, S., Linster, S., Kourlimpinis, I., Papadokostantakis, S., Hungerbühler, K. (2016). Sustainability assessment of succinic acid production technologies from biomass using metabolic engineering. *Energy & Environmental Science*, *9*(9), 2794–2805. <http://doi.org/10.1039/C6EE00634E>
- Mussatto, S. I., Aguiar, L. M., Marinha, M. I., Jorge, R. C., & Ferreira, E. C. (2015). Economic analysis and environmental impact assessment of three different fermentation processes for fructooligosaccharides production. *Bioresource Technology*, *198*, 673–681. <http://doi.org/10.1016/j.biortech.2015.09.060>
- Nieder-Heitmann, M., Haigh, K., & Görgens, J. F. (2019). Process design and economic evaluation of integrated, multi-product biorefineries for the co-production of bio-energy, succinic acid, and polyhydroxybutyrate (PHB) from sugarcane bagasse and trash lignocelluloses. *Biofuels, Bioproducts and Biorefining*, *13*(3), 599–617. <http://doi.org/10.1002/bbb.1972>
- OECD/FAO. (2018). Sugar. In *OECD-FAO Agricultural Outlook 2018-2027* (2018th ed., pp. 139–276). Rome: OECD. http://doi.org/10.1787/agr_outlook-2018-en
- Pateraki, C., Almqvist, H., Ladakis, D., Lidén, G., Koutinas, A. A., & Vlysidis, A. (2016). Modelling succinic acid fermentation using a xylose based substrate. *Biochemical Engineering Journal*, *114*, 26–41. <http://doi.org/10.1016/j.bej.2016.06.011>

- Reid, B. M. J., & Rein, P. W. (1983). Steam balance for the new Felixton II mill. In *Proceedings of The South African Sugar Technologists' Association* (pp. 85–91).
- Reid, M. J. (2006). Why do we continue to burn so much coal? In *Proceedings of The South African Sugar Technologists' Association* (Vol. 80, pp. 353–363).
- Rein, P. (2007). *Cane Sugar Engineering*. (A. Bartens, Ed.). Berlin.
- Renó, M. L. G., Olmo, O. A. del, Palacio, J. C. E., Lora, E. E. S., & Venturini, O. J. (2014). Sugarcane biorefineries: Case studies applied to the Brazilian sugar–alcohol industry. *Energy Conversion and Management*, 86, 981–991. <http://doi.org/10.1016/j.enconman.2014.06.031>
- Shin, H. (2004). Production of fructo-oligosaccharides from molasses by *Aureobasidium pullulans* cells. *Bioresource Technology*, 93(1), 59–62. <http://doi.org/10.1016/j.biortech.2003.10.008>
- Soltanian, S., Aghbashlo, M., Farzad, S., Tabatabaei, M., Mandegari, M., & Görgens, J. F. (2019). Exergoeconomic analysis of lactic acid and power cogeneration from sugarcane residues through a biorefinery approach. *Renewable Energy*, 143, 872–889. <http://doi.org/10.1016/j.renene.2019.05.016>
- Soni, M. G., & Tsai, H. (2016). GRAS notification for Fructooligosaccharides. *CFR, 17036*. Retrieved from <https://www.fda.gov/downloads/Food/IngredientsPackagingLabeling/GRAS/NoticeInventory/ucm504609.pdf>
- Starzak, M., & Davis, S. (2016). MATLAB ® modelling of a sugar mill : model development and validation. In *Proc S Afr Sug Technol Ass* (pp. 517–536).
- Thakker, C., Martínez, I., San, K.-Y., & Bennett, G. N. (2012). Succinate production in *Escherichia coli*. *Biotechnology Journal*, 7(2), 213–24. <http://doi.org/10.1002/biot.201100061>
- Ur Rehman, A., Kovacs, Z., Quitmann, H., Ebrahimi, M., & Czermak, P. (2016). Enzymatic production of fructo-oligosaccharides from inexpensive and abundant substrates using a membrane reactor system. *Separation Science and Technology*, 01496395.2016.1167740. <http://doi.org/10.1080/01496395.2016.1167740>
- Vaňková, K., Onderková, Z., Antořová, M., & Polakovič, M. (2008). Design and economics

of industrial production of fructooligosaccharides. *Chemical Papers*, 62(4), 375–381.
<http://doi.org/10.2478/s11696-008-0034-y>

Vaňková, K., & Polakovič, M. (2012). Design of Fructooligosaccharide Separation Using Simulated Moving-Bed Chromatography. *Chemical Engineering & Technology*, 35(1), 161–168. <http://doi.org/10.1002/ceat.201100254>

Vaswani, S. (2010). *Bio-based succinic acid*. Menlo Park, California 94025. Retrieved from https://www.ihs.com/pdf/RW2010-14_220240110917062932.pdf

Ziegler, J. (1978). Sugar Boiling Some Useful Strategies. *Sugar Journal*, 40(8), 8–15. Retrieved from <http://www.zieglerassociates.com/strategy.pdf>

Ziyai, M. R., Mehrpooya, M., Aghbashlo, M., Omid, M., Alsagri, A. S., & Tabatabaei, M. (2019). Techno-economic comparison of three biodiesel production scenarios enhanced by glycerol supercritical water reforming process. *International Journal of Hydrogen Energy*, 44(33), 17845–17862. <http://doi.org/10.1016/j.ijhydene.2019.05.017>

Chapter 6

Aggregated system exergoeconomic methodology for the selection of cost-effective integrated sugarcane biorefineries

Objective of dissertation in this chapter and summary of findings

This chapter was concerned with the selection of a cost-effective and energy-efficient biorefinery for integration into the sugar mill using exergoeconomic analysis (objective 5). In addition to three of the plausible biorefinery scenarios presented in Chapter 5 to valorise A-molasses based on exergy principles, (i.e., scenarios SA-S, S-F_P and SA70-S-F_S for the production of succinic acid (SA), short-chain fructooligosaccharides (scFOS) powder and co-production of SA and scFOS syrup, respectively), one more scenario was simulated to co-utilise C-molasses and lignocellulosic residues as first-and second-generation (1G-2G) feedstock to produce SA. The choice of C-molasses was considered to limit the effect of overproduction on the economics of this scenario (Chapter 5) and to present an alternative scenario to compare the A-molasses based scenarios with. Moreover, waste heat recovery was considered for the supply of chilled water for the biorefinery (Chapter 4).

It has been shown in the literature that exergoeconomic methodology can be used to design and optimize processes for cost-effectiveness by revealing the sources of costs losses of a system that a simple energy/exergy and economic evaluations would not reveal. Moreover, it assesses the costs associated with system inefficiencies and the investment costs needed to reduce them to assist designers with decision making. The methodology is mostly applied to individual components of a system to prioritise them for improvement. In this chapter, the exergoeconomic methodology was applied to the biorefinery scenarios at the aggregated system level as a selection tool to identify the preferred one for implementation by locating the

cost sources and their significance based on the exergoeconomic performance parameters. The scenarios were first analysed using the exergy methodology developed in Chapter 3.

The main contribution of this study is the use of the exergoeconomic methodology for the selection of the most cost-effective process option by quantifying the cost associated with the system inefficiencies relative to capital investment. The key results were that while scenario SA70-S-F_S was preferred based on economics (Chapter 5), scenario S-F_P showed the best exergy (lowest relative irreversibility of 4.8) and exergoeconomic performances with exergoeconomic factor and relative cost difference of 0.56 and 0.19, respectively. Therefore, scenario S-F_P was the preferred biorefinery for a cost-effective integration into the sugar mill because of a good balance between the irreversibility- and investment-related costs. On the other hand, scenario 1G-2G SA-S showed the worst performance in all three analysis. However, the exergoeconomic analysis revealed that the investment cost rate of this scenario was more significant than its irreversibility cost rate and opportunities to reduce capital investment at the expense of efficiency should be pursued to improve the exergoeconomic performance. Thus, exergoeconomics provided more information for a better design of this scenario than the conventional exergy and economic analysis did.

Declaration by the candidate

With regards to Chapter 6, pages 199 - 243, the nature and scope of my contribution were as follows:

Nature of contribution	Extent of contribution (%)
Project and scope definition, Aspen simulation work, analysis of data, interpretation of results and writing of manuscript	80

The following co-authors have contributed to Chapter 6, pages 199 - 243:

Name	e-mail address	Nature of contribution	Extent of contribution (%)
J.F. Görgens	jgorgens@sun.ac.za	Provided writing assistance through review and proof reading of manuscript and general discussion	8
M.A. Mandegari	mandegari@sun.ac.za	Reviewed Aspen simulations, and provided writing assistance through suggestions, review and proof reading of manuscript and general discussions	12

Signature of candidate:

Date:

Declaration by co-authors:

The undersigned hereby confirm that

1. the declaration above accurately reflects the nature and extent of the contributions of the candidate and the co-authors to Chapter 6, pages 199 - 243,
2. no other authors contributed to Chapter 6, pages 199 – 243 besides those specified above, and
3. potential conflicts of interest have been revealed to all interested parties and that the necessary arrangements have been made to use the material in Chapter 6, pages 199 - 243, of this dissertation.

Signature	Institutional affiliation	Date
	Stellenbosch University	
	Stellenbosch University	

Aggregated system exergoeconomic methodology for the selection of cost-effective integrated sugarcane biorefineries

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Abstract

This study presents an aggregated system exergoeconomic methodology that enables the comparison and selection of a preferred biorefinery system for the integration in an existing sugar mill. The study involved to three biorefineries developed in Chapter 5 (scenarios SA-S, S-F_P and SA70-S-F_S) to valorise A-molasses for the production of succinic acid (SA), short-chain fructooligosaccharides (scFOS) powder and co-production of SA and scFOS syrup, respectively. In addition to the three, one more scenario (1G-2G SA-S) was simulation in this chapter to co-utilizing C-molasses and available sugarcane lignocellulose residue as first- and second-generation (1G-2G) feedstocks for the production of SA. Both the economic and exergy performance worst for scenario 1G-2G SA-S with the highest MSP for SA of 2 318 US\$/t and the highest relative irreversibility of 5.9. While scenario SA70-S-F_S was preferred based on economics (Chapter 5), scenario S-F_P showed the best exergy performance (lowest relative irreversibility of 4.8). However, the exergoeconomic evaluation presented scenario S-F_P with the lowest total cost rate of 1029 US\$/h as the preferred choice for a cost-effective integration into the sugar mill. With exergoeconomic factor and relative cost difference of 0.56 and 0.19, respectively, the scenario is almost optimized because of a good balance between the irreversibility- and investment-related costs. The exergoeconomic factor and relative cost difference for scenario 1G-2G SA-S were 0.76 and 0.94, respectively revealing that it requires significant improvement and optimization for a cost-effective implementation compared to the other scenarios. By these results, opportunities to reduce capital investment at the expense of efficiency should be pursued to improve the exergoeconomic performance of the 1G-2G SA-S scenario. Thus, the exergoeconomic methodology provided insightful design and cost information that assisted with the selection of the preferred biorefinery for integration, and can be applied in other studies as a preliminary tool to shortlist a wide range of biorefinery scenarios for implementation.

Keywords: Biorefinery; C-molasses; Exergoeconomic analysis; 1G-2G succinic acid; Lignocellulosic residue; Relative irreversibility.

6.1. Introduction

The sugar industry is an energy-intensive agro-processing industry typically requiring more than 300 kg steam in addition to about 32 kWh electricity per ton of sugar cane processed (Chapter 3). Due to rising energy prices and oversupply of sugar in the international market, it has become necessary to produce sugar in an energy-efficient and cost-efficient way to sustain the industry economically (Somayeh Farzad, Mandegari, Guo, et al., 2017). Traditional sugar mills were designed with energy-supply inefficiencies in order to dispose of excess bagasse, which now is an important feedstock for the emerging bio-based economy. Consequently, opportunities for improving the energy efficiency of existing sugar mills towards cost-effective production are limited while working with the old boiler systems and the improvements achieved are quite marginal (Chapter 4). Alternatively, valorisation of sugar industry by-products for promising high-value products in the sugar mill could refurbish the economics of the industry (Somayeh Farzad, Mandegari, Guo, et al., 2017). The sugarcane biorefinery concept has been the vehicle driving this economic sustainability agenda of the sugar industry. The success is widely demonstrated in the Brazilian and other sugar industries through the co-production of ethanol with sugar (M. O. de S. Dias et al., 2015b; Krajnc & Glavič, 2009). Many other products have been investigated for the sugarcane biorefinery due to the recent global shift toward bio-based chemicals to replace environmentally unfriendly fossil ones (M. Mandegari et al., 2018; Nieder-Heitmann et al., 2019).

Recent research efforts on sugarcane biorefineries have considered techno-economic and environmental feasibilities of various biofuels, biochemicals and electricity (Cavalett et al., 2012; Somayeh Farzad, Mandegari, & Görgens, 2017; Görgens & Mandegari, 2018; Klein et al., 2017; M. Mandegari et al., 2018). These studies showed that energy cost has a strong influence on the economics of sugarcane-based biorefineries (Mkwananzi et al., 2019). Therefore, designing a thermodynamically more efficient biorefinery should be an important consideration for investment decision making and implementation since improving energy efficiency implies lower environmental impact due to reduced energy wastage and better utilization of biomass resources, which reduces input energy cost and consequently improve the economics (Chapter 4). The preferred methodology for the evaluation of the true thermodynamic state of energy systems is exergy analysis rather than the conventional energy

conservation analysis because it reveals the location, magnitude and cause of thermodynamic inefficiencies of energy systems for improvement (Chapter 3). Moreover, assessment of the costs associated with system inefficiencies and the investment costs needed to reduce them provide a more beneficial insight to investors than only identifying the inefficiencies and their locations through exergy analysis, and that is the goal of exergoeconomic analysis, simply referred to as exergoeconomics (George Tsatsaronis, 2007). Therefore, exergoeconomics can be used to evaluate the holistic performance of biorefineries and to identify the preferred one for investment.

Exergoeconomics is the branch of engineering that combines exergy analysis with economic tools to provide useful information for the design and operation of a cost-effective system that is not provided by simple energy or exergy and economic analyses (Bejan, Tsatsaronis, & Moran, 1996; George Tsatsaronis, 2007). The methodology has been used to design, evaluate the performance, and optimize thermal systems in the last decade (Vučković, Stojiljković, Vukić, Stefanović, & Dedeić, 2014) by evaluating the costs associated with system exergy destructions in addition to the investment-related costs (making the total cost rate of the system) and finding cost-effective ways to minimise them (George Tsatsaronis, 2007). Recent exergoeconomic studies have mostly focused on determining the cost rates at the level of individual components of the system (Aghbashlo et al., 2019; Atmaca & Yumrutaş, 2014b). To the best of the authors' knowledge, the only exergoeconomics study on biorefineries was reported by Soltanian et al. (2019) who used the component level exergoeconomic approach to identify the hotspots of cost losses in a sugarcane-based lactic acid and electricity biorefinery. The total cost rate of the steam generation unit was found responsible for 43.73 % of the overall system's total cost rate, while its irreversibility cost rate (more than 70 % of its total cost rate) was 52.50 % of the overall system's irreversibility cost rate as expected. This implies that the irreversibility-related cost was more significant than the capital investment cost of the system and that improvement efforts should be towards minimizing the exergy destruction at the expense of a higher capital investment. Thus, the exergoeconomic evaluation of the system makes the sources of the costs more visible to assist design decisions than a simple energy/exergy and economic evaluations would reveal.

Much as the component evaluation is necessary to understand the cost formation and flow in a particular system, the aggregated system methodology presented in Bejan et al. (1996) may be more robust for determining the holistic performance of different configurations of a system in order to select the most cost-effective one. Therefore, this study applies the exergoeconomic

methodology to evaluate and compare the performance of various biorefinery options for integration into an existing sugar mill, through the aggregated system approach. The methodology is different from other studies because it focuses on evaluating and comparing the total cost rates and exergoeconomic performance of overall systems rather than assessing the performance of individual components of a particular system. Ultimately, the method developed in this study provides holistic design and cost information about an entire system at a glance beyond what simple energy/exergy and economic analyses provide, without a detailed assessment of its components. This method, therefore, serves as a single holistic criteria in a preliminary process to eliminate uneconomical systems or select a preferred one from a pool of options for implementation.

