

# Techno-economic Analysis of Xylitol, Citric Acid and Glutamic Acid Biorefinery Scenarios Utilizing Sugarcane Lignocellulose

*by*

H. M. Raoul. Özüdoğru

Thesis presented in partial fulfilment  
of the requirements for the Degree

*of*

MASTER OF ENGINEERING  
(CHEMICAL ENGINEERING)



in the Faculty of Engineering  
at Stellenbosch University

The financial assistance of the National Research Foundation (NRF) towards this research is hereby acknowledged. Opinions expressed and conclusions arrived at, are those of the author and are not necessarily attributed to the NRF.

Supervisor

(Prof. Johann. F. Görgens)

Co-Supervisor

(Dr. Kathleen Haigh)

March 2018

## Declaration

By submitting this thesis electronically, I declare that the entirety of the work contained therein is my own, original work, that I am the sole author thereof (save to the extent explicitly otherwise stated), that reproduction and publication thereof by Stellenbosch University will not infringe any third party rights and that I have not previously in its entirety or in part submitted it for obtaining any qualification.

Date: .....

## Abstract

A decrease in global sugar prices, lower crop yields (caused by drought) and the continued use of aging and inefficient infrastructure have been listed as external factors contributing to decreasing profitability of the South African (SA) sugar milling industry. In an effort to diversify product revenues and ensure future economic competitiveness within the sugar industry, it was proposed that second-generation (2G), lignocellulosic biomass (sugarcane bagasse and trash) be valorised for the co-production of chemical products and electricity in biorefineries annexed to existing mills.

Detailed techno-economic analyses were conducted on three separate bioproduct scenarios with electricity co-generation, for a total feed capacity of 113.5 tons/h of sugarcane bagasse and trash. Products for biorefinery scenarios were selected based on potential applications, global demands and technological maturities, and included catalytic xylitol production, fermentative citric acid (CA) production (solvent extraction (SE) route) and fermentative glutamic acid (GA) production (hybrid membrane-integrated route). Detailed process simulations for each product scenario were developed using reported technology performances and Aspen Plus® software. Simulation results were used to define equipment, processes and utility requirements, and determine the total operating cost (TOC) and capital investment (TCI) for discounted cash flow (DCF) analyses at a 9.7% hurdle rate (real term basis). Profitability was characterised over a 25-year project lifespan in terms of financial parameters such as the internal rate of return (IRR), net-present (NPV), pay-back period (PBP). The economic profitability (IRR's) of product scenarios were compared to a baseline scenario with electricity-only production from combined heat and power (CHP) plant, to assess whether it was more profitable to continue burning sugarcane biomass for electricity or to convert biomass to bioproducts with electricity co-production in integrated biorefineries.

The co-production of xylitol (utilising a Raney-nickel catalyst) and electricity was achieved through a 46% bypass of the lignocellulose feed to the CHP plant. This scenario was deemed to be profitable, based on an average xylitol market price of \$4538/ton, achieving an IRR of 30% (> 9.7% minimum real term rate) and an NPV of \$405 million for a TCI of \$220 million. CA production (SE purification route) requiring a 45% biomass bypass to the CHP plant for electricity co-production, was deemed to be the least profitable, based on an average CA market price of \$1102/ton, achieving an IRR of 13%, above the minimum acceptable rate of 9.7%. A NPV of \$77 million was achieved within a 13 year pay-back period, for a TCI of \$291 million. The GA-electricity co-production biorefinery (utilizing a hybrid membrane integrated fermenter) required a 35% biomass bypass to the CHP plant, was deemed to be

the most profitable product scenario, based on an average GA price of \$3625/ton, resulting an NPV of \$866 million and an IRR of 32%. The TCI (\$422 million) was paid back over a period of 5 years.

All product scenarios (xylitol, CA and GA) were able to achieve IRR's in excess of the 8.3% achieved by the CHP baseline scenario, with electricity-only production (i.e. 100% biomass bypass). The profitability of all biorefinery scenarios were highly dependent on 3 main variables, namely the product selling price, FCI and the TOC. In addition to these variables, the xylitol biorefinery profitability was highly affected by catalyst price, whereas GA and CA profitability were sensitive to enzyme and solvent costs respectively.