Motivated by energy-saving prospects of single-stage crystallization leading to reduced sugar production but the opportunity to diversify the sugar industry (Chapter 3), various biorefinery scenarios (referred to hereafter as 1G scenarios) were developed in Chapter 5 to valorise A-molasses for the production of short-chain fructooligosaccharides (scFOS) and succinic acid (SA) annexed an existing sugar mill. ScFOS is a product of enzymatic transformation of sucrose (Bedzo, Mandegari, et al., 2019), which have prebiotic properties and many health benefits (Dominguez, Rodrigues, Lima, & Teixeira, 2014b). Rising awareness of personal health and wellness have contributed to increasing consumer demand for scFOS leading to its promising market. On the other hand, succinic acid (SA), produced by bacterial fermentation of plant biomass sugars is one of the top platform chemicals. SA has a wide range applications including high-value niche in the food and beverage industry and large-volume applications such as plasticizers, polyurethanes, resins, coatings and chemical intermediate for the production of bio-butanediol (BDO) (Taylor et al., 2015). These large volume applications of SA, especially for the production of BDO are the main driver for the growing global SA market, estimated at 40 kt/y in 2014 and expected to 600 kt/y by 2020 (Klein et al., 2017).

This chapter presents an aggregated exergoeconomic analysis of three of the biorefinery scenarios developed in Chapter 5, in addition to one more scenario developed that utilises all the final (C-) molasses of the sugar process (instead of A-molasses) in addition to all lignocellulosic residues available, as first- and second-generation (1G-2G) feedstocks to produce succinic acid. This scenario was considered as a way to completely valorise the sugarcane plant in the current configuration of existing sugar mills in a complex all-inclusive sugarcane biorefinery. The four scenarios (three 1G biorefineries plus one 1G-2G SA

biorefinery) are evaluated and compared through the exergoeconomic assessment for the selection of a preferred one for implementation in the South African sugar industry.

6.2. Materials and methods

6.2.1. The first generation (1G) biorefinery scenarios from A-molasses

Three of the integrated biorefineries developed in Chapter 5 for the valorisation of A-molasses are chosen as representative 1G scenarios for this study. These include single-product biorefineries producing SA (scenario SA-S) and scFOS powder (scenario S-F_p), each from all available A-molasses and scenarios SA70-S-F_s, and a multi-product biorefinery producing SA from 70 % of the A-molasses and scFOS syrup (as well as invert sugar by-product) from the remaining 30 % based on their global market sizes and growth rates (Chapter 5).

6.2.2. Combined first- and second-generation (1G-2G) succinic acid biorefinery

The sugarcane-based feedstock for commercial succinic acid biorefinery includes sucrose, but preferably glucose (Vaswani, 2010), which is more expensive. Efe et al. (2013) presented a techno-economic study for SA production cane sugar (sucrose) and recommended the use of a cheaper feedstock such as molasses with a suitable purification technique to improve the economics of the process. On the other hand, the use of another a less-expensive lignocellulosic residue (2G) feedstock presents the challenge of expensive pre-treatment requirement (Nieder-Heitmann et al., 2019). Alternatively, the use of purer molasses, A-molasses (69.8 % sucrose, 15.1 % glucose and 15.1 % fructose on total solids basis and 22 % water) (values from Aspen simulation, not published) for SA production in an integrated biorefinery annexed to an existing sugar mill presented a more profitable SA biorefinery than the 2G biorefinery (Chapter 5).

Another possibility exist to combine molasses and sugarcane hemicellulose residues as 1G-2G feedstock for SA production, which has not been reported in the literature. This scenario was motivated by the possibility of mixing molasses stream with highly dilute sugar streams from pre-treated and hydrolysed hemicellulose for dilution instead of using water. This consideration is important to improve the exergy efficiency of the 1G-2G process compared to a 2G process due to the reduction in exergy losses of evaporation (Chapter 3). Therefore, the 1G-2G SA scenario integrated into the sugar mill presents a complex sugarcane biorefinery that fully valorises the sugarcane plant in an energy-efficient way.

6.2.2.1. First- (1G) and second-generation (2G) feedstocks

The 1G feedstock considered for the 1G-2G SA scenario in this chapter was 9.6 t/h of C-molasses available from the current configuration of the existing sugar mill (Chapter 3) instead of 21.2 t/h of sucrose-rich A-molasses in the 1G-2G scenario was to avoid the overproduction of SA considering the current market size of approximately 440 kt/y (Chapter 5). Besides, the C-molasses comprising of 20.89 % water, 32.44 % sucrose and 46.67 % non-sucrose (i.e., glucose and fructose) on mass basis (Guest et al., 2019) has a higher content of readily fermentable sugars (glucose and fructose) than A-molasses, which contains has more sucrose content (Chapter 5). However, it is worth noting that the use of C-molasses may require a much more thorough SA purification than using A-molasses

The 2G feedstock comprises of the total lignocellulosic feedstock (including bagasse, tops and trash) available from harvesting and processing 250 t/h of sugarcane (Chapter 3). This is a mixture of 77 t/h wet bagasse (Chapter 3) and 19 t/h tops and trash, which represent 50 % of the total harvesting residue, leaving the rest on the fields to condition the soil (Ali Mandegari et al., 2017). The composition of the combined bagasse (Chapter 5) and trash (Petersen, Aneke, & Görgens, 2014) as a 2G feedstock was assumed as 40.4 % cellulose, 24.3% hemicellulose, 24.6 % lignin 7.27 % extractive and 3.4 % ash (dry basis) and total water content of 44.0 %. The hemicellulose fraction was assumed as 88.7 % xylan, 7.3 % arabinan, 3.3 % galactan and 0.7 % mannan carbohydrates.

For the 1G-2G SA scenario, no energy was saved since no reconfiguration of the sugar mill was required to use C-molasses. As a result, a new boiler was necessary for the biorefinery. A small low-pressure boiler was considered for the biorefinery instead of replacing the existing sugar mill boiler with a bigger and more expensive but more efficient one that supplies energy to both the sugar mill and biorefinery (Nieder-Heitmann et al., 2019) because surplus electricity generation for sale was not an option for this study. Therefore, sixty per cent (60%) of the 2G feedstock was used in addition to the 1G feedstock (C-molasses) for 1G-2G SA production. The remaining 40% was shared between the existing sugar mill boiler (in place of only bagasse) and as a supplement to solid residues from cellulose hydrolysis (from 2G feed pre-treatment plant) for the 1G-2G biorefinery boiler to supply heat and power demands. It worth mentioning that the feedstock portion to biorefinery boiler was iterated to ensure the energy self-sufficiency of the integrated plant.

The 1G-2G feedstock was used for the 1G-2G SA process simulated in Aspen Plus software. The processes involved pre-treatment and enzymatic hydrolysis of the lignocelluloses in separate hydrolysis and fermentation (SHF) process where the enzymatic hydrolysate was mixed with C-molasses for the fermentation of hexoses and the pre-treated hydrolysate containing pentoses was fermented separately. The fermentation broths from the separate fermenters were mixed after the removal of cells for further downstream processing to purify the product.

6.2.2.2. Pre-treatment and hydrolysis of lignocellulosic feedstock

Dilute acid pre-treatment was used as one of the most well-established methods for lignocellulose pre-treatment (Pachón et al., 2018). The lignocellulosic feedstock diluted to 30 % solids loading (Humbird et al., 2011) and pre-heated by direct injection of high-pressure (HP) steam to a pre-treatment temperature of 180 °C (Nieder-Heitmann et al., 2019). The feed was fed into the reactor (modelled with RStoic in Aspen) together with 2 % v/v dilute sulphuric acid (93 %). The pre-treated mixture containing xylose-rich liquid hydrolysate, and cellulose and lignin solid fraction in addition to microorganism inhibitors such as furfural and 5-hydroxymethyl furfural (HMF) was flashed and separated in a centrifuge (Humbird et al., 2011). The solid fraction was washed to reduce xylose and inhibitors contents of the solid and diluted to 20 % solid for the enzymatic hydrolysis. The wash water was added to the liquid fraction and detoxified using granular activated carbon column to remove the inhibitors (furfural and HMF) before fermentation (R. Liu et al., 2013; Nieder-Heitmann et al., 2019). The enzymatic hydrolysis reactor was simulated with RStoic reactor operated at 50 °C with a cellulase loading of 20 mg/g cellulose to produce glucose hydrolysate and a lignin-rich solid residue.

6.2.2.3. Succinic acid production.

Different genetically engineered strains of *E. coli* were confirmed to successfully convert sugar mixtures of sucrose, fructose and glucose as well as xylose and arabinose (major pentoses in pre-treatment hydrolysate) to succinic acid with minimal by-products concentrations (Andersson, Hodge, Berglund, & Rova, 2007; R. Liu et al., 2013; H. Wang et al., 2014; J. Wang, Zhu, Bennett, & San, 2011). However, there is a lack of experimental data for the fermentation of mixed sugars consisting of sucrose, xylose, glucose and fructose for the production of succinic acid. Therefore, in this study, the glucose stream from enzymatic hydrolysis of cellulose/lignin solid was co-fermented with C-molasses as hexoses mixture (S.

Chan et al., 2012) separately from the xylose-rich hydrolysate obtained directly from hemicelluloses pre-treatment (containing mainly pentoses) (R. Liu et al., 2013), using genetically engineered *E. coli* strains KJ122-KJSUC-24T and BA204, respectively. The reaction conditions in both fermenters were 37 °C, pH of 7.0 and using NaHCO₃ as CO₂ source and neutralizer. The initial sugars concentrations were 150 g/L and 30 g/L for the fermentation of hexoses and pentoses, respectively. The conversion of sugars to succinic acid were calculated to ensure the succinate titre of 56 g/L at 72 hours in the hexoses fermentation broth (S. Chan et al., 2012) and 18.8 g/L at 48 hours for pentoses (R. Liu et al., 2013), using 10 % of the sugars in each case for microorganism growth. It was assumed that, both *E. coli* strains produced free succinic acid at the defined fermentation conditions to simplify the process (Morales et al., 2016).

6.2.2.4. Downstream processing (DSP)

The first downstream process for the purification of the product succinic acid was the removal of *E. coli* cells and solid impurities from the fermentation broth through filtration (Nieder-Heitmann et al., 2019; Vaswani, 2010). The broths from the hexose and pentose sugars fermenters were filtered separately and the filtrates mixed afterwards for further processing as described in Chapter 5, Section 5.2.2.1.3.

6.2.3. Exergoeconomic methodology

The purpose of the exergoeconomic analysis in this study was to evaluate on aggregated system level, the cost of inefficiencies and the associated investment costs of the 1G-2G SA biorefinery and three 1G biorefinery scenarios (Section 6.2.1) developed in Chapter 5. This analysis was to provide clearer design and cost information about the biorefineries than the outcomes of simple energy/exergy and economic analyses, to assist the selection of a preferred scenario for implementation. The steps necessary in conducting exergoeconomic analysis include defining the fuel and product exergy, followed by setting up the exergy balance from exergy analysis and then, setting up cost balance equations through exergy costing using data from economic analysis (Vučković et al., 2014). Figure 6-1 depict the exergoeconomic methodology for the evaluation of the biorefinery systems in this study.

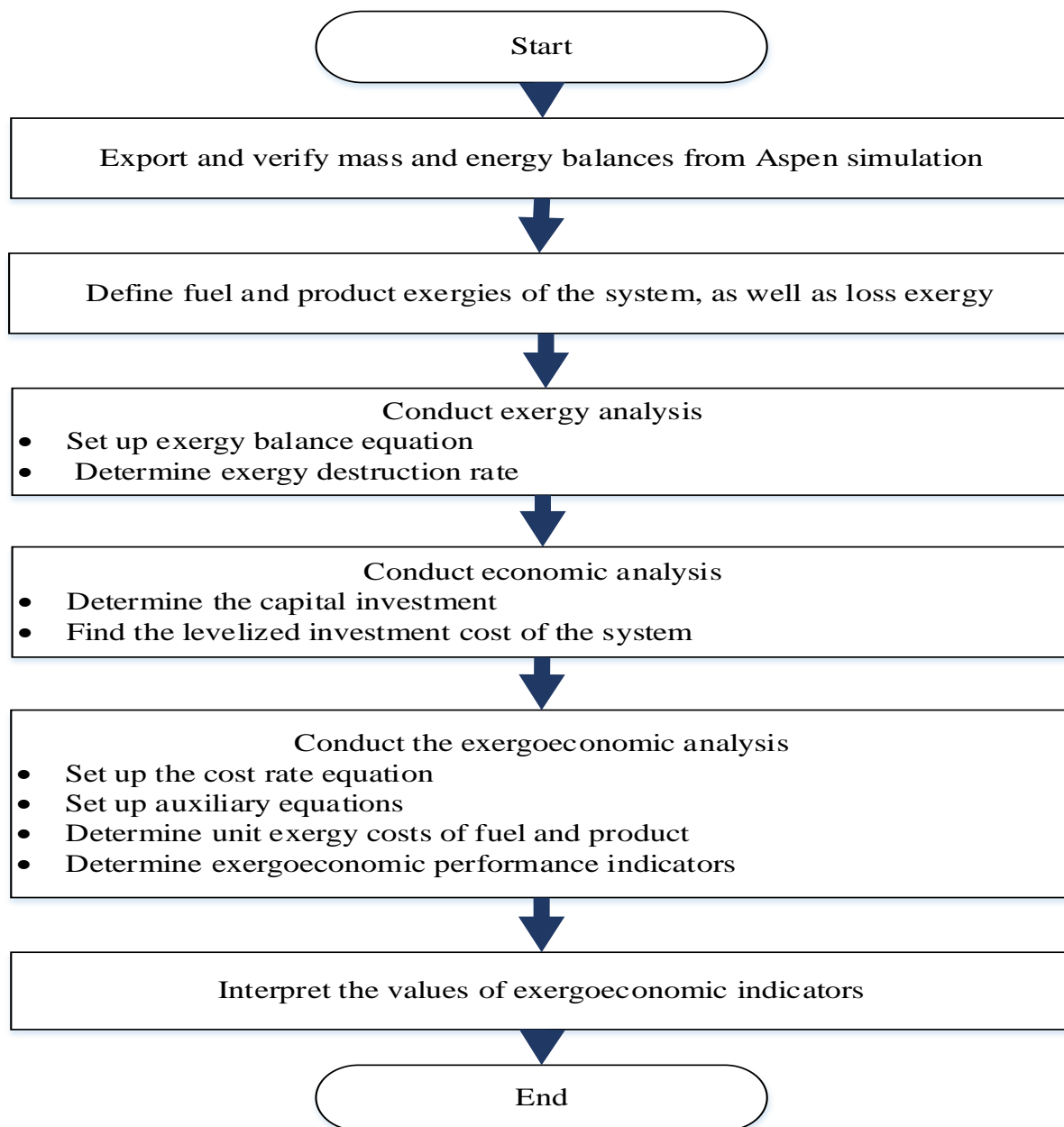


Figure 6-1: The aggregated system exergoeconomic methodology for the biorefineries

6.2.2.5. Fuel and product exergy

The first step of the exergoeconomic methodology was to classify all exergy streams entering and leaving the biorefinery system as fuel, product or loss (waste). The fuel exergy is defined as the sum of all exergy values entering the system or component and exergy decreases of respective materials, minus exergy increases that do not serve the purpose of the system or component. The product exergy is all exergy values exiting a system in addition to all exergy increases that serve the purpose of the system or component (Lazzaretto & Tsatsaronis, 2006).

The exergies of streams that do not appear in the definition of either the product or fuel represent exergy losses of the system. These definitions are based on the specific exergy costing (SPECOC) approach presented by Lazzaretto and Tsatsaronis (2006) as summarised in Table 6-1, with the streams numbers defined on Figure 6-2 to 6-5, which illustrate the schematics of the biorefineries system boundaries. These fuel, product and loss exergy definitions were used in exergy balance equation of the biorefinery systems assessed to determine the system exergy destruction and subsequently, the associated cost.

6.2.2.6. Exergy analysis

The second step of the exergoeconomic methodology was the setting up of exergy balance equation through exergy analysis, to determine the exergy destruction of the systems. The exergy rate of all streams in and out of the biorefineries were determined as described in Chapter 3.

The exergy destruction rate, $\dot{E}x_D$ (the same as irreversibility rate, \dot{I} , in Chapter 3) was evaluated by the exergy balance equation defined with respect to the definitions of fuel, product and loss exergies in this chapter (equation (6-1)),

$$\dot{E}x_F = \dot{E}x_P + \dot{E}x_L + \dot{E}x_D \quad (6-1)$$

where $\dot{E}x_F$, $\dot{E}x_P$ and $\dot{E}x_L$ are the exergy rates of fuel, product and losses, respectively.

The exergy balance equation (6-1) is the basis for the cost balance equation in the exergoeconomic methodology, established through exergy costing.

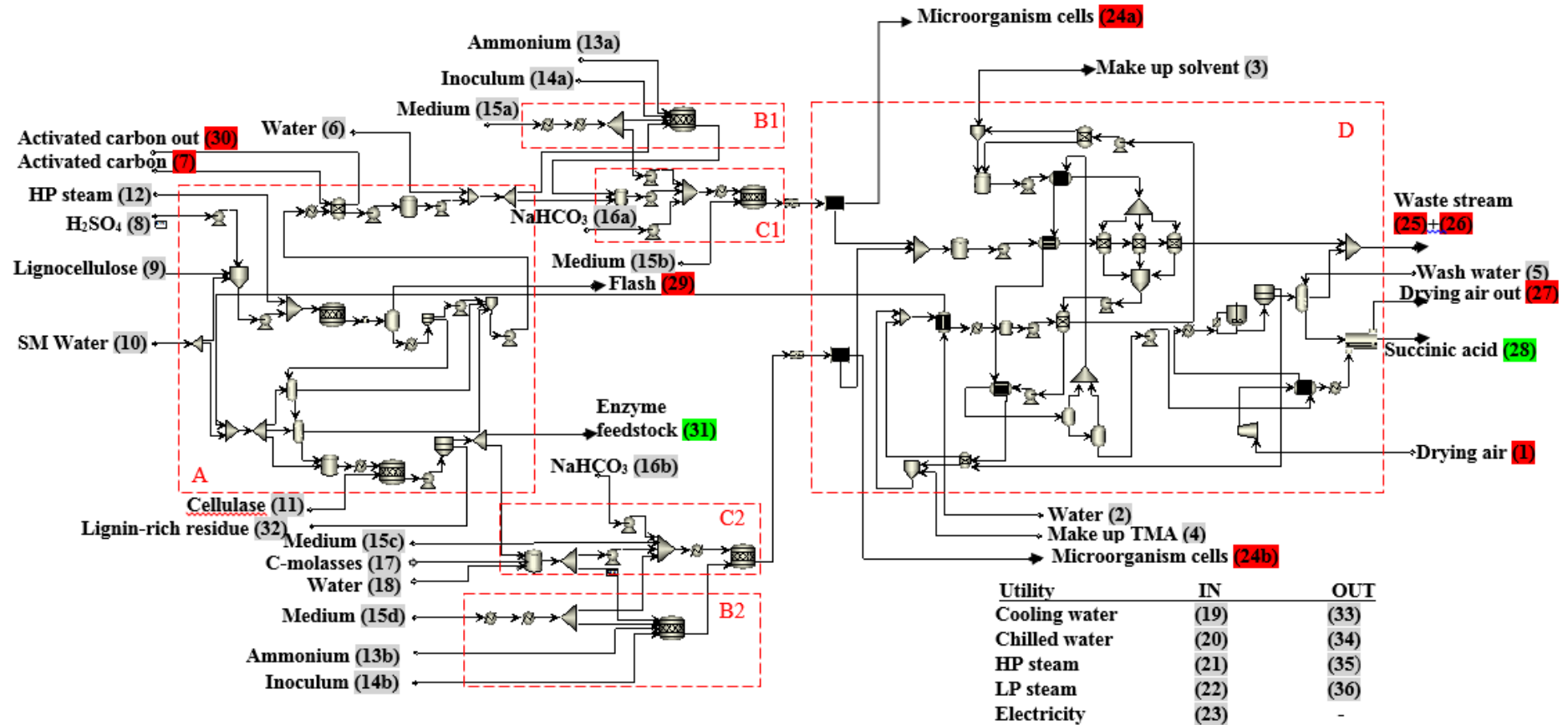


Figure 6-2: Schematic of the multi-product 1G-2G SA biorefinery (scenario 1G-2G SA-S) showing all exergy streams entering and leaving the system. Colour codes: grey - fuel exergy; green - product exergy; red - waste exergy. Plant sections: A - pre-treatment and enzymatic hydrolysis; B1 – pentoses seed train; C1 – pentoses fermentation; B2-hexoses seed train; C2 – hexoses fermentation; D - downstream processing

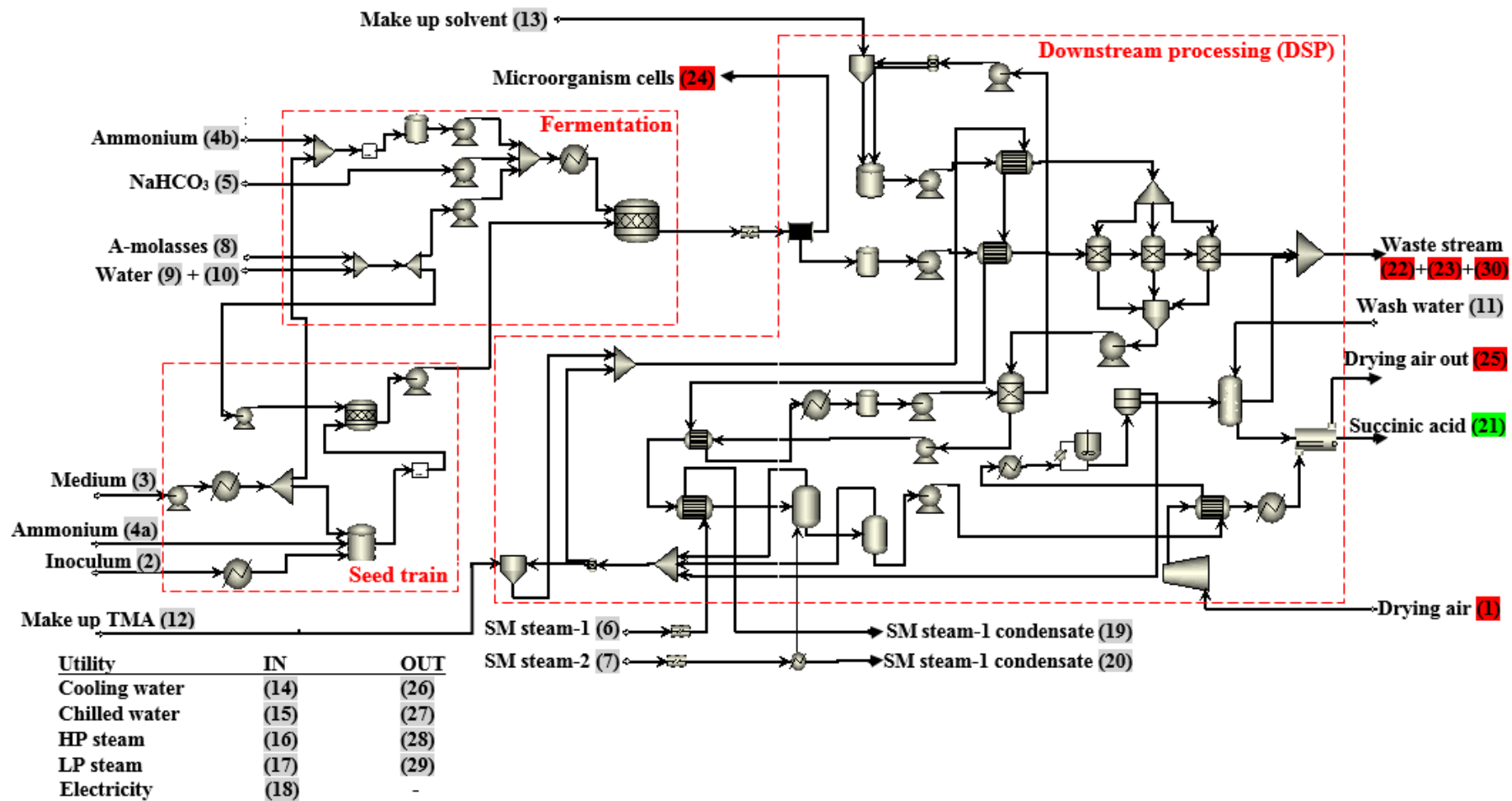


Figure 6-3: Schematic of the 1G SA biorefinery (scenario SA-S) showing all exergy streams entering and leaving the system. Colour codes: grey - fuel exergy; green - product exergy; red - waste exergy.

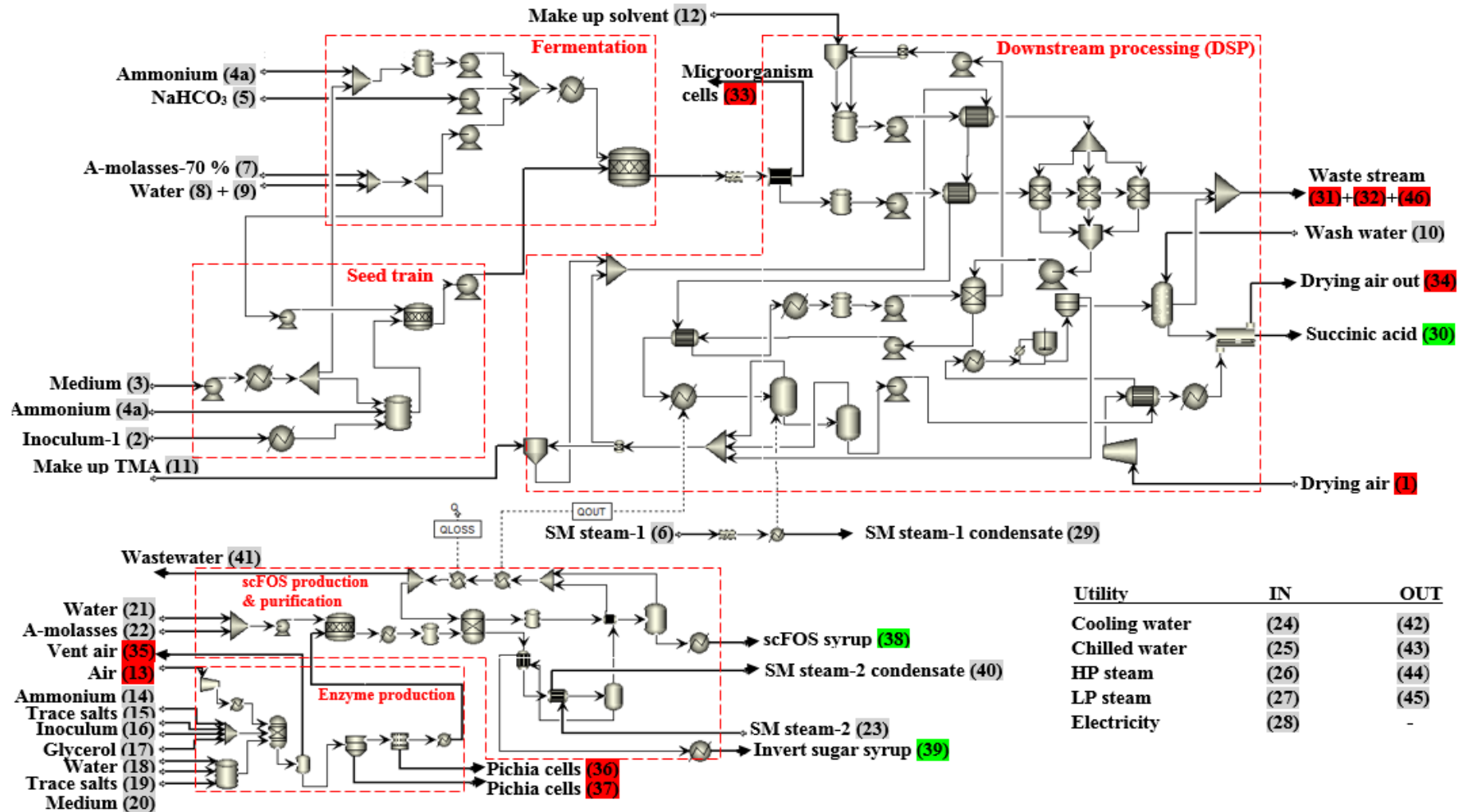


Figure 6-4: Schematic of the multi-product 1G SA (blue black box) and scFOS syrup (yellow black box) biorefinery (scenario SA70-S-Fs) showing all exergy streams entering and leaving the system. Colour codes: grey - fuel exergy; green - product exergy; red - waste exergy.

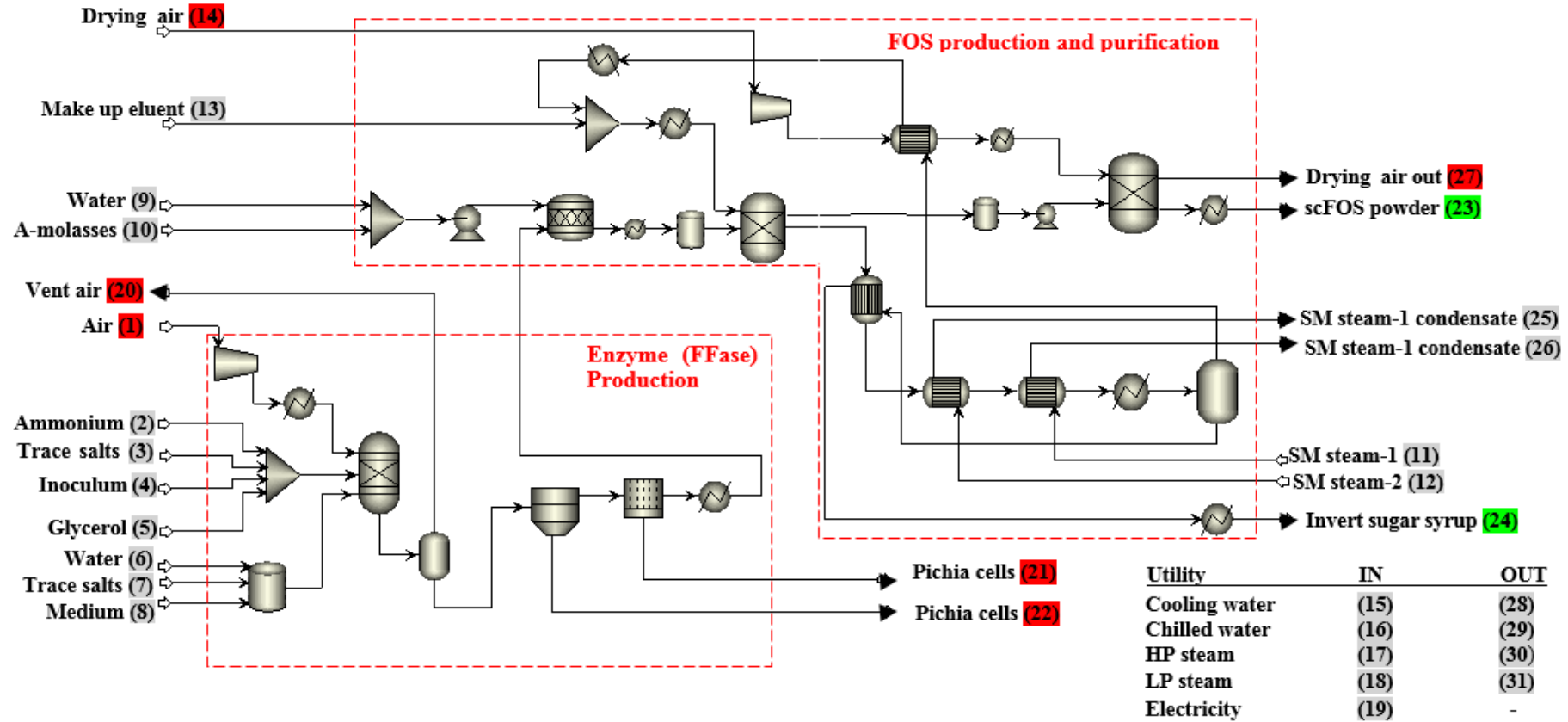


Figure 6-5: Schematic of the scFOS biorefinery (scenario S-F_P) showing all exergy streams entering and leaving the system. Colour codes: grey - fuel exergy; green - product exergy; red - waste exergy.