In terms of environmental impact, the GA scenario consumed the most process water and produced the highest quantity of solid waste and non-biogenic carbon emissions, whereas the CHP baseline scenario had the lowest environmental impact. Although the investigated xylitol and GA scenarios were deemed profitable, the proposed production scale of xylitol (38 789 tons/annum), based on lignocellulose availability, represented 20% of the world's total xylitol demand; a reduction in the production scale to a more realistic market penetration will negatively affect profitability. The production scales for GA and CA represented much smaller and more realistic market penetration values of 3% (83 005 tons/annum) and 5% (97 893 tons/annum) of global product demand, respectively. Thus, whereas the GA scenario offered the most compelling investment case, its negative environmental impacts warrant further consideration.

## Acknowledgements

Acknowledgement and gratitude is expressed to the following parties for their various financial and academic contributions and support in delivering this thesis:

- To the Sugarcane Technology Enabling Programme for Bio-Energy (STEP-Bio), supported by the public-private partnership between the South African (SA) Department of Science and Technology's (DST) Sector Innovation Fund and the SA sugar milling industry (Sugar Milling Research Institute – SMRI) for their funding and sponsorship of this research
- The National Research Foundation (NRF) for their financial assistance towards this research
- To Professor Johann Görgens for your help in acquiring funding for this research as well as the many constructive critique and guidance you have provided to help produce this work
- To Dr Kate Haigh for all your guidance, patience and support reading my various document drafts and submissions
- To the BRTEM SMRI biorefinery research team members (Kutemba Kapanji and Mieke Nieder-Heitmann) for all your great questions, motivation and humour
- To my family and friends for all their undying support, motivation, memories and love

# I. Table of Contents

Abstract.....	ii
Acknowledgements.....	iv
I. Table of Contents.....	v
II. List of Figures.....	xii
III. List of Tables.....	xv
IV. Nomenclature.....	xix
1. Introduction.....	1
1.1 Background.....	1
1.2 Project Scope.....	2
1.3 Novelty of Study.....	4
1.4 Study Limitations.....	5
2. Literature Review.....	6
2.1 Lignocellulosic Feedstock from Sugarcane.....	6
2.1.1 Cellulose.....	7
2.1.2 Hemicellulose.....	8
2.1.3 Lignin.....	8
2.2 Pre-treatment Technologies.....	9
2.2.1 Physical pre-treatment.....	10
2.2.2 Chemical pretreatment.....	11
2.2.3 Physio-chemical pretreatment.....	15
2.2.4 Comparative Summary of Pretreatment Technologies.....	16
2.3 Enzymatic Hydrolysis.....	18
2.4 Purification and Detoxification Technologies.....	20
2.4.1 Activated carbon adsorption.....	20
2.4.2 Evaporation.....	21

2.4.3 Ion-exchange resins .....	21
2.4.4 Centrifugation .....	22
2.4.5 Membrane Extraction Filtration .....	23
2.4.6 Neutralisation .....	23
2.4.7 Summary of Detoxification Technology Applications .....	23
2.5 Products Valorised from Sugarcane Lignocellulose.....	24
2.5.1 Xylitol .....	24
2.5.2 Citric Acid .....	28
2.5.3 Glutamic Acid .....	33
2.6 Comparison of Microorganism Strains for CA and GA Fermentation .....	37
2.6.1 Citric Acid Microorganism Comparison .....	37
2.6.2 Glutamic Acid Microorganism Comparison .....	40
3. Study Objectives .....	42
4. Methodology.....	44
4.1 Software.....	44
4.2 Thermodynamic Model .....	44
4.3 Techno-economic Analysis Steps.....	44
4.4 Biorefinery Process Specifications.....	46
4.4.1 Biorefinery Lignocellulosic Feedstock.....	46
4.4.2 Catalyst and Microorganism Selection for Biorefinery Scenarios.....	46
4.4.3 Process Decisions .....	49
4.4.3 Process Assumptions .....	50
4.5 Biorefinery Scenario Descriptions .....	51
4.5.1 Standalone CHP Baseline Scenario .....	51
4.5.2 Xylitol and Electricity Co-production Scenario.....	54
4.5.3 Citric Acid (CA) and Electricity Co-production Scenario.....	68
4.5.4 Glutamic Acid (GA) and Electricity Co-production Scenario .....	79
4.6 Economic Evaluation.....	84

4.6.1 Capital Cost Estimation .....	84
4.6.2 Operating Cost Estimation .....	86
4.6.3 Cash Flow Sheet Assumptions .....	89
5. Results and Discussion .....	92
5.1 Utility and Bioproduct Production.....	92
5.1.1 Xylitol Equipment Utility Consumption.....	92
5.1.2 CA Equipment Utility Consumption .....	94
5.1.3 GA Equipment Utility Consumption.....	97
5.2 Bioproduct and Electricity Production.....	99
5.2.1 Product Quantity and Purity Comparison.....	99
5.2.2 Electricity Production Comparison .....	100
5.3 Operating Expenditure .....	101
5.3.1 Xylitol OPEX.....	101
5.3.2 Citric Acid (CA) Production Scenario OPEX .....	102
5.3.3 GA OPEX .....	104
5.3.4 OPEX Comparison (Including CHP Baseline Scenario).....	106
5.4 Capital Expenditure .....	106
5.4.1 Xylitol Biorefinery CAPEX .....	106
5.4.2 Citric Acid Biorefinery CAPEX.....	108
5.4.3 Glutamic Acid Biorefinery CAPEX.....	110
5.4.4 CAPEX Comparison (Including CHP Baseline Scenario).....	112
5.5 Financial Performance of Biorefinery Scenarios.....	113
5.5.1 Profitability Indicators (NPV, IRR, TCI, PBP) .....	113
5.5.2 Economic Comparison between different Biorefinery Scenarios.....	115
5.6 Cost Sensitivity Analysis of Scenarios .....	116
5.6.1 Xylitol Biorefinery Cost Sensitivity Analysis .....	116
5.6.2 CA Biorefinery Cost Sensitivity Analysis.....	117
5.6.3 GA Biorefinery Cost Sensitivity Analysis .....	118



5.6.4 Summary .....	119
5.7 Environmental Impact of Biorefinery Scenarios .....	119
5.8 Viability of Results in Comparison to Existing Facilities.....	121
5.8.1 Biorefinery Capacity and Market Representation .....	121
5.8.2 Techno-economic Comparison to Previous Studies .....	122
6. Conclusion.....	125
7. Recommendations and Future Scope.....	126
8. References .....	128
Appendix A. Xylitol Calculation Summary.....	142
A1. Xylitol Component Summary .....	142
A2. Xylitol Aspen Simulation Process Flow Diagrams.....	143
A2.1. Overall Xylitol Biorefinery.....	143
A2.2. Area 100 Pretreatment (DA, Neutralization, Washing) .....	143
A2.3. Area 200 Detoxification (Solids Separation and Sugar Washing Recovery) .....	144
A2.4. Area 300 Catalytic Reaction.....	144
A2.5. Area 400 Purification and Drying.....	145
A2.6. Area 500 Combined Heat and Power Plant .....	145
A2.7. Area 600 Wastewater Treatment Plant.....	146
A3. Xylitol Mass and Energy Balance Streams.....	147
A4. Xylitol Utility Consumption Summary .....	150
A5. Xylitol Equipment Sizing Summaries .....	152
A5.1 Pumps .....	152
A5.2. Conveyors .....	153
A5.3. Compressors, Blowers and Turbines .....	154
A5.4. Heat Exchangers .....	154
A5.5. Separators and Filters.....	155
A5.6. Reactors and Vessels .....	157
A5.7. Agitators .....	158