Table 6-1: Definition of fuel, product and loss exergy of the developed biorefineries

Production system	Fuel	Product	Losses
1G-2G SA-S	$\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + \dot{E}x_6 + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{11} + \dot{E}x_{12} + \dot{E}x_{13} + \dot{E}x_{14} + \dot{E}x_{15} + \dot{E}x_{16} + \dot{E}x_{17} + \dot{E}x_{18} + (\dot{E}x_{19} - \dot{E}x_{33}) + (\dot{E}x_{20} - \dot{E}x_{34}) + (\dot{E}x_{21} - \dot{E}x_{35}) + (\dot{E}x_{22} - \dot{E}x_{36}) + \dot{E}x_{23}$	$\dot{E}x_{28} + \dot{E}x_{31}$	$\dot{E}x_{24} + \dot{E}x_{25} + \dot{E}x_{26} + (\dot{E}x_{27} - \dot{E}x_1) + \dot{E}x_{29} + (\dot{E}x_{30} - \dot{E}x_7) + \dot{E}x_{32}$
SA-S	$\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + (\dot{E}x_6 - \dot{E}x_{19}) + (\dot{E}x_7 - \dot{E}x_{20}) + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{11} + \dot{E}x_{12} + \dot{E}x_{13} + (\dot{E}x_{14} - \dot{E}x_{26}) + (\dot{E}x_{15} - \dot{E}x_{27}) + (\dot{E}x_{16} - \dot{E}x_{28}) + (\dot{E}x_{17} - \dot{E}x_{29}) + \dot{E}x_{18}$	$\dot{E}x_{21}$	$\dot{E}x_{22} + \dot{E}x_{23} + \dot{E}x_{24} + (\dot{E}x_{25} - \dot{E}x_1) + \dot{E}x_{30}$
SA70-S-Fs	$\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + (\dot{E}x_6 - \dot{E}x_{29}) + \dot{E}x_7 + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{11} + \dot{E}x_{12} + \dot{E}x_{14} + \dot{E}x_{15} + \dot{E}x_{16} + \dot{E}x_{17} + \dot{E}x_{18} + \dot{E}x_{19} + \dot{E}x_{20} + \dot{E}x_{21} + \dot{E}x_{22} + (\dot{E}x_{23} - \dot{E}x_{40}) + (\dot{E}x_{24} - \dot{E}x_{42}) + (\dot{E}x_{25} - \dot{E}x_{43}) + (\dot{E}x_{26} - \dot{E}x_{44}) + (\dot{E}x_{27} - \dot{E}x_{45}) + \dot{E}x_{28}$	$\dot{E}x_{30} + \dot{E}x_{38} + \dot{E}x_{39}$	$\dot{E}x_{31} + \dot{E}x_{32} + \dot{E}x_{33} + (\dot{E}x_{34} - \dot{E}x_1) + (\dot{E}x_{35} - \dot{E}x_{13}) + \dot{E}x_{36} + \dot{E}x_{37} + \dot{E}x_{41} + \dot{E}x_{46}$
S-Fp	$\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + \dot{E}x_6 + \dot{E}x_7 + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + (\dot{E}x_{11} - \dot{E}x_{25}) + (\dot{E}x_{12} - \dot{E}x_{26}) + \dot{E}x_{13} + (\dot{E}x_{15} - \dot{E}x_{28}) + (\dot{E}x_{16} - \dot{E}x_{29}) + (\dot{E}x_{17} - \dot{E}x_{30}) + (\dot{E}x_{18} - \dot{E}x_{31}) + \dot{E}x_{19}$	$\dot{E}x_{23} + \dot{E}x_{24}$	$(\dot{E}x_{20} - \dot{E}x_1) + \dot{E}x_{21} + \dot{E}x_{22} + (\dot{E}x_{27} - \dot{E}x_{14})$

6.2.2.7. Exergy costing and cost balance equations

Exergy costing involves assigning costs to exergy streams that flow in and out of a system as well as the exergy destruction of the system and this is the core of the exergoeconomic analysis (Bejan et al., 1996). The cost rate (US\$/h) associated with the fuel, product or waste exergy streams \dot{C} is proportional to the exergy of the stream as expressed by equation (6-2).

$$\dot{C} = c\dot{E}x \quad (6-2)$$

Where c denotes average cost per unit exergy (US\$/GJ), also referred to as unit exergy cost of each stream. Following the definitions of a fuel and product exergies and exergy losses, a system has a cost balance of the form expressed in equation (6-3)

$$\sum_F c_k \dot{E}x_k + \dot{Z} = \sum_P c_k \dot{E}x_k + \sum_L c_k \dot{E}x_k \quad (6-3)$$

where k denote exergy streams entering and leaving the system (including material and energy flows). The unit exergy cost, c of each fuel stream is calculated as the stream price multiplied by its specific exergy rate, while the unit exergy cost of products and losses are evaluated by solving the cost balance equation (6-3). The term \dot{Z} is the cost rate (US\$/h) due to capital investment and the operating and maintenance expenses of the system and it is calculated using equation (6-4).

$$\dot{Z} = \frac{Z \cdot CRF \cdot \emptyset}{H} \quad (6-4)$$

Where Z , CRF , \emptyset and H are the capital cost of the system (US\$), capital recovery factor, maintenance factor and annual operation hours (h), respectively. The capital recovery factor is determined by equation (6-5).

$$CRF = \frac{i(i+1)^N}{(i+1)^N - 1} \quad (6-5)$$

where i and N are interest rate and plant life (years), respectively. The interest rate, i and the maintenance factor, \emptyset are given as 6.75 % and 1.06 (Soltanian et al., 2019). A plant life of 20 years was considered for the biorefineries, operating at the same annual operational hours (5000 h/y) as the sugar mill (Chapter 5).

For uniformity, the system boundaries of each biorefinery scenario were limited to the inside battery limit (IBL) to focus on the exergoeconomic performance of the production processes considering their integration points in the sugar mill, rather than the auxiliary utility units. The

capital investments of the inside battery limit (IBL) for the previously developed biorefinery scenarios SA-S, SA70-S-F_S and S-F_P were evaluated as 36.34, 33.52 and 29.56 MUS\$, respectively (Chapter 5). The IBL capital investment of 1G-2G SA scenario was determined from a techno-economic assessment as 154 MUS\$.

To solve the cost balance equation (equation (6-3)) for each biorefinery presented in Tables 6-2 and 6-3, auxiliary costing equations based on the F-rule and P-rule of the SPECO approach for the biorefinery systems were formulated and also reported in presented in Tables 6-2 and 6-3. The F rule states that the cost associated with the removal of unit exergy from a fuel stream is equal to the average cost at which the exergy was supplied to the upstream of the same stream (Lazzaretto & Tsatsaronis, 2006). For example, the unit exergy cost, c of steam entering a system is equal to that of its condensate leaving the system. When no exergy difference is considered in the fuel definition, no auxiliary costing equation is required for the F rule (Lazzaretto & Tsatsaronis, 2006). On the other hand, the P rule states that unit exergy is supplied to all product streams at the same average cost, c_p .

Other auxiliary equations necessary to solve the cost balance equation relate to exergy losses of each system. Exergy losses are costed by assigning to them the unit exergy cost of the fuel supplied to the system (Soltanian et al., 2019) as expressed in equation (6-6), assuming the loss streams will be reused by the system (Bejan et al., 1996).

$$\dot{C}_L = c_F \dot{E}x_L \quad (6-6)$$

Where the parameters c_F , \dot{C}_L and $\dot{E}x_L$ are the unit cost of fuel exergy of the system, cost rate and exergy rate of losses, respectively of each biorefinery system. Otherwise, the cost rate of exergy loss (last term in equation (6-3)) is set to zero as losses to the environment, where the product bears the full burden of the system cost (Bejan et al., 1996). Considering the nature of streams defined as losses in this study according to the F and P rule and their possible use in other applications such as biogas production and waste heat recovery applications, the former approach is applied for the overall system analysis as presented by equations (6-7) to (6-10).

Table 6-2: Cost balance and auxiliary equations for exergoeconomic analysis of scenarios 1G-2G SA-S and SA-S

Production system	Cost Balance	F rule	P rule
1G-2G SA-S	$c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + c_6\dot{E}x_6 + c_8\dot{E}x_8 + c_9\dot{E}x_9 + c_{10}\dot{E}x_{10} + c_{11}\dot{E}x_{11} +$ $c_{12}\dot{E}x_{12} + c_{13}\dot{E}x_{13} + c_{14}\dot{E}x_{14} + c_{15}\dot{E}x_{15} + c_{16}\dot{E}x_{16} + c_{17}\dot{E}x_{17} + c_{18}\dot{E}x_{18} +$ $(c_{19}\dot{E}x_{19} - c_{33}\dot{E}x_{33}) + (c_{20}\dot{E}x_{20} - c_{34}\dot{E}x_{34}) + (c_{21}\dot{E}x_{21} - c_{35}\dot{E}x_{35}) + (c_{22}\dot{E}x_{22} -$ $c_{36}\dot{E}x_{36}) + c_{23}\dot{E}x_{23} + \dot{Z}_{1G-2G SA} = c_{28}\dot{E}x_{28} + c_{31}\dot{E}x_{31} + c_{32}\dot{E}x_{32} + c_{24}\dot{E}x_{24} + c_{25}\dot{E}x_{25} +$ $c_{26}\dot{E}x_{26} + (c_{27}\dot{E}x_{27} - c_1\dot{E}x_1) + c_{29}\dot{E}x_{29} + (c_{30}\dot{E}x_{30} - c_7\dot{E}x_7)$	$c_{19} = c_{33}$ $c_{20} = c_{34}$ $c_{21} = c_{35}$ $c_{22} = c_{36}$	$c_{28} = c_{31}$
SA-S	$c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + (c_6\dot{E}x_6 - c_{19}\dot{E}x_{19}) + (c_7\dot{E}x_7 - c_{20}\dot{E}x_{20}) + c_8\dot{E}x_8 +$ $c_9\dot{E}x_9 + c_{10}\dot{E}x_{10} + c_{11}\dot{E}x_{11} + c_{12}\dot{E}x_{12} + c_{13}\dot{E}x_{13} + (c_{14}\dot{E}x_{14} - c_{26}\dot{E}x_{26}) +$ $(c_{15}\dot{E}x_{15} - c_{27}\dot{E}x_{27}) + (c_{16}\dot{E}x_{16} - c_{28}\dot{E}x_{28}) + (c_{17}\dot{E}x_{17} - c_{29}\dot{E}x_{29}) + c_{18}\dot{E}x_{18} + \dot{Z}_{SA} =$ $c_{21}\dot{E}x_{21} + c_{22}\dot{E}x_{22} + c_{23}\dot{E}x_{23} + c_{24}\dot{E}x_{24} + (c_{25}\dot{E}x_{25} - c_1\dot{E}x_1) + c_{30}\dot{E}x_{30}$	$c_6 = c_{19}$ $c_7 = c_{20}$ $c_{14} = c_{26}$ $c_{15} = c_{27}$ $c_{16} = c_{28}$ $c_{17} = c_{29}$	-

Table 6-3: Cost balance and auxiliary equations for exergoeconomic analysis of scenarios SA70-S-Fs and S-Fp

SA70-S-Fs	$ \begin{aligned} & c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + (c_6\dot{E}x_6 - c_{29}\dot{E}x_{29}) + c_7\dot{E}x_7 + c_8\dot{E}x_8 + c_9\dot{E}x_9 + \\ & c_{10}\dot{E}x_{10} + c_{11}\dot{E}x_{11} + c_{12}\dot{E}x_{12} + c_{14}\dot{E}x_{14} + c_{15}\dot{E}x_{15} + c_{16}\dot{E}x_{16} + c_{17}\dot{E}x_{17} + c_{18}\dot{E}x_{18} + \\ & c_{19}\dot{E}x_{19} + c_{20}\dot{E}x_{20} + c_{21}\dot{E}x_{21} + c_{22}\dot{E}x_{22} + (c_{23}\dot{E}x_{23} - c_{40}\dot{E}x_{40}) + (c_{24}\dot{E}x_{24} - c_{42}\dot{E}x_{42}) + \\ & (c_{25}\dot{E}x_{25} - c_{43}\dot{E}x_{43}) + (c_{26}\dot{E}x_{26} - c_{44}\dot{E}x_{44}) + (c_{27}\dot{E}x_{27} - c_{45}\dot{E}x_{45}) + c_{28}\dot{E}x_{28} + \\ & \dot{Z}_{SA+scFOS(s)} = c_{30}\dot{E}x_{30} + c_{38}\dot{E}x_{38} + c_{39}\dot{E}x_{39} + c_{31}\dot{E}x_{31} + c_{32}\dot{E}x_{32} + c_{33}\dot{E}x_{33} + \\ & (c_{34}\dot{E}x_{34} - c_1\dot{E}x_1) + (c_{35}\dot{E}x_{35} - c_{13}\dot{E}x_{13}) + c_{36}\dot{E}x_{36} + c_{37}\dot{E}x_{37} + c_{41}\dot{E}x_{41} + c_{46}\dot{E}x_{46} \end{aligned} $	$ \begin{aligned} c_6 &= c_{29} \\ c_{23} &= c_{40} \\ c_{24} &= c_{42} \\ c_{25} &= c_{43} \\ c_{26} &= c_{44} \\ c_{27} &= c_{45} \end{aligned} $	$c_{30} = c_{38} = c_{39}$
S-Fp	$ \begin{aligned} & c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + c_6\dot{E}x_6 + c_7\dot{E}x_7 + c_8\dot{E}x_8 + c_9\dot{E}x_9 + c_{10}\dot{E}x_{10} + \\ & (c_{11}\dot{E}x_{11} - c_{25}\dot{E}x_{25}) + (c_{12}\dot{E}x_{12} - c_{26}\dot{E}x_{26}) + c_{213}\dot{E}x_{13} + (c_{15}\dot{E}x_{15} - c_{28}\dot{E}x_{28}) + \\ & (c_{16}\dot{E}x_{16} - c_{29}\dot{E}x_{29}) + (c_{17}\dot{E}x_{17} - c_{30}\dot{E}x_{30}) + (c_{18}\dot{E}x_{18} - c_{31}\dot{E}x_{31}) + c_{19}\dot{E}x_{19} + \\ & \dot{Z}_{scFOS(p)} = c_{23}\dot{E}x_{23} + c_{24}\dot{E}x_{24} + (c_{20}\dot{E}x_{20} - c_1\dot{E}x_1) + c_{21}\dot{E}x_{21} + c_{22}\dot{E}x_{22} + \\ & (c_{27}\dot{E}x_{27} - c_{14}\dot{E}x_{14}) \end{aligned} $	$ \begin{aligned} c_{11} &= c_{25} \\ c_{12} &= c_{26} \\ c_{15} &= c_{28} \\ c_{16} &= c_{29} \\ c_{17} &= c_{30} \\ c_{18} &= c_{31} \end{aligned} $	$c_{23} = c_{24}$

Auxiliary cost equation for exergy loss in overall 1G-2G SA plant

$$\begin{aligned}
 c_{24} = c_{25} = c_{26} &= \frac{(c_{27}\dot{E}x_{27} - c_1\dot{E}x_1)}{\dot{E}x_{27} - \dot{E}x_1} = c_{29} = \frac{(c_{30}\dot{E}x_{30} - c_7\dot{E}x_7)}{\dot{E}x_{30} - \dot{E}x_7} = c_{32} = c_F \\
 &= \frac{c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + c_6\dot{E}x_6 + c_8\dot{E}x_8 + c_9\dot{E}x_9 + c_{10}\dot{E}x_{10} + c_{11}\dot{E}x_{11} + c_{12}\dot{E}x_{12} + c_{13}\dot{E}x_{13} + c_{14}\dot{E}x_{14} + c_{15}\dot{E}x_{15} + c_{16}\dot{E}x_{16} + c_{17}\dot{E}x_{17} + c_{18}\dot{E}x_{18} + (c_{19}\dot{E}x_{19} - c_{33}\dot{E}x_{33}) + (c_{20}\dot{E}x_{20} - c_{34}\dot{E}x_{34}) + (c_{21}\dot{E}x_{21} - c_{35}\dot{E}x_{35}) + (c_{22}\dot{E}x_{22} - c_{36}\dot{E}x_{36}) + c_{23}\dot{E}x_{23}}{\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + \dot{E}x_6 + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{11} + \dot{E}x_{12} + \dot{E}x_{13} + \dot{E}x_{14} + \dot{E}x_{15} + \dot{E}x_{16} + \dot{E}x_{17} + \dot{E}x_{18} + (\dot{E}x_{19} - \dot{E}x_{33}) + (\dot{E}x_{20} - \dot{E}x_{34}) + (\dot{E}x_{21} - \dot{E}x_{35}) + (\dot{E}x_{22} - \dot{E}x_{36}) + \dot{E}x_{23}}
 \end{aligned} \tag{6-7}$$