A6. Xylitol Equipment Cost Summary .....	159
A7. Xylitol Pricing from Literature and Industry .....	161
A8. Xylitol Discount Cash Flow Sheet .....	163
A9. Xylitol Cumulative Cash Flow Sheet .....	166
Appendix B. Citric Acid (CA) Calculation Summary .....	167
B1. CA Component Summary .....	167
B2. Citric Acid Aspen Simulation Process Flow Diagrams.....	170
B2.1 Overall CA Biorefinery - Solvent Extraction Route.....	170
B2.2 Area 100 Pretreatment .....	170
B2.3. Area 200 Enzymatic Hydrolysis .....	171
B2.4. Area 300 Fermentation.....	171
B2.5. Area 400 Purification .....	172
B2.6. Area 500 Combined Heat and Power Plant .....	172
B2.7. Area 600 Water Treatment Plant .....	173
B2.8. Area 700 Candida Oleophila Seed Train .....	173
B3. CA Mass and Energy Balance Streams.....	174
B4. CA Utility Consumption Summary .....	178
B5. CA Equipment Sizing Summaries .....	181
B5.1. Pumps .....	181
B5.2. Conveyors .....	182
B5.3. Compressors, Blowers and Turbines .....	182
B5.4. Heat Exchangers .....	183
B5.5. Separators and Filters .....	184
B5.6. Vessels and Reactors.....	185
B5.7. Agitators.....	186
B6. CA Equipment Cost Summary.....	188
B7. CA Pricing from Literature and Industry.....	190
B8. CA Discount Cash Flow Sheet .....	192

B9. CA Cumulative Cash Flow Sheet .....	195
Appendix C. Glutamic Acid (GA) Calculation Summary .....	196
C1. GA Component Summary .....	196
C2. Glutamic Acid Aspen Simulation Process Flow Diagrams.....	198
C2.1. Overall GA Biorefinery - Hybrid Membrane Reactor System .....	198
C2.2. Area 100 Pretreatment .....	198
C2.3. Area 200 Enzymatic Hydrolysis .....	199
C2.4. Area 300 Fermentation .....	199
C2.5. Area 400 Purification .....	200
C2.6. Area 500 CHP Plant .....	200
C2.7. Area 600 WWTW .....	201
C2.8. Area 700 Brevibacterium Seed Train .....	201
C3. GA Mass and Energy Balance Streams .....	202
C4. GA Utility Consumption Summary.....	206
C5. GA Equipment Sizing Summaries.....	208
C5.1. Pumps .....	208
C5.2. Conveyors .....	209
C5.3. Compressors, Blowers and Turbines.....	210
C5.4. Heat Exchangers.....	210
C5.5. Separators and Filters .....	212
C5.6. Vessels and Reactors.....	213
C5.7. Agitators.....	215
C5. GA Equipment Cost Summary .....	216
C6. GA Pricing from Literature and Industry .....	219
C7. GA Discount Cash Flow Sheet.....	220
C8. GA Cumulative Cash Flow Sheet.....	223
Appendix D. CHP Baseline Aspen Simulation Process Flow Diagram .....	224
Appendix E. Equipment Purchase Cost Reference Sheet .....	225

Appendix F. Feed and Cleaning Schedules.....	227
F1. Xylitol Reactor Feed and Cleaning Schedule.....	227
F2. CA Fermenter Feed and Cleaning Schedule.....	227
F3. GA Fermenter Feed and Cleaning Schedule .....	227