Auxiliary cost equation for exergy loss in overall SA plant

$$\begin{aligned}
 c_{22} = c_{23} = c_{24} &= \frac{(c_{25}\dot{E}x_{25} - c_1\dot{E}x_1)}{\dot{E}x_{25} - \dot{E}x_1} = c_{30} = c_F \\
 &= \frac{c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + (c_6\dot{E}x_6 - c_{19}\dot{E}x_{19}) + (c_7\dot{E}x_7 - c_{20}\dot{E}x_{20}) + c_8\dot{E}x_8 + c_9\dot{E}x_9 + c_{10}\dot{E}x_{10} + c_{11}\dot{E}x_{11} + c_{12}\dot{E}x_{12} + c_{13}\dot{E}x_{13} + (c_{14}\dot{E}x_{14} - c_{26}\dot{E}x_{26}) + (c_{15}\dot{E}x_{15} - c_{27}\dot{E}x_{27}) + (c_{16}\dot{E}x_{16} - c_{28}\dot{E}x_{28}) + (c_{17}\dot{E}x_{17} - c_{29}\dot{E}x_{29}) + c_{18}\dot{E}x_{18}}{\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + (\dot{E}x_6 - \dot{E}x_{19}) + (\dot{E}x_7 - \dot{E}x_{20}) + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{11} + \dot{E}x_{12} + \dot{E}x_{13} + (\dot{E}x_{14} - \dot{E}x_{26}) + (\dot{E}x_{15} - \dot{E}x_{27}) + (\dot{E}x_{16} - \dot{E}x_{28}) + (\dot{E}x_{17} - \dot{E}x_{29}) + \dot{E}x_{18}}
 \end{aligned} \tag{6-8}$$

Auxiliary cost equation for exergy loss in overall SA + scFOS syrup plant

$$\begin{aligned}
 c_{31} = c_{32} = c_{33} &= \frac{(c_{34}\dot{E}x_{34} - c_1\dot{E}x_1)}{\dot{E}x_{34} - \dot{E}x_1} = \frac{(c_{35}\dot{E}x_{35} - c_{13}\dot{E}x_{13})}{\dot{E}x_{35} - \dot{E}x_{13}} = c_{36} = c_{37} = c_{41} = c_{46} = c_F \\
 &= \frac{c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + (c_6\dot{E}x_6 - c_{29}\dot{E}x_{29}) + c_7\dot{E}x_7 + c_8\dot{E}x_8 + c_9\dot{E}x_9 + c_{10}\dot{E}x_{10} + c_{11}\dot{E}x_{11} + c_{12}\dot{E}x_{12} + c_{14}\dot{E}x_{14} + c_{15}\dot{E}x_{15} + c_{16}\dot{E}x_{16} + c_{17}\dot{E}x_{17} + c_{18}\dot{E}x_{18} + c_{19}\dot{E}x_{19} + c_{20}\dot{E}x_{20} + c_{21}\dot{E}x_{21} + c_{22}\dot{E}x_{22} + (c_{23}\dot{E}x_{23} - c_{40}\dot{E}x_{40}) + (c_{24}\dot{E}x_{24} - c_{42}\dot{E}x_{42}) + (c_{25}\dot{E}x_{25} - c_{43}\dot{E}x_{43}) + (c_{26}\dot{E}x_{26} - c_{44}\dot{E}x_{44}) + (c_{27}\dot{E}x_{27} - c_{45}\dot{E}x_{45}) + c_{28}\dot{E}x_{28}}{\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + (\dot{E}x_6 - \dot{E}x_{29}) + \dot{E}x_7 + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + \dot{E}x_{11} + \dot{E}x_{12} + \dot{E}x_{14} + \dot{E}x_{15} + \dot{E}x_{16} + \dot{E}x_{17} + \dot{E}x_{18} + \dot{E}x_{19} + \dot{E}x_{20} + \dot{E}x_{21} + \dot{E}x_{22} + (\dot{E}x_{23} - \dot{E}x_{40}) + (\dot{E}x_{24} - \dot{E}x_{42}) + (\dot{E}x_{25} - \dot{E}x_{43}) + (\dot{E}x_{26} - \dot{E}x_{44}) + (\dot{E}x_{27} - \dot{E}x_{45}) + \dot{E}x_{28}}
 \end{aligned} \tag{6-9}$$

Auxiliary cost equation for exergy loss in overall scFOS plant

$$\begin{aligned}
 \frac{(c_{20}\dot{E}x_{20} - c_1\dot{E}x_1)}{\dot{E}x_{20} - \dot{E}x_1} &= c_{21} = c_{22} = \frac{(c_{27}\dot{E}x_{27} - c_{14}\dot{E}x_{14})}{\dot{E}x_{27} - \dot{E}x_{14}} = c_F \\
 &= \frac{c_2\dot{E}x_2 + c_3\dot{E}x_3 + c_4\dot{E}x_4 + c_5\dot{E}x_5 + c_6\dot{E}x_6 + c_7\dot{E}x_7 + c_8\dot{E}x_8 + c_9\dot{E}x_9 + c_{10}\dot{E}x_{10} + (c_{11}\dot{E}x_{11} - c_{25}\dot{E}x_{25}) + (c_{12}\dot{E}x_{12} - c_{26}\dot{E}x_{26}) + c_{213}\dot{E}x_{13} + (c_{15}\dot{E}x_{15} - c_{28}\dot{E}x_{28}) + (c_{16}\dot{E}x_{16} - c_{29}\dot{E}x_{29}) + (c_{17}\dot{E}x_{17} - c_{30}\dot{E}x_{30}) + (c_{18}\dot{E}x_{18} - c_{31}\dot{E}x_{31}) + c_{19}\dot{E}x_{19}}{\dot{E}x_2 + \dot{E}x_3 + \dot{E}x_4 + \dot{E}x_5 + \dot{E}x_6 + \dot{E}x_7 + \dot{E}x_8 + \dot{E}x_9 + \dot{E}x_{10} + (\dot{E}x_{11} - \dot{E}x_{25}) + (\dot{E}x_{12} - \dot{E}x_{26}) + \dot{E}x_{13} + (\dot{E}x_{15} - \dot{E}x_{28}) + (\dot{E}x_{16} - \dot{E}x_{29}) + (\dot{E}x_{17} - \dot{E}x_{30}) + (\dot{E}x_{18} - \dot{E}x_{31}) + \dot{E}x_{19}}
 \end{aligned} \tag{6-10}$$

To solve the costing equations in Tables 6-2 and 6-3, the unit exergy costs of all entering streams were evaluated using their prices. The unit exergy cost of air supply was considered zero (Bejan et al., 1996; Soltanian et al., 2019) as it has no exergy value. Moreover, it was assumed that the biorefinery provides hydrolysate as feedstock for an independent near-site enzyme production unit in exchange for cellulase enzyme at a 50 % discount price considering that the carbon source for enzyme production makes up 57 % of the enzyme cost (Humbird et al., 2011). Furthermore, the unit exergy costs of cooling water, chilled water, steam and electricity were determined based on the utility usage and considering the process integration with the sugar mill. Using the electricity price of 0.08 US\$/kWh (Dogbe et al., 2019), the average cost per exergy of electricity was calculated and the same assigned to steam as both are products from the same cogeneration system (Bejan et al., 1996; Soltanian et al., 2019). On the other hand, the steam and electricity for the A-molasses-based scenarios provided by an existing cogeneration system of the sugar mill due to the integration; were paid for by charging the cost of surplus bagasse added to the existing boiler as an extra operation cost to the biorefinery (Chapter 5) and accounted for in the pricing of A-molasses. Therefore, no price was assigned to them.

6.2.2.8. Exergoeconomic indicators for system evaluation

The indication of exergoeconomic performance of a system or component is usually guided by three parameters: the irreversibility cost rate, the exergoeconomic factor and the relative cost difference (Atmaca & Yumrutaş, 2014a). As irreversibility indicates inefficiency of a system, knowing its cost is important in improving the system for a cost-effective production (Bejan et al., 1996). The irreversibility cost rate, \dot{C}_D is calculated using the cost rate equation (6-2). It is determined by assigning the unit exergy cost of the fuel to the irreversibility as expressed by Equation (6-11). This parameter reveals the hidden cost associated with the exergy destruction of the process.

$$\dot{C}_D = c_F \dot{E}x_D \quad (6-11)$$

The variables, c_F and $\dot{E}x_D$ are the unit exergy cost of fuel and irreversibility of the system, respectively. The irreversibility cost rate forms part of the total cost rate of a system, which includes the capital investment cost rate. A high irreversibility cost rate and hence a high total cost rate indicates a high priority of a system for improvement (Ozbilen, Dincer, & Rosen, 2016). However, the manner of improvement depends on which component of the cost rate is significant and is signified by another parameter called the exergoeconomic factor.

The exergoeconomic factor expressed by Equation (6-12) indicates the contribution of the investment cost rate to the total (investment and irreversibility) cost rate of the system. A high f value indicates

that investment cost is dominant and should be reduced to improve the system at the expense of irreversibility, and vice versa.

$$f = \frac{\dot{Z}}{\dot{Z} + \dot{C}_D} \quad (6-12)$$

Finally, the relative cost difference shows the relative increase in the average cost per exergy unit between the fuel and product of the system, expressed by Equation (6-13)

$$r = \frac{c_P - c_F}{c_F} \quad (6-13)$$

where c_p is the unit exergy cost of products. This parameter is important for optimizing the system. A high r value indicates a high potential to reduce the unit exergy cost of the product with little effort and vice versa (Aghbashlo et al., 2019). The improvement is achieved by either reducing the irreversibility or capital investment, depending on which is dominant.

The exergoeconomic indicators were evaluated for all four biorefineries considered in this study to assess their actual cost-effectiveness. The results and discussions are presented in the subsequent section.

6.3. Results and discussions

6.3.1. Techno-economic analysis

The economic analysis results of the scenario 1G-2G SA-S compared to the results determined for the A-molasses-based scenarios in Chapter 5 are summarised in Table 6-4. The production rate of succinic acid from the simulation of the scenario 1G-2G SA-S was 18.6 t/h, which is 7.0 t/h higher than scenario SA-S (Chapter 5). The total capital investment and operation costs, 340 MUS\$ and 65 MUS\$, respectively are the highest. The capital investment and annual operation costing results are presented in Table A.6-1 and Table A.6-2 of the chapter Appendix, respectively. Given the prices of trash (53.95 US\$/t), bagasse (26 US\$/t), C-molasses (192 US\$/t) and SA (2500 US\$/t) (Ali Mandegari et al., 2017; Dogbe et al., 2019; OECD/FAO, 2016), and the discount and income tax rates of 9.7 % and 28 %, the profitability of the scenario was determined. The IRR and MSP of SA is lower (22.42 vs. 57.1 % and 2 318 vs. 1 348 US\$/t) compared to scenario SA-S producing 11.6 t/h of succinic acid from A-molasses (Chapter 5). This is largely attributed to the cost of a new cogeneration system for the 1G-2G SA biorefinery since it was not possible to share the existing cogeneration unit with the sugar mill due to the plant size, no surplus steam from the sugar mill and additional energy intensity of the lignocellulose pre-treatment. The other principal reason for higher MSP of 1G-2G SA scenario is the contribution of larger fermenter sizes required for hexose fermentation due to low

sugars concentration of 30.0 g/L, compared to 150 g/L for hexoses fermentation (Section 6.2.2.3). Contrary to what was expected, 1G-2G SA scenario also performs worse than a 13.5 t/h stand-alone 2G SA biorefinery reported in the literature with IRR of 22.4 vs. 36.4 % (Nieder-Heitmann et al., 2019). This is attributed to the production of saleable electricity in the 2G scenario, which provided extra revenue even though a more expensive boiler was used for this purpose (Nieder-Heitmann et al., 2019). Therefore, where a new boiler is required for biorefinery (the case of scenario 1G-2G SA) and electricity production for sale is viable and is a priority (which is not the case currently for South African sugar mills), it will be more economical to replace the existing sugar mill boiler with a more efficient one to supply utilities to both the sugar mill and biorefinery rather than using two separate inefficient boiler (the case of 1G-2G SA scenario). Overall, the scFOS scenario S-F_P present the best economic performance with IRR of 61.1 % compared to other scenarios considering the selling prices assumed for the study, which implies that the scFOS is cheaper to produce than succinic acid.

Table 6-4: Production and economic results of the integrated biorefinery scenarios

	Production rate of SA	TCI	OPEX	NPV	IRR	Payback	SA MSP @ 20 % IRR
	t/h	MUS\$	MUS\$	MUS\$	%	years	US\$/t
1G-2G SA-S	18.6	340	65	381	22.4	4.43	2 318
SA-S	11.6	81	40	422	57.1	1.7	1 348
SA70-S-F_S	8.1	75	36	384	56.6	1.7	1 222
S-F_P	-	65	28	367	61.1	1.8	-

One important parameter from the economic analysis which is necessary for the exergoeconomic methodology used in this study is the installed equipment cost of the inside battery limit (IBL) of the biorefineries. Due to the size of the plant, increased number of process units and the use of separate fermenters, scenario 1G-2G SA-S has the highest IBL of 154 MUS\$ compared to 37 MUS\$, 34 MUS\$ and 30 MUS\$ for scenarios SA-S, SA70-S-F_S and S-F_P, respectively.

6.3.2. Exergy analysis

The true thermodynamic performance of the biorefineries developed is also compared through exergy analysis results presented in Table 6-4. The comparison of exergy performance of the biorefineries (in addition to economic performance) enhances the understanding of the exergoeconomic assessments in this study. The 1G-2G SA-S scenario has the highest irreversibility (22.7 MW) and lowest exergetic efficiency (31.7 %) showing the worst exergy performance. No exergy studies of SA biorefineries were reported in the literature to compare these results to. However, a higher exergetic efficiency (44.73 %) was reported for a 2G biorefinery annexed a sugar mill producing

lactic acid and electricity (Aghbashlo, Mandegari, et al., 2018a), probably because it produced saleable electricity, which is a high exergy product, electricity. The comparison is however not entirely fair since lactic acid production and purification processes are different from SA. In general, the low exergy performance of the 1G-2G SA scenario is attributed to the complexity of the process involving more units (due to the additional fermentation and pre-treatment units) compared to the other biorefineries (Chapter 3). Moreover, the lignin-rich unhydrolysed stream from the pre-treatment unit that was defined as waste for this scenario, is highly exergetic and contributed to the low exergetic efficiency, even though this waste stream was used outside the system boundary as a boiler feed, whereas in the lactic acid scenario reported, the boiler section was inside the boundary analysed (Aghbashlo, Mandegari, et al., 2018a). The impact of useable waste streams on the exergoeconomic analysis is discussed in Section 6.3.3.

The results (Table 6-5) also present the scFOS powder scenario (S-F_P) with the best exergetic performance with irreversibility and exergetic efficiency of 6.3 MW and 85.8 %, respectively. This is because scFOS production involves simpler process units, fewer chemical reactions, and milder process condition than SA production. Consequently, better exergy performance is obtained from scFOS than the SA biorefineries. It is worth noting that, even though the multi-product SA70-S-F_S scenario has more process units than the single product SA scenario (SA-S), the later performs poorer exergetically with the irreversibility of 8263 vs. 9713 kW and efficiency of 61.8 vs. 48.7 %. The lower irreversibility of scenario SA70-S-F_S is due to the reduced production size of SA (70 vs. 100 % in scenario SA-S) combined with 30 % scFOS (versus 100 % in a less irreversible scenario S-F_P) because irreversibility is an extensive property (Chapter 3). Moreover, the absence of the energy-intensive spray dryer (present in scenario S-F_P) from the scFOS syrup plant of scenario SA70-S-F_S further reduces the irreversibility. Higher exergetic efficiency for this multi-product scenario than the SA-S scenario is attributed to the integration of a highly efficient scFOS process with the SA process, which offers an improved efficiency than the lone SA process (scenario SA-S).

The integration effect on the exergetic efficiency is further observed when the system boundary includes the sugar mill, into which the biorefineries are integrated. Based on the fuel-product definitions used in this study, the exergetic efficiency of the sugar mill process without the boiler is higher (91.6 %) than the biorefinery scenarios, because it involves purely physical process, which is usually less exergy-intensive than chemical processes (Kotas, 1995). Thus, the combined system (biorefinery plus sugar mill) performs better (69.4, 81.3, 84.2 and 89.1 %) compared to the corresponding individual biorefineries (37.7, 48.7, 61.8 and 85.8 %) 1G-2G SA -S, SA-S, SA70-S-F_S and S-F_P scenarios, respectively (Table 6-5). In other words, the sugar mill itself performs worse

by the integration. However, the trade-off is the economic benefits from the biorefineries, which also have improved exergy performance by the integration.

Table 6-5: Irreversibility, exergetic efficiency and relative irreversibility of the biorefinery systems and the combined sugar mill-biorefinery systems.