## II. List of Figures

Figure 1 – Investigated Biorefinery Scenarios Annexed to Existing Sugar Mills (Xylitol, CA and GA with Electricity Co-production) .....	3
Figure 2 - Sugarcane Lignocellulosic Structural Breakdown from Lignocellulose to Sugars redrawn from ( Wyman & Yang, 2009); (Mussatto & Teixeira, 2010); (Bhayani & Ramarao, 2013)) .....	6
Figure 3 - Three Major Lignin Polymer Components (permission obtained from (Fisher & Fong, 2014) under Creative Commons Attribution License) .....	9
Figure 4- Hydrolysis of cellulose onto cellulase active site into monomeric glucose redrawn from (Quiroz-Castañeda & Folch-Mallol, 2013; Mussatto & Teixeira, 2010) .....	19
Figure 5 - Different Types of Inhibitors Formed during Pretreatment (redrawn with permission from (Jönsson, et al., 2013)) .....	20
Figure 6 - Xylitol structural formula .....	25
Figure 7 - Hydrolysis and Hydrogenation Steps in Xylitol Production, redrawn from (Tomishige, et al., 2016) .....	26
Figure 8 - Citric Acid Anhydrous Structural Formula .....	28
Figure 9 –Simplified Citric Acid Cycle (redrawn from (Shuler & Kargi, 2002); (Ciriminna, et al., 2017)) .....	31
Figure 10 - Glutamic acid chemical structure .....	33
Figure 11 - Cell permeability of glutamic acid in relation to phospholipids in the cell, redrawn from (Kumar, et al., 2014) .....	35
Figure 12 - Regulatory pathway for the biosynthesis of glutamic acid redrawn and adapted from (Kumar, et al., 2014) .....	36
Figure 13 - Steps Required to Perform Economic Analyses.....	45
Figure 14 - CA microorganism weighted desirability for batch and continuous production.....	47
Figure 15 - GA microorganism weighted desirability for batch and continuous production .....	47
Figure 16 - CHP Baseline Flow Sheet.....	53
Figure 17 - Xylitol Biorefinery Pretreatment Section (Area 100) .....	56
Figure 18 - Xylitol Detoxification Section (Area 200) .....	58
Figure 19 - Xylitol Catalytic Reaction and Purification Section (Area 300 & Area 400) .....	60
Figure 20 - Xylitol WWT Plant (Area 600) .....	64
Figure 21 - Xylitol CHP Plant (Area 500).....	67
Figure 22 - CA Pretreatment Section (Area 100) .....	69
Figure 23 - CA Enzymatic Hydrolysis Section (Area 200) .....	71
Figure 24 - CA Fermentation Section (Area 300) .....	74

Figure 25 - CA Purification Section (Area 40).....	75
Figure 26 - CA CHP Plant Section (Area 500) .....	76
Figure 27 - CA WWT Plant Section (Area 600) .....	77
Figure 28 - CA Seed Train Section (Area 700) .....	78
Figure 29 - GA Pretreatment Section (Area 100) .....	81
Figure 30 - GA Enzymatic Hydrolysis Section (Area 200) .....	82
Figure 31 - GA Fermentation and Purification Sections (Area 300 and 400).....	83
Figure 32 - Xylitol Electricity Distribution by Plant Area .....	92
Figure 33 - Xylitol Equipment Electricity Distribution (Top 7 Consumers) .....	93
Figure 34 - CA Electricity Distribution by Plant Area.....	95
Figure 35 - CA Equipment Electricity Distribution .....	95
Figure 36 - GA Electricity Distribution by Plant Area .....	97
Figure 37 -GA Equipment Electricity Distribution .....	98
Figure 38 – Comparison of Bioproduct Production (Xylitol, CA, GA, and Baseline) .....	99
Figure 39 - Comparison of Electricity Production (Xylitol, CA, GA, and Baseline) .....	100
Figure 40 - Xylitol Variable Operating Cost Distribution.....	101
Figure 41 - CA Variable Operating Cost Distribution .....	103
Figure 42 - GA Variable Operating Cost Distribution .....	105
Figure 43 – Xylitol Equipment Cost Distribution per Area .....	107
Figure 44 - CA Capital Cost Distribution.....	109
Figure 45 - GA Capital Cost Distribution .....	111
Figure 46 - Xylitol Profitability Range for Max, Ave and Min Xylitol Prices .....	114
Figure 47 - CA Profitability for different CA Prices (Max, Ave and Min).....	114
Figure 48 - GA Profitability Range for Max, Ave and Min GA Prices.....	115
Figure 49 - Comparison of Key Profitability and Investment Indicators among Biorefinery Scenarios .....	116
Figure 50 - Xylitol Sensitivity Analysis.....	117
Figure 51 - CA Sensitivity Analysis.....	118
Figure 52 - GA Sensitivity Analysis (% Change in IRR).....	118
Figure 53 - Environmental Impact of Biorefinery Scenarios (Carbon Emissions, Solid Waste and Water Consumption).....	120
Figure 54 - Representation of Current Production to Total World Demand .....	122
Figure 55 - Aspen Xylitol Biorefinery Flowsheet .....	143
Figure 56 - Aspen Xylitol Pretreatment Flowsheet (Area 100) .....	143

Figure 57 - Aspen Xylitol Detoxification Flowsheet (Area 200) .....	144
Figure 58 - Aspen Xylitol Catalytic Reactor Flowsheet (Area 300).....	144
Figure 59 - Aspen Xylitol Purification and Drying Flowsheet (Area 400) .....	145
Figure 60 - Aspen Xylitol CHP Plant Flowsheet (Area 500) .....	145
Figure 61 - Aspen Xylitol WWT Flowsheet (Area 600) .....	146
Figure 82 - Xylitol Cumulative Cash Flow .....	166
Figure 62 - Aspen CA Biorefinery Flowsheet.....	170
Figure 63 - Aspen CA Pretreatment Flowsheet (Area 100).....	170
Figure 64 - Aspen CA Enzymatic Hydrolysis Flowsheet (Area 200).....	171
Figure 65 - Aspen CA Fermentation Flowsheet (Area 300).....	171
Figure 66 - CA Aspen Purification Flowsheet (Area 400) .....	172
Figure 67 - CA Aspen CHP Plant Flowsheet (Area 500).....	172
Figure 68 - CA Aspen WWT Flowsheet (Area 600).....	173
Figure 69 - CA Aspen Candida oleophila Seed Train Flowsheet (Area 700).....	173
Figure 83 - CA Cumulative Cash Flow .....	195
Figure 70 - Aspen GA Biorefinery Flowsheet .....	198
Figure 71 – Aspen GA Pretreatment Flowsheet .....	198
Figure 72 – Aspen GA Enzymatic Hydrolysis Flowsheet (Area 200).....	199
Figure 73 - Aspen GA Fermentation Flowsheet (Area 300) .....	199
Figure 74 – Aspen GA Purification Flowsheet (Area 400).....	200
Figure 75 – Aspen GA CHP Plant Flowsheet (Area 500).....	200
Figure 76 – Aspen GA WWT Plant Flowsheet (Area 600) .....	201
Figure 77 – Aspen GA Brevibacterium Seed Train Flowsheet (Area 700).....	201
Figure 84 - GA Cumulative Cash Flow .....	223
Figure 78 - CHP Baseline Aspen PFD .....	224
Figure 79 - Xylitol Catalytic Reactor (x3) feed and Cleaning Schedule (half-hour schedule).....	227
Figure 80 - CA Fermenter (x3) Feed and Cleaning Schedule (half-day schedule).....	227
Figure 81 - GA Fermenter (x8) Feed and Cleaning Schedule (hourly schedule) .....	227

### III. List of Tables

Table 1 – Average Polysaccharide Distribution for Different Varieties of SA Bagasse and Trash (Görgens, et al., 2015; Petersen, et al., 2014) .....	7
Table 2 - Hemicellulose hydrolysate composition for different feedstock using dilute acid pre-treatment, adapted from (Chandel, et al., 2012) .....	12
Table 3 - Effect of different pretreatment technologies on the structure of lignocellulose adapted from ( Mosier, et al., 2005); (Alvira, et al., 2010)) .....	17
Table 4 - Capital cost of five pretreatment processes adapted from (Eggeman & Elander, 2005) .....	17
Table 5 - Summary of conversion data for six different pretreatment technologies using switch grass as the lignocellulosic feed adapted from (Tao, et al., 2011).....	18
Table 6 - Enzymatic Hydrolysis Reactions and Conversions for Cellulosic material adapted from (Humbird, et al., 2011).....	19
Table 7 - Advantages and disadvantages of different detoxification methods of hemicellulosic hydrolysate adapted from (Canilha, et al., 2012) .....	24
Table 8 - Main and side reactions for the hydrogenation of xylose adapted from ( Mikkola, et al., 2003); (Yadav, et al., 2011)) .....	27
Table 9 - Main stoichiometric reactions for citric acid production via fermentation adapted from (Heinzle, et al., 2006) .....	30
Table 10 - Comparison of Various Microorganisms for the Production of Citric Acid.....	38
Table 11 - Comparison of Various Microorganisms for the Production of Glutamic Acid (Concentration, Productivity, Yield).....	40
Table 12 - Biorefinery Scenarios Investigated.....	42
Table 13 - Composition of Lignocellulosic Bagasse and Trash Feedstock for Biorefinery Scenarios (Petersen, et al., 2014).....	46
Table 14 - Comparison of Xylitol Selectivity and Yields Using Different Catalysts adapted from (Yadav, et al., 2011) .....	48
Table 15 - CHP Plant Baseline Combustion Reactions .....	51
Table 16 - DA Hydrolysis Chemical Reactions and Conversions (Humbird, et al., 2011) .....	55
Table 17 - Catalytic Reactions of Xylose and Arabinose to Xylitol and Arabinitol .....	59
Table 18 - Anaerobic Digestion Reactions ( Bilitewski, et al., 1997); (Kalyuzhnyi, 1997); (Ostrem, 2004); (Rajendran, et al., 2014)) .....	62
Table 19 - Aerobic Digestion Reactions (Tchobanoglous, et al., 2004).....	63
Table 20 - Boiler Combustion Reactions (Görgens, et al., 2015) .....	66
Table 21 - Enzymatic Hydrolysis Reactions and Conversions (Humbird, et al., 2011).....	70