	Irreversibility, kW		Exergy efficiency, %		Relative irreversibility, %	
	<i>Individual processes</i>	<i>Sugar mill + biorefinery</i>	<i>Individual processes</i>	<i>Sugar mill + biorefinery</i>	<i>Individual processes</i>	<i>Sugar mill + biorefinery</i>
Sugar mill	15.1	-	91.6	-	3.7	-
1G-2G SA-S	22.7	38.7	37.7	69.4	8.6	5.9
SA-S	9.7	25.6	48.7	81.3	10.8	5.8
SA70-S-F _s	8.3	24.2	61.8	84.2	9.3	5.5
S-F _p	6.3	21.2	85.8	89.1	7.0	4.8

The biorefineries were also compared based on relative irreversibility (Atmaca & Yumrutaş, 2014a) defined as the percentage of fuel exergy destroyed in each system. Scenarios S-F_s has the lowest relative irreversibility (7.0 %) after the base sugar mill (3.7 %), showing the highest exergy performance among the A-molasses biorefineries. Then, the SA70-S-F_s scenario (9.3 %) followed by the SA-S scenario (10.8 %), respectively. Following the trend of irreversibility and exergetic efficiency in Table 6-5, the 1G-2G SA-S scenario was expected to have the highest relative irreversibility, but it showed a lower value than scenarios SA70-S-F_s and SA-S (8.6 % versus 9.3 % and 10.8 %, respectively). The unexpected result could be attributed to a latent exergetic effect of acid and enzyme catalysts on the chemical conversion of the lignocellulose polysaccharides (mainly cellulose and xylan) to simple sugars in the pre-treatment and hydrolysis reactors. This effect was not quantified since the enzyme and acid remains chemically unchanged in the process, resulting in lower fuel exergy evaluated (and lower irreversibility) than was actually expended (Goldberg, 2003). Accordingly, the assessment of the individual units shows the pre-treatment has the lowest relative irreversibility of 0.6 %. On the contrary, the effect was not obvious in the enzymatic transfructosylation of sucrose in scFOS production probably due to the nature of the reaction mechanism. Further studies are required to investigate the exergetic effect of catalyst and enzymes in the exergy performance of energy systems and of course, the cost of such effects through exergoeconomics.

It is, however, worth noting that the 1G-2G SA-S scenario performs worse than the 1G scenarios with the highest relative irreversibility (5.9 %) when the boundary is extended to include the sugar mill. This is because the 1G processes use A-molasses, eliminating the components in the second and third crystallization stages whereas the 1G-2G SA-S scenario does not involve any such restructuring. This

shows that valorising A-molasses leads to an improved exergy performance of the integrated system compared to lignocellulose feedstock, which is a useful result for the design of biorefineries.

It can be concluded, the relative irreversibility provides a more objective basis for exergy comparison since it is not influenced by the process size and purpose (bias associated with the definition of products and wastes) that affect the usual exergy indicators irreversibility and exergetic efficiency, respectively. This notwithstanding, scenario S-F_P shows the best exergy performance by all indications. However, for a holistic comparison, the cost-based exergy performance is required to judge the most cost-effective biorefinery for decision making using the exergoeconomic results.

6.3.3. Exergoeconomic analysis

Exergoeconomics was conducted at the aggregated system level and around the inside battery limit of the biorefinery (excluding the sugar mill and utility plants) to allow comparison of the of biorefinery production processes.

Table 6-6 summarises the exergoeconomic analysis results of the 1G-2G SA biorefinery presenting the thermodynamic properties, total exergy rates, exergy cost rates and unit exergy costs of all streams entering and exiting the system. Similar results for scenarios SA-S, SA70-S-F_S and S-F_P are presented in Tables A.6-3 - A.6-8 in the chapter Appendix. The difference between the rates of streams entering and exiting the system represent the investment cost rate according to the cost balance equation (Equation (6-3)). The most important streams entering the 1G-2G SA system based on the exergy cost rate are chilled water (10 212.48 US\$/h) and the lignocellulose feedstock (3 493.87 US\$/h). Similarly, A-molasses and chilled water are the most important fuel streams in the SA-S, SA70-S-F_S and S-F_P (Tables A.6-3 - A.6-8). The unit exergy costs of fuel (c_F) and product (c_P) are also evaluated for each scenario. However, the exergoeconomic performance of the system is assessed based on the cost rates related to the capital investment, (\dot{Z}) and irreversibility (\dot{C}_D), the exergoeconomic factor (f) and the relative cost difference (r). These parameters were evaluated using the methodology presented in Section 6.2.3 and compared for all developed biorefineries.

Table 6-6: Thermodynamic properties, total exergy rate, exergy cost rate and unit exergy cost of all streams entering and exiting the 1G-2G SA biorefinery

		\dot{m} , kg/h	T , °C	P , bar	$\dot{E}x$, kW	\dot{C} , US\$/h	c , US\$/GJ
<i>Input Streams</i>							
1	Air	79361	25.0	1.00	500.57	0.00	0.00
2	Water	88430	25.0	1.00	1227.11	141.78	32.09
3	Solvent make-up	626	25.0	1.00	5665.39	584.32	28.65
4	TMA make-up	27	25.0	1.00	72.96	21.43	81.58
5	Water	9950	25.0	1.00	138.07	15.95	32.09
6	Water	214000	25.0	1.00	2969.60	343.10	32.09
7	Activated carbon	2400	25.0	1.00	22771.35	0.00	0.00
8	Sulphuric acid	716	25.0	1.00	331.35	75.29	63.12
9	Lignocellulose feedstock	57710	25.0	1.00	178332.79	3493.87	5.44
10	Sugar mill export water	51425	55.4	3.08	808.29	82.45	28.33
11	Cellulase enzyme	3633	25.0	1.00	23233.98	1353.29	16.18
12	HP steam	19160	232.0	28.60	5556.53	444.52	22.22
13	Ammonium	403	25.0	1.00	2220.30	154.07	19.27
14	Microorganism	1441	40.0	1.00	20.03	0.00	0.00
15	Growth medium	344	25.0	1.00	60.88	49.93	227.86
16	Sodium carbonate	7214	50.0	1.00	515.26	1628.14	877.74
17	C-molasses	9670	25.0	1.00	36757.97	1859.45	14.05
18	Water	43105	25.0	1.00	598.15	69.11	32.09
19	Cooling water	1189679	25.0	1.01	16509.30	1907.36	32.09
20	Chilled water	543217	2.0	1.00	8021.40	10212.48	353.65
21	HP steam	2044	231.2	28.60	683.15	54.65	22.22
22	LP steam	30310	130.0	2.62	5773.86	461.91	22.22
23	Electricity				1016.74	81.34	22.22
	Total	2354864	-	-	313785.03	23034.00	-
<i>Output Stream</i>							
24	Dead cells	1380	37.0	1.00	86.14	3.67	11.83
25	Waste water	436659	49.6	1.00	31862.16	1357.06	11.83
26	Waste stream	6065	82.2	1.00	26186.69	1115.34	11.83
27	Air	81213	60.0	1.00	581.82	3.46	1.65
28	Succinic acid	18512	0.0	0.00	70095.97	5785.58	22.93
29	Flash steam	12727	100.2	1.00	1955.25	83.28	11.83
30	Activated carbon	3044	79.4	1.00	23589.19	34.83	0.41
31	Hydrolysate for cellulase	10522	50.0	2.00	29631.64	2445.74	22.93
32	Lignin-rich residue	19492	50.0	2.00	81160.06	3456.74	11.83
33	Cooling water	1189679	40.0	1.01	17090.54	1974.51	32.09
34	Chilled water	543217	15.0	1.00	7613.34	9692.96	353.65
35	HP steam condensate	2044	231.2	28.60	194.88	15.59	22.22
36	LP steam condensate	30310	129.0	2.62	1026.64	82.13	22.22
	Total	2354864	-	-	291074.33	26050.89	-

Figure 6-6 illustrates the unit exergy costs of fuel and product of the studied biorefineries. The unit exergy cost of the product was evaluated first by considering that the wastes produced are exergetically useful outside the system boundary and as a result have exergetic costs. On the other hand, focusing on the biorefinery alone, the wastes are losses and bear no exergy cost, in which case the products bear the costs of losses (Bejan et al., 1996). These two approaches are presented in Figure 6-6, to show the effect of waste streams in each of the biorefineries. It can be concluded from the results that valorising waste streams in biorefineries present a more cost-effective design from the exergoeconomic point of view than discarding them to the environment even though it will result in increased exergy destruction (Soltanian et al., 2019). Therefore, subsequent evaluations focus on the first approach, considering that the major waste streams are valorised elsewhere outside the boundary of the biorefinery as boiler feed and for biogas production (Nieder-Heitmann et al., 2019).

Scenario 1G-2G SA-S has the lowest unit exergy cost of fuel and consequently, the lowest unit exergy cost for product. The low unit exergy of cost fuel stems mainly from the low cost rate chilled water (10 212 US\$/h) compared to other biorefinery. Scenario SA-S has the highest cost rate of chilled water (38 868 US\$/h) and consequently, the highest unit exergy cost of fuel and product of the system. This is because in producing succinic acid from A-molasses, sucrose is first hydrolysed into glucose and fructose in an exothermic reaction before fermentation. Therefore, a higher rate of chilled water is required to cool the fermenter than in the case of fermenting lignocellulose hydrolysates (mainly fermentable sugars), which is the dominant feed for scenario 1G-2G SA-S. The exergoeconomic analysis, therefore, reveals that succinic acid production from lignocellulose hydrolysate (having simple sugars) is more cost-effective. Because a lower production cost is obtained from lignocellulosic process compared to sucrose based substrates, even though the pre-treatment of lignocellulose could be expensive (Nieder-Heitmann et al., 2019).

It can be concluded that even though the economics of the 1G-2G SA scenario is the highest, the use of much lower exergy streams (lignocelluloses and C-molasses) with lower rates of the most important utility for the SA production (chilled water) makes it an attractive scenario to consider from the exergoeconomic point of view. On the other hand, cheaper source of chilled water should be sort for a cost-effective design of the A-molasses/sucrose based scenarios. A detailed evaluation of the absorption chiller considered to provide the chilled water will reveal its actual cost compared to what is assumed in this study.

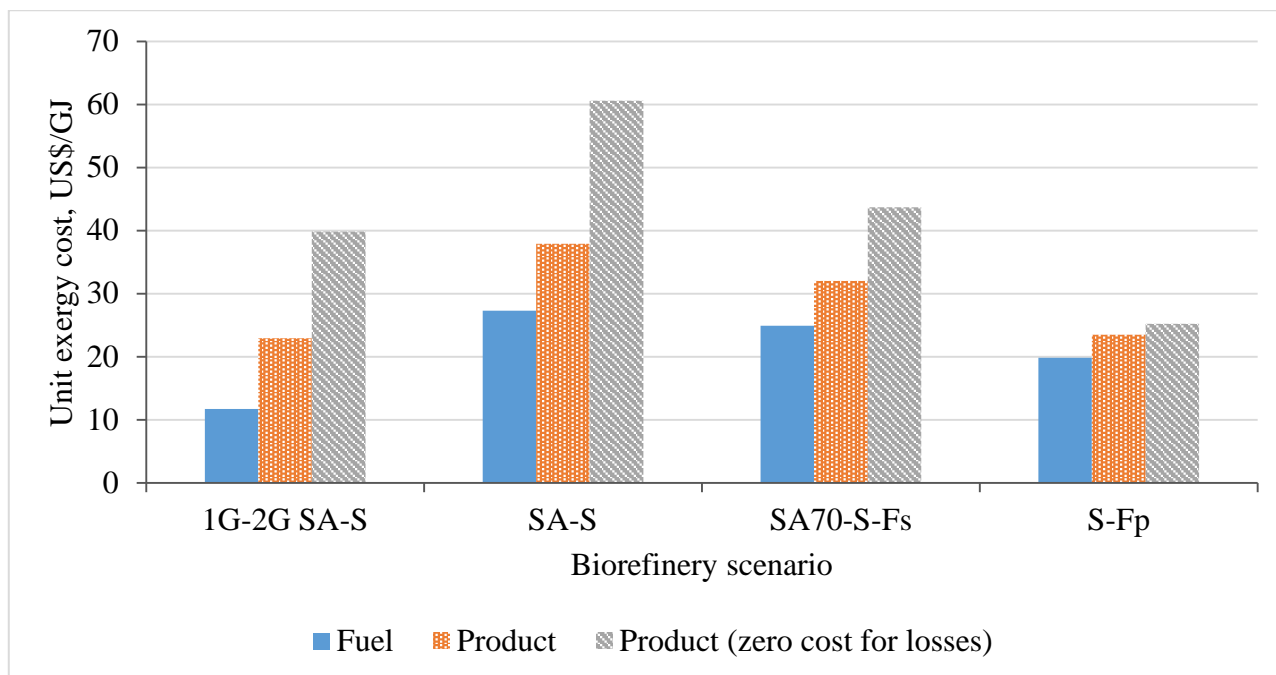


Figure 6-6: Unit exergy costs of fuel and product of the aggregated biorefinery systems

The cost rates of the biorefinery composing of the irreversibility cost rates and the investment cost rate are also presented in Figure 6-7. This parameter illustrates at a glance which effect is important in improving the holistic performance of the system; exergy or economics (Aghbashlo et al., 2019). The 1G-2G SA plant has the highest values for the total cost rate (3 983 US\$/h), the investment cost rate (3 016 US\$/h) and the irreversibility cost rate of 967 US\$/h. The trend in the cost associated with irreversibility is determined by the different levels of process complexity and exergy intensity of the biorefineries as observed for irreversibility results (see Table 6-5). Though 1G-2G SA-S biorefinery significantly is more complex, with several more process units and reactions than the SA-S plant, the associated irreversibility cost rate is only slightly higher (967 US\$/h vs. 956 US\$/h), which is attributed to the catalytic and enzymatic effect on the exergy of the pre-treatment and hydrolysis reactions as discussed previously. Therefore, catalytic reactions could be good options for improving the exergy performance of high exergy destructive systems involving chemical reactions, with economic consideration of course.

The cost rate results show that the 1G-2G SA-S biorefinery need more improvement in its design than the other scenarios. The extent and means by which the system should be improved is indicated by the exergoeconomic factor.

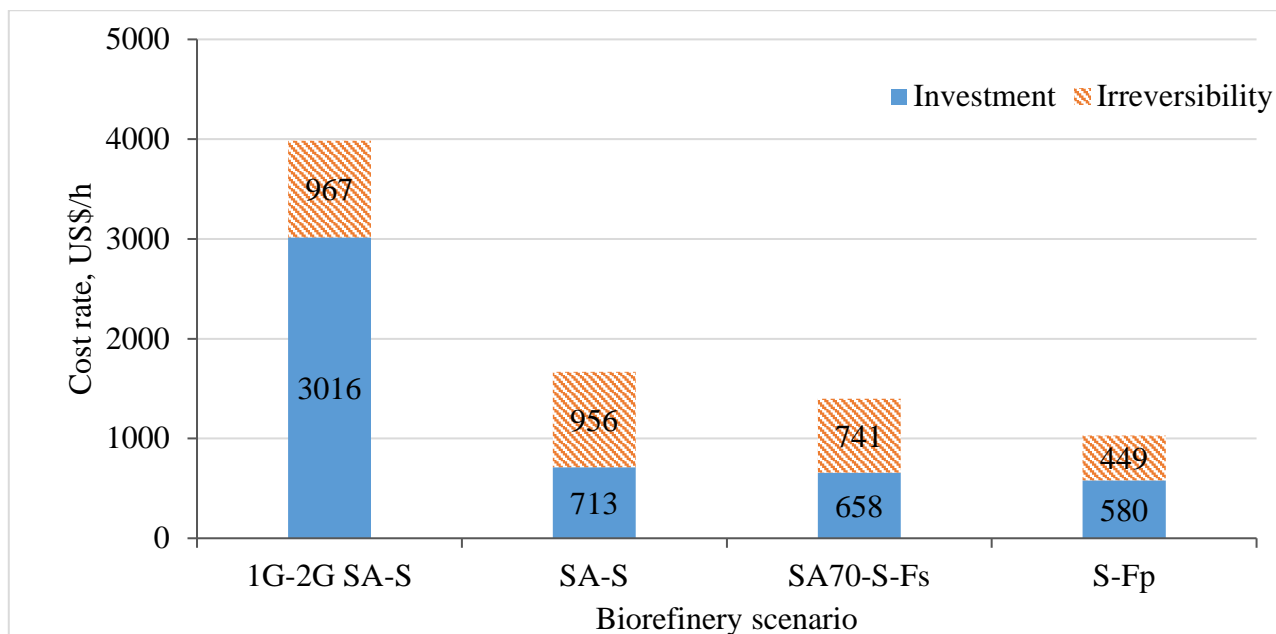


Figure 6-7: Investment cost rate and irreversibility cost rate of the aggregated biorefinery systems

Table 6-7 presents the exergoeconomic factor of the biorefineries compared in this study. A high value of the exergoeconomic factor suggests that the investment cost rate contributes more to the system's performance than the irreversibility cost rate. Therefore, reducing investment cost rate, even at the expense of exergy efficiency will improve the system and vice versa (Bejan et al., 1996). Scenario 1G-2G SA-S shows the highest exergoeconomic factor of 0.76 followed by S-F_P (0.56) proving that the investment cost rate contributes more to the system than the irreversibility cost rates of these systems (following the observation from Figure 6-7). The factor is lower for scenarios SA-S and SA70-S-F_S because the irreversibility cost rates contribute more to the total cost rates of these biorefineries (even much more for scenario SA-S) than the capital investment cost rates (see Figure 6-7). For an optimized system, the exergoeconomic factor of 0.6 indicates an appropriate balance between the investment- and irreversibility-related cost rates (Soltanian et al., 2019). In other words, for an exergoeconomically optimal system, the investment cost rate should be approximately 50 % more than the irreversibility cost rate. Therefore, scenario S-F_P is almost optimized while the 1G-2G SA-S scenario can be improved by reducing the capital investment cost even if at the expense of the process efficiency. A practical way to do this is to consider a fed-batch process for the fermentation of hydrolysate or find suitable microorganism strains that tolerate high initial sugars concentration for the fermentation of xylose.