Table 22 - CA fermentation medium sugar and nutrient concentration, table adapted from (Anastassiadis & Rehm, 2006) .....	72
Table 23 - CA Fermenter Stoichiometric Reactions (Heinzle, et al., 2006).....	73
Table 24 – GA Fermentation Feed Medium Concentrations (Miesher & Haute, 1975) .....	79
Table 25 - GA Chemical Reactions ( (Lederberg, 2000); (Delaunay, et al., 2002); (Pandey, et al., 2015)) .....	80
Table 26 - Equipment Capacity Scaling Components (Humbird, et al., 2011) .....	84
Table 27 - Equipment Installation Factors (Humbird, et al., 2011).....	85
Table 28 - CEPCI Values for Different Years .....	86
Table 29 - Chemical Consumables Cost Summary for Xylitol Biorefinery Scenario (Görgens, et al., 2015) .....	86
Table 30 - Chemical Consumables Cost Summary for CA Biorefinery Scenario (Anastassiadis & Rehm, 2006) .....	87
Table 31 - Chemical Consumables Cost Summary for GA Biorefinery Scenario (Pal, et al., 2016; Miesher & Haute, 1975).....	88
Table 32 - Waste Removal Costs.....	88
Table 33 - Staff Salary Summary (Görgens, et al., 2015) .....	89
Table 34 - Maintenance, Insurance and Tax Operational Costs (Görgens, et al., 2015) .....	89
Table 35 – Main Cash Flow Sheet Assumptions ( (Görgens, et al., 2015); (Farzad, et al., 2017)) .....	90
Table 36 –Summary of Electricity Consumption According to Plant Sections for Xylitol Biorefinery ..	92
Table 37 – Summary of Highest Electricity Consumers for Xylitol Biorefinery .....	93
Table 38 - Summary of LPS Consumed by Equipment for Xylitol Biorefinery.....	94
Table 39 - Summary of Cooling Water an Air Consumed by Equipment for Xylitol Biorefinery.....	94
Table 40 - Summary of Electricity Distribution among Plant Sections for Xylitol Biorefinery .....	95
Table 41 - Summary of Highest Electricity Consumption among Equipment for CA Biorefinery .....	96
Table 42 - Summary of LPS Consumed by Equipment for CA Biorefinery .....	96
Table 43 - Summary of Cooling Water Consumed by Equipment for CA Biorefinery.....	96
Table 44 - Summary of Electricity Usage among Plant Sections for GA Biorefinery.....	97
Table 45 - Summary of Highest Electricity Consumption among Equipment for GA Biorefinery.....	98
Table 46 - Summary of LPS Consumed by Equipment for GA Biorefinery .....	98
Table 47 - Summary of Cooling Water Consumed by Equipment for GA Biorefinery .....	99
Table 48 - Xylitol Variable Operating Costs.....	102
Table 49 - Xylitol TOC Summary.....	102
Table 50 - CA Variable Operating Costs .....	103