The relative cost difference, which indicates the potential of decreasing the unit exergy cost of product (Soltanian et al., 2019) is also presented in Table 6-7. Scenario 1G-2G SA-S has the highest relative cost difference of 0.94 signifying that the unit exergy cost its product can be reduced with ease, even though it is the lowest compared to the other biorefineries (Figure 6-6). Based on the discussions of

the exergoeconomic factor, this improvement will be possible if the capital investment can be reduced at the expense of the exergy efficiency due to its significant contribution to the exergoeconomic performance of the scenario. On the other hand, scenario S-F_P with r value of 0.19 has almost no chance to reduce its product cost.

Table 6-7: Exergoeconomic factor and relative cost difference of the developed biorefineries

Biorefinery scenario	f	r
1G-2G SA-S	0.76	0.94
SA-S	0.43	0.39
SA70-S-F _S	0.47	0.28
S-F _P	0.56	0.19

It can be concluded from the results that the design of scenario S-F_P is almost optimized with little or no further improvements required for its implementation. On the other hand, better designs are possible for scenarios SA-S and SA70-S-F_S by reducing their irreversibility, while less capital-intensive processes are desired for a better design of scenario 1G-2G-SA-S, even at the expense of increasing inefficiencies (Soltanian et al., 2019). A component level exergoeconomic studies for each of the scenarios, especially scenario 1G-2G-SA-S, will be required to identify most important components to focus on for the required improvements.

6.4. Conclusion

The study compares four integrated biorefinery scenarios annexed to an existing sugar mill using an aggregated system exergoeconomic methodology. The performances of the biorefineries were evaluated based on various exergoeconomic parameters including capital investment cost rate and irreversibility cost rate making up the total cost rate, exergoeconomic factor and relative cost difference. With the lowest total cost rate of 1029 US\$/h and an exergoeconomic factor of 0.56, indicating a good balance between the irreversibility- and investment-related costs, the single-product scFOS powder scenario S-F_P emerged the most cost-effective scenario for integration into the sugar mill. Moreover, its relative cost difference of 0.19 showed the system is almost optimized. On the other hand, scenario 1G-2G SA-S recorded the worst performance, mainly due to high investment cost rate. However, the exergoeconomic methodology revealed that it can be optimized by minimizing the capital investment at the expense of process efficiency. Further studies can consider a suitable microorganism strain that utilises all sugar types in the combined 1G-2G sugar stream and withstand higher initial sugars concentrations, while using a fed-batch fermentation instead of batch

to reduce the investment-related cost of scenario 1G-2G SA-S. In addition, component exergoeconomic analysis should be conducted to reveal the most important component of the system to optimize.

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References

- Aghbashlo, M., Mandegari, M., Tabatabaei, M., Farzad, S., Mojarab Soufiyan, M., & Görgens, J. F. (2018). Exergy analysis of a lignocellulosic-based biorefinery annexed to a sugarcane mill for simultaneous lactic acid and electricity production. *Energy*, *149*, 623–638. <http://doi.org/10.1016/j.energy.2018.02.063>
- Aghbashlo, M., Tabatabaei, M., Soltanian, S., Ghanavati, H., & Dadak, A. (2019). Comprehensive exergoeconomic analysis of a municipal solid waste digestion plant equipped with a biogas genset. *Waste Management*, *87*, 485–498. <http://doi.org/10.1016/j.wasman.2019.02.029>
- Ali Mandegari, M., Farzad, S., & Görgens, J. F. (2017). Economic and environmental assessment of cellulosic ethanol production scenarios annexed to a typical sugar mill. *Bioresource Technology*, *224*, 314–326. <http://doi.org/10.1016/j.biortech.2016.10.074>
- Andersson, C., Hodge, D., Berglund, K. A., & Rova, U. (2007). Effect of Different Carbon Sources on the Production of Succinic Acid Using Metabolically Engineered Escherichia coli. *Biotechnology Progress*, (23), 381–388. <http://doi.org/10.1021/bp060301y>
- Atmaca, A., & Yumrutaş, R. (2014a). Thermodynamic and exergoeconomic analysis of a cement plant: Part I – Methodology. *Energy Conversion and Management*, *79*, 790–798. <http://doi.org/10.1016/j.enconman.2013.11.053>
- Atmaca, A., & Yumrutaş, R. (2014b). Thermodynamic and exergoeconomic analysis of a cement plant: Part II – Application. *Energy Conversion and Management*, *79*, 799–808. <http://doi.org/10.1016/j.enconman.2013.11.054>
- Bedzo, O. K. K., Mandegari, M., & Görgens, J. F. (2019). Comparison of immobilized and free enzyme systems in industrial production of short-chain fructooligosaccharides from sucrose using a techno-economic approach. *Biofuels, Bioproducts and Biorefining*, bbb.2025.

<http://doi.org/10.1002/bbb.2025>

- Bejan, A., Tsatsaronis, G. (George), & Moran, M. J. (1996). *Thermal design and optimization*. Wiley. Retrieved from <https://www.wiley.com/en-us/Thermal+Design+and+Optimization-p-9780471584674>
- Cavalett, O., Junqueira, T. L., Dias, M. O. S., Jesus, C. D. F., Mantelatto, P. E., Cunha, M. P., Bonomi, A. (2012). Environmental and economic assessment of sugarcane first generation biorefineries in Brazil. *Clean Technologies and Environmental Policy*, *14*(3), 399–410. <http://doi.org/10.1007/s10098-011-0424-7>
- Chan, S., Kanchanatawee, S., & Jantama, K. (2012). Production of succinic acid from sucrose and sugarcane molasses by metabolically engineered *Escherichia coli*. *Bioresource Technology*, *103*(1), 329–336. <http://doi.org/10.1016/j.biortech.2011.09.096>
- Dias, M. O. de S., Maciel Filho, R., Mantelatto, P. E., Cavalett, O., Rossell, C. E. V., Bonomi, A., & Leal, M. R. L. V. (2015). Sugarcane processing for ethanol and sugar in Brazil. *Environmental Development*, *15*, 35–51. <http://doi.org/10.1016/j.envdev.2015.03.004>
- Dogbe, E. S., Mandegari, M., & Görgens, J. F. (2019). Assessment of the thermodynamic performance improvement of a typical sugar mill through the integration of waste-heat recovery technologies. *Applied Thermal Engineering*, *158*, 113768. <http://doi.org/10.1016/j.applthermaleng.2019.113768>
- Dominguez, A. L., Rodrigues, L. R., Lima, N. M., & Teixeira, J. A. (2014). An Overview of the Recent Developments on Fructooligosaccharide Production and Applications. *Food and Bioprocess Technology*, *7*(2), 324–337. <http://doi.org/10.1007/s11947-013-1221-6>
- Efe, Ç., van der Wielen, L. A. M., & Straathof, A. J. J. (2013). Techno-economic analysis of succinic acid production using adsorption from fermentation medium. *Biomass and Bioenergy*, *56*, 479–492. <http://doi.org/10.1016/j.biombioe.2013.06.002>
- Farzad, S., Mandegari, M. A., & Görgens, J. F. (2017). Integrated techno-economic and environmental analysis of butadiene production from biomass. *Bioresource Technology*, *239*, 37–48. <http://doi.org/10.1016/j.biortech.2017.04.130>
- Farzad, S., Mandegari, M. A., Guo, M., Haigh, K. F., Shah, N., & Görgens, J. F. (2017). Multi-product biorefineries from lignocelluloses: a pathway to revitalisation of the sugar industry? *Biotechnology for Biofuels*, *10*. <http://doi.org/10.1186/s13068-017-0761-9>
- Goldberg, R. N. (2003). Enzyme catalysis (thermodynamics). *Access Science*.

<http://doi.org/10.1036/1097-8542.YB030795>

- Görgens, J. F., & Mandegari, M. (2018). Comparative assessment of non-ethanol biofuel production from sugarcane lignocelluloses in Africa, including synfuels, butanol and jet fuels. In *Sugarcane Bioenergy for Sustainable Development* (pp. 97–108). Routledge. <http://doi.org/10.4324/9780429457920-9>
- Guest, K., Stark, A., & Starzak, M. (2019). Development and verification of an ASPEN Plus® model of a sugarcane biorefinery. *Proc S Afr Sug Technol Ass*, 92(August), 254–273.
- Humbird, D., Davis, R., Tao, L., Kinchin, C., Hsu, D., & Aden, A. (2011). *Process Design and Economics for Biochemical Conversion of Lignocellulosic Biomass to Ethanol. Technical Report NREL/TP-5100-47764*. Golden, Colorado.
- Klein, B. C., Silva, J. F. L., Junqueira, T. L., Rabelo, S. C., Arruda, P. V., Ienczak, J. L., Bonomi, A. (2017). Process development and techno-economic analysis of bio-based succinic acid derived from pentoses integrated to a sugarcane biorefinery. *Biofuels, Bioproducts and Biorefining*, 11(6), 1051–1064. <http://doi.org/10.1002/bbb.1813>
- Kotas, T. J. (1995). *The Exergy Method of Thermal Plant Analysis* (Reprint Ed). Florida, USA: Krieger Publishing Company, Malabar. <http://doi.org/10.1016/B978-0-408-01350-5.50003-9>
- Krajnc, D., & Glavič, P. (2009). Assessment of different strategies for the co-production of bioethanol and beet sugar. *Chemical Engineering Research and Design*, 87(9), 1217–1231. <http://doi.org/10.1016/j.cherd.2009.06.014>
- Lazzaretto, A., & Tsatsaronis, G. (2006). SPECO: A systematic and general methodology for calculating efficiencies and costs in thermal systems. *Energy*, 31(8–9), 1257–1289. <http://doi.org/10.1016/j.energy.2005.03.011>
- Liu, R., Liang, L., Cao, W., Wu, M., Chen, K., Ma, J., ... Ouyang, P. (2013). Succinate production by metabolically engineered *Escherichia coli* using sugarcane bagasse hydrolysate as the carbon source. *Bioresource Technology*, 135, 574–577. <http://doi.org/10.1016/J.BIORTECH.2012.08.120>
- Mandegari, M., Farzad, S., & Görgens, J. F. (2018). A new insight into sugarcane biorefineries with fossil fuel co-combustion: Techno-economic analysis and life cycle assessment. *Energy Conversion and Management*, 165, 76–91. <http://doi.org/10.1016/j.enconman.2018.03.057>
- Mkwanzani, T., Mandegari, M., & Görgens, J. F. (2019). Disturbance modelling through steady-state value deviations: The determination of suitable energy indicators and parameters for energy

consumption monitoring in a typical sugar mill. *Energy*, 176, 211–223. <http://doi.org/10.1016/j.energy.2019.03.191>

Morales, M., Ataman, M., Badr, S., Linster, S., Kourlimpinis, I., Papadokonstantakis, S., Hungerbühler, K. (2016). Sustainability assessment of succinic acid production technologies from biomass using metabolic engineering. *Energy & Environmental Science*, 9(9), 2794–2805. <http://doi.org/10.1039/C6EE00634E>

Nieder-Heitmann, M., Haigh, K., & Görgens, J. F. (2019). Process design and economic evaluation of integrated, multi-product biorefineries for the co-production of bio-energy, succinic acid, and polyhydroxybutyrate (PHB) from sugarcane bagasse and trash lignocelluloses. *Biofuels, Bioproducts and Biorefining*, 13(3), 599–617. <http://doi.org/10.1002/bbb.1972>

OECD/FAO. (2016). Sugar. In *OECD-FAO Agricultural Outlook 2016-2025* (pp. 1–13). Paris: OECD Publishing, Paris. http://doi.org/10.1787/agr_outlook-2016-9-en

Ozbilen, A., Dincer, I., & Rosen, M. A. (2016). Development of a four-step Cu–Cl cycle for hydrogen production – Part I: Exergoeconomic and exergoenvironmental analyses. *International Journal of Hydrogen Energy*, 41(19), 7814–7825. <http://doi.org/10.1016/j.ijhydene.2015.12.184>

Pachón, E. R., Vaskan, P., Raman, J. K., & Gnansounou, E. (2018). Transition of a South African sugar mill towards a biorefinery. A feasibility assessment. *Applied Energy*, 229, 1–17. <http://doi.org/10.1016/J.APENERGY.2018.07.104>

Petersen, A. M., Aneke, M. C., & Görgens, J. F. (2014). Techno-economic comparison of ethanol and electricity coproduction schemes from sugarcane residues at existing sugar mills in Southern Africa. *Biotechnology for Biofuels*, 7(1), 105. <http://doi.org/10.1186/1754-6834-7-105>

Soltanian, S., Aghbashlo, M., Farzad, S., Tabatabaei, M., Mandegari, M., & Görgens, J. F. (2019). Exergoeconomic analysis of lactic acid and power cogeneration from sugarcane residues through a biorefinery approach. *Renewable Energy*, 143, 872–889. <http://doi.org/10.1016/j.renene.2019.05.016>

Taylor, R., Natrass, L., Alberts, G., Robson, P., Chudziak, C., Bauen, A., Ree, R. van. (2015). *From the Sugar Platform to biofuels and biochemicals*. Retrieved from https://ec.europa.eu/energy/sites/ener/files/documents/EC_Sugar_Platform_final_report.pdf

Tsatsaronis, G. (2007). Definitions and nomenclature in exergy analysis and exergoeconomics. *Energy*, 32(4), 249–253. <http://doi.org/10.1016/j.energy.2006.07.002>

Vaswani, S. (2010). *Bio-based succinic acid*. Menlo Park, California 94025. Retrieved from

https://www.ihs.com/pdf/RW2010-14_220240110917062932.pdf

- Vučković, G. D., Stojiljković, M. M., Vukić, M. V., Stefanović, G. M., & Dedeić, E. M. (2014). Advanced exergy analysis and exergoeconomic performance evaluation of thermal processes in an existing industrial plant. *Energy Conversion and Management*, 85, 655–662. <http://doi.org/10.1016/j.enconman.2014.03.049>
- Wang, H., Pan, J., Wang, J., Wang, N., Zhang, J., Li, Q., Zhou, X. (2014). Succinic acid production from xylose mother liquor by recombinant *Escherichia coli* strain. *Biotechnology, Biotechnological Equipment*, 28(6), 1042–1049. <http://doi.org/10.1080/13102818.2014.952501>
- Wang, J., Zhu, J., Bennett, G. N., & San, K.-Y. (2011). Succinate production from different carbon sources under anaerobic conditions by metabolic engineered *Escherichia coli* strains. *Metabolic Engineering*, 13(3), 328–335. <http://doi.org/10.1016/j.ymben.2011.03.004>

Appendix

Table A.6-1: Costing of capital investment

Cost component		US\$
<i>Plant area</i>		<i>Equipment Installed cost</i>
Pre-treatment		16 452 000
Pentose – seed train		10 490 000
Pentose - fermentation		23 718 000
Hexose – seed train		8 386 000
Hexose - fermentation		56 330 000
Downstream processing		38 336 000
<i>IBL</i>		<i>153 712 000</i>
Chiller		1 483 963
Cooling Tower system		628 150
Cogeneration System		11 883 182
<i>OBL</i>		<i>13 996 000</i>
TEPC		167 708 000
Fixed Capital Investment (FCI)		323 178 000
Working Capital (WC)	0.05 of FCI	16 158 900
Total Capital Investment (TCI)	FCI + WC	339 337 000

Table A.6-2: Annual manufacturing costs

Annual Manufacturing Cost	US\$
Total Variable Costs	41 972 000
Fixed Operating Costs	7 712 000
Annual capital charge	2 061 000
Marginal change in sugar mill revenue	12 693 000
Total Product Cost (OPEX)	64 438 000

Table A.6-3: Thermodynamic properties, total exergy rate, exergy cost rate and unit exergy cost of all streams entering the SA-S biorefinery

		\dot{m} , kg/h	T , °C	P , bar	$\dot{E}x$, kW	\dot{C} , US\$/h	c , US\$/GJ
<i>Input Streams</i>							
1	Drying air	43850	25.00	1.00	276.58	0.00	0.00
2	Inoculum	371	25.00	1.00	5.15	0.00	0.00
3	Medium	73	25.00	1.00	10.96	10.57	267.74
4	Ammonium	85	25.00	1.00	469.36	32.57	19.27
5	NaHCO ₃	5860	25.00	1.00	418.54	1322.52	877.74
6	Sugar mill steam - 1	4333	122.73	2.13	786.83	0.00	0.00
7	Sugar mill steam - 2	2844	115.63	1.69	488.48	0.00	0.00
8	A-molasses	21194	56.00	1.00	80958.84	5199.70	17.84
9	Water	51425	55.41	3.08	808.29	82.45	28.33
10	Water	47466	25.00	1.00	658.66	76.10	32.09
11	Wash water	9950	25.00	1.00	138.07	11.05	22.22
12	Make up TMA	17	25.00	1.00	45.65	3.65	22.22
13	Make up solvent	162	25.00	1.00	1466.12	151.21	28.65
14	Cooling water	275317	25.00	1.01	3820.60	441.40	32.09
15	Chilled water	2067466	2.00	1.00	30529.19	38868.37	353.65
16	HP steam	1124	231.23	28.60	375.70	0.00	0.00
17	LP steam	10727	130.00	2.62	2043.42	0.00	0.00
18	Electricity	-	-	-	538.25	0.00	0.00
Total		2542264	-	-	123838.70	46199.58	-

Table A.6-4: Thermodynamic properties, total exergy rate, exergy cost rate and unit exergy cost of all streams exiting the SA-S biorefinery

		\dot{m} , kg/h	T , °C	P , bar	$\dot{E}x$, kW	\dot{C} , US\$/h	c , US\$/GJ
<i>Output Stream</i>							
19	Sugar mill steam – 1 condensate	4333	121.97	2.13	126.12	0.00	0.00
20	Sugar mill steam – 2 condensate	2844	114.90	1.69	82.11	0.00	0.00
21	Succinic acid	11576	0.00	0.00	43832.73	5983.69	37.92
22	Waste stream	5367	20.00	1.00	25585.47	2517.69	27.33
23	Waste stream	117941	43.36	1.00	7925.48	779.89	27.33
24	Microorganism cells	573	37.00	1.00	2842.49	279.71	27.33
25	Drying air out	44847	60.02	1.00	314.48	3.73	3.29
26	Cooling water	275317	40.00	1.01	3955.11	456.94	32.09
27	Chilled water	2067466	15.00	1.00	28976.11	36891.06	353.65
28	HP steam	1124	231.23	28.60	107.17	0.00	0.00
29	LP steam	10727	128.97	2.62	363.34	0.00	0.00
30	Waste stream	108	23.83	1.00	14.89	0.00	0.00
Total		2542223	-	-	114125.50	46912.72	-

Table A.6-5: Thermodynamic properties, total exergy rate, exergy cost rate and unit exergy cost of all streams entering the SA-S-Fs biorefinery

		\dot{m} , kg/h	T , °C	P , bar	$\dot{E}x$, kW	\dot{C} , US\$/h	c , US\$/GJ
<i>Input Streams</i>							
1	Drying air	41000.0	25.00	1.00	258.61	0.00	0.00
2	Inoculum - 1	266.5	25.00	1.00	3.70	0.00	0.00
3	Medium - 1	51.0	25.00	1.00	7.67	7.40	267.74
4	Ammonium	59.6	25.00	1.00	328.55	0.00	0.00
5	NaHCO ₃	3615.5	25.00	1.00	258.23	815.97	877.74
6	Sugar mill steam - 1	4333.5	122.73	2.13	786.83	0.00	0.00
7	A-molasses	14827.1	56.00	1.00	56671.10	3637.66	17.83
8	Water	50094.2	55.41	3.08	787.37	80.32	28.34
9	Water	21040.0	25.00	1.00	291.96	33.73	32.10
10	Wash water	9950.0	25.00	1.00	138.07	0.00	0.00
11	Make up TMA	10.9	25.00	1.00	29.46	0.00	0.00
12	Make up solvent	114.0	25.00	1.00	1031.72	106.41	28.65
13	Air	124.6	25.00	1.00	1.57	0.00	0.00
14	Ammonium	1.9	25.00	1.00	0.03	0.67	7261.76
15	Trace salts	0.2	25.00	1.00	0.00	0.00	0.00
16	Inoculum - 2	1.1	30.00	1.00	0.02	0.00	0.00
17	Glycerol	17.1	25.00	1.00	91.13	11.98	36.50
18	Water	11.0	25.00	1.00	0.15	0.02	32.09
19	Trace salts	0.1	25.00	1.00	0.00	0.00	0.00
20	Medium - 2	1.3	25.00	1.00	2.66	1.22	127.18
21	Water	1330.7	55.41	3.08	20.92	2.13	28.34
22	A-molasses	6358.3	56.00	1.00	24287.65	1559.94	17.84
23	Sugar mill steam - 2	2843.7	115.63	1.69	488.48	0.00	0.00
24	Cooling water	438013.4	25.00	1.01	6078.36	702.29	32.09
25	Chilled water	1761407.8	2.00	1.00	26009.78	33114.47	353.65
26	HP steam	1077.0	231.23	28.60	359.92	0.00	0.00
27	LP steam	11481.0	130.00	2.62	2187.07	0.00	0.00
28	Electricity	-	-	-	501.65	40.13	22.22
Total		2368031.4	-	-	120622.67	40114.34	-

Table A.6-6: Thermodynamic properties, total exergy rate, exergy cost rate and unit exergy cost of all streams exiting the SA-S-Fs biorefinery

		\dot{m} , kg/h	T , °C	P , bar	$\dot{E}x$, kW	\dot{C} , US\$/h	c , US\$/GJ
<i>Output Stream</i>							
29	Sugar mill steam – 1 condensate	4333.5	121.97	2.13	126.12	0.00	0.00
30	Succinic acid	8103.5	0.00	0.00	30683.69	3536.41	32.01
31	Waste stream	3756.6	20.00	1.00	17908.77	1606.92	24.92
32	Waste stream	86666.4	42.76	1.00	5589.21	501.51	24.92
33	Microorganism cells	401.3	37.00	1.00	1988.85	178.46	24.92
34	Drying air out	41996.1	59.62	1.00	294.83	3.25	3.06
35	Vent air	131.5	30.00	1.00	3.75	0.20	14.51
36	Pichia cells	4.8	30.00	1.00	33.30	2.99	24.92
37	Pichia cells	0.1	30.00	1.00	0.68	0.06	24.92
38	scFOS syrup	2681.9	35.00	1.00	10157.14	1170.65	32.01
39	Invert sugars syrup	4251.1	35.00	1.00	13971.77	1610.30	32.01
40	Sugar mill steam – 2 condensate	2843.7	114.90	1.69	82.11	0.00	0.00
41	Waste water	776.9	78.00	1.00	38.98	3.50	24.92
42	Cooling water	438013.4	40.00	1.01	6292.36	727.02	32.09
43	Chilled water	1761407.8	15.00	1.00	24686.62	31429.87	353.65
44	HP steam	1077.0	231.23	28.60	102.67	0.00	0.00
45	LP steam	11481.0	128.97	2.62	388.88	0.00	0.00
46	Waste stream	77.7	23.45	1.00	10.45	0.94	24.92
Total		2368004.3	-	-	112360.16	40772.06	-

Table A.6-7: Thermodynamic properties, total exergy rate, exergy cost rate and unit exergy cost of all streams entering the S-F_P biorefinery

		\dot{m} , kg/h	T , °C	P , bar	$\dot{E}x$, kW	\dot{C} , US\$/h	c , US\$/GJ
<i>Input Streams</i>							
1	Air	417.4	25.00	1.00	5.24	0.00	0.00
2	Ammonium	6.2	25.00	1.00	0.09	2.26	7261.76
3	Trace salts	0.6	25.00	1.00	0.01	0.00	0.00
4	Inoculum	3.8	30.00	1.00	0.05	0.00	0.00
5	Glycerol	57.4	25.00	1.00	305.36	40.13	36.50
6	Water	36.8	25.00	1.00	0.51	0.06	32.09
7	Trace salts	0.2	25.00	1.00	0.00	0.00	0.00
8	Medium	4.5	25.00	1.00	8.92	4.08	127.18
9	Water	4424.3	55.41	3.08	69.54	7.09	28.34
10	A-molasses	21194.8	56.00	1.00	80958.85	5199.89	17.84
11	Sugar mill steam - 1	4332.8	122.73	2.13	786.72	0.00	0.00
12	Sugar mill steam - 2	2843.9	115.63	1.69	488.51	0.00	0.00
13	Make up eluent	3646.1	25.00	1.00	50.60	5.85	32.09
14	Drying air out	228000.0	25.00	1.00	1438.07	0.00	0.00
15	Cooling water	369729.5	25.00	1.01	5130.78	592.81	32.09
16	Chilled water	985741.3	2.00	1.00	14555.92	18531.94	353.65
17	HP steam	10844.4	231.23	28.60	3624.02	0.00	0.00
18	LP steam	9093.0	130.00	2.62	1732.17	0.00	0.00
19	Electricity	-	-	-	2732.74	218.62	22.22
Total		1640377.1	-	-	111888.11	24602.72	-

Table A.6-8: Thermodynamic properties, total exergy rate, exergy cost rate and unit exergy cost of all streams exiting the S-F_P biorefinery

		\dot{m} , kg/h	T , °C	P , bar	$\dot{E}x$, kW	\dot{C} , US\$/h	c , US\$/GJ
<i>Output Stream</i>							
20	Vent air	440.6	30.00	1.00	12.55	0.52	11.53
21	Pichia cells	16.1	30.00	1.00	111.58	7.96	19.81
22	Pichia cells	0.3	30.00	1.00	2.28	0.16	19.81
23	scFOS powder	6157.2	35.00	1.50	30608.19	2591.58	23.52
24	Invert sugars syrup	14119.5	35.00	1.00	46400.72	3928.73	23.52
25	Sugar mill steam – 1 condensate	4332.8	121.97	2.13	135.57	0.00	0.00
26	Sugar mill steam – 2 condensate	2843.9	114.90	1.69	82.11	0.00	0.00
27	Drying air out	237058.1	90.22	1.50	7764.14	451.05	16.14
28	Cooling water	369729.5	40.00	1.01	5311.41	613.68	32.09
29	Chilled water	985741.3	15.00	1.00	13815.44	17589.18	353.65
30	HP steam	10844.4	231.23	28.60	1033.80	0.00	0.00
31	LP steam	9093.0	128.97	2.62	308.00	0.00	0.00
Total		1640376.7	-	-	105585.80	25182.87	-

Chapter 7

Conclusions and recommendations

7.1. Conclusions

The economic sustainability of the South African sugar industry is currently challenged. Though the industry is one of the world's best producers of high-quality sugar, the dropping world sugar market prices induced by overproduction of major sugar-producing countries require lower costs of production of sugar for economic competitiveness. It is expected that improving the energy efficiency of the industry would reduce the cost of sugar production. Moreover, the increasing trend of global warming and energy insecurity worldwide demand efficient use of energy, especially in energy-intensive industries such as the sugar mill. On the other hand, introducing higher-value products through the biorefinery concept would revamp the industry and ensure its economic sustainability. In view of these, the aim of this project was to improve the energy efficiency and economic sustainability and competitiveness of the South African sugar industry.

A typical South African sugar mill with a capacity of 250 tons of cane crushed per hour, simulated in Aspen Plus chemical process simulator was employed for the study. The approach included first assessing the energy inefficiency of the mill and then integrating energy-efficient technologies and biorefineries to improve the thermodynamic efficiency and economic sustainability, respectively.

The results of the exergy assessment of the sugar mill showed that the cogeneration system was responsible for 86 % of the total exergy destruction of the sugar mill in agreement with the findings of other sugar mill exergy studies in the literature. This indicates that efforts towards improving the thermodynamic efficiency of the sugar mill should majorly focus on the cogeneration system. Moreover, the crystallization unit was found to use energy most inefficiently due to the complexity of the three-stage crystallization process, having exergetic efficiency of 9.6 %. This was a major outcome of this study that has not been previously revealed in the literature. Simpler sugar crystallization configuration such as a single-stage crystallization was recommended and further explored in Objective 5 of this study. Therefore, exergy analysis was successfully used to identify the sources, magnitude and causes of thermodynamic inefficiency of the sugar mill (Objective 1) for better design and operation of sugar mill and integrated biorefineries.

Following the results of exergy analysis, absorption heat pump (AHP) and organic Rankine cycle (ORC) were selected and integrated into the sugar mill for waste heat recovery towards improving

the efficiency, including their economic feasibility evaluation. The integration of AHP reduced the overall sugar mill irreversibility by 0.14 % while ORC integration improved the cogeneration system exergetic efficiency by 1.7 %. Though the improvements achieved by these integrations are only minimal, both were found to be economically feasible and should be considered for implementation. This implies that exergy analysis could adequately inform the selection and integration of economically viable technologies into the sugar mill for energy improvement (Objectives 2 and 3).

Based on the results of the crystallization unit exergy performance, an integrated biorefinery concept was developed to produce succinic acid and (SA) and short-chain fructooligosaccharides (scFOS), two promising biorefinery products annex the sugar mill to valorise A-molasses, thus diverting sucrose away from the crystallization units. The biorefineries were found to be more economical than the stand-alone plants reported in the literature due to the integration strategy used and worthwhile when sugar prices remain low. Moreover, the biorefinery concept was extended to utilise surplus lignocellulose residue of the sugarcane plant as a second-generation (2G) feedstock in addition to the by-product of sugar production, C-molasses, a first-generation feedstock (1G) to produce succinic acid as an integrated biorefinery option to obtain maximum economic value from the sugarcane plant considering the current sugar mill configuration. All developed scenarios were economically feasible, indicating the success of exergy principle for the design of economically viable biorefineries for the sugar industry (Objective 4). Particularly, the multi-product SA and scFOS syrup scenario SA70-S-F_s showed the best economic performance with considering the products market sizes to avoid oversupplying one particular market, while also providing diversification of revenue, and easy market access while the 1G-2G SA scenario was the worst.

Following economic and exergy results of all the biorefineries developed, an aggregated system exergoeconomic methodology was applied to select the most cost-effective biorefinery for implementation. The single-product scFOS powder scenario (S-F_P) was the preferred biorefinery for a cost-effective integration into the sugar mill. With exergoeconomic factor and relative cost difference of 0.56 and 0.19, respectively, the scenario showed a good balance between the irreversibility- and investment-related costs indicating an optimum performance. Therefore, the exergoeconomic methodology provides insightful information for the design, optimization and selection of a cost-effective biorefinery compared to simple exergy and economic parameters (objective 5).

7.2. Recommendations

The findings of the study showed that the cogeneration system is the principal unit to prioritise for the thermodynamic efficiency improvement of the sugar mill. However, the study did not consider any cogeneration improvement technology since they are currently not of interest to the South African sugar industry currently due to their high costs. However, further studies should investigate the trade-offs between the cost of integrating advanced cogeneration systems and the improvements attained in thermodynamic efficiency using exergoeconomic methodology, to establish a motivation or otherwise for their implementation or not in the South African sugar industry.

Moreover, the integration of ORC and AHP were both found to minimally improve the thermodynamic efficiency of the sugar mill even though they are economically feasible to implement. It is recommended that these technologies are optimized technically for their specific application in this study to improve their performance. On the other hand, exergoeconomic performances of integrating ORC and AHP could be studied and compared to that of more expensive energy-efficient technologies such as the advanced cogeneration technologies to select the most cost-effective integration to improve the efficiency of the sugar mill.

Furthermore, the scFOS and succinic acid processes were simulated using literature data with reasonable assumptions to model the integrated biorefineries developed in this study. Further work may include specific experimental work using A-molasses as the feedstock to establish some important assumptions such as sugars conversion to products and the purification requirements. In addition, the market dynamics of bioproducts involving the relationship between plant capacity and product global market size as well as their effect on the product price need to be investigated to confirm the economic results obtained for the integrated biorefineries in this study.

Finally, it would be insightful to evaluate the exergoeconomic performance of the sugar mill together with the integrated biorefineries compared to that of the existing sugar mill that produces only sugar. However, the challenge of assessing the economic data of an existing South African sugar mill on time limited the study in this regard. This can be explored when the data becomes available.