Table 51 - CA TOC Summary .....	104
Table 52 - GA Variable Operating Costs.....	105
Table 53 - GA TOC Summary .....	105
Table 54 - OPEX and Revenue Comparison Summary (\$million/year) .....	106
Table 55 - Installed Cost Distribution (Xylitol-Electricity Co-production).....	107
Table 56 - Xylitol Capital Expenditure Summary.....	108
Table 57 - Installed Cost Distribution (CA-Electricity Co-production Scenario).....	109
Table 58 - CA Capital Expenditure Summary .....	110
Table 59 - Installed Cost Distribution (GA-Electricity Co-production Scenario) .....	111
Table 60 - GA Capital Expenditure Summary.....	112
Table 61 - Comparison of Capital Investment (\$ million) .....	112
Table 62 – Weighted profitability and environmental composite score .....	120
Table 63 - Citric Acid Production Facilities from Different Locations ( (SRI Consulting, 2000) (Ciriminna, et al., 2017)) .....	121
Table 64 - Biorefinery Market Representation .....	122
Table 65 - Economic Comparison of Present Xylitol Scenario to Past Studies .....	123
Table 66 - GA Economic Comparison to Past Studies (Pal, et al., 2015).....	124
Table 67 - Xylitol Aspen Component Summary .....	142
Table 70 - Xylitol Utility Summary.....	150
Table 74 - Xylitol Equipment Sizing and Cost Summary.....	159
Table 81 - Xylitol Pricing According to Literature .....	161
Table 82 - Xylitol Pricing by Industry and Country.....	162
Table 77 - Xylitol 25-Year Cumulative Cash Flow.....	166
Table 68 - CA Aspen Component Summary.....	167
Table 71 - CA Utility Summary .....	178
Table 75 - CA Equipment Sizing and Cost Summary .....	188
Table 83 - CA Pricing According to Literature .....	190
Table 84 - CA Pricing by Industry and Country .....	191
Table 78 - CA 25-Year Cumulative Cash Flow .....	195
Table 69 - GA Component Summary .....	196
Table 72 - GA Utility Summary.....	206
Table 76 - GA Equipment Sizing and Cost Summary.....	216
Table 85 - GA Pricing According to Literature.....	219
Table 79 - GA 25-Year Cumulative Cash Flow .....	223

Table 73 - Equipment Purchase Costs for Common Plant Equipment (adapted from (Towler & Sinnott, 2012)) ..... 225

## IV. Nomenclature

Acronym, Symbol or Term	Description or Definition
SA	South African
CHP	Combined Heat and Power
NCD	Non-communicable diseases
PFD	Process Flow Diagram
SSF	Solid State Fermentation
SE	Steam Explosion
GA	Glutamic acid
CA	Citric acid
SLE	Solid-Liquid Extraction
5-HMF	5-Hydroxymethyl Furfural
DP	Degree of polymerisation
Bagasse	Agricultural residue formed after squeezing cane
Lignocellulose	Component of cell wall in plants (consists of cellulose, hemicellulose and lignin)
Cellulose	Crystalline polysaccharide consisting of long chain glucose monomers
Hemicellulose	Non-crystalline polysaccharide found in lignocellulose consisting of various pentosan and hexosan sugar monomers
Lignin	Random structure poly-alcohol found in lignocellulose consisting of grouped aromatic compounds
Polymer	Long-chain compound composed of repeated chemical units called monomers
Oligomer	Shorter chain polymer with a defined degree of polymerisation that can be further broken down into monomers
Monomer	Individual molecules that may combine to form longer-chain polymers
5-HMF	5-hydroxymethyl-furfural
H-Unit	refers to p-hydroxyphenyl oligomer unit in lignin
S-Unit	refers to syringyl oligomer unit in lignin
G-Unit	refers to guaiacyl oligomer unit in lignin
Pentosans	5 carbon atom monomer units found in hemicellulose
Hexosans	6 carbon atom monomer units found in cellulose and hemicellulose polymers
DCFROR/IRR	Internal Rate of Return or Discounted Cash Flow Rate on Return. The rate for which NPV = 0
Hurdle Rate	Minimum internal rate for which a project is accepted as being profitable by investors
NPV	Net Present Value
PBP	Pay-Back-Period (Time period before an investment is recovered)
ELEC-NRTL	Electrolyte non-random two-liquid activity coefficient model
IE	Ion exchange
WBA	Weak basic anionic
SAC	Strong acid cationic
CASP	Citric acid selling price
GASP	Glutamic selling price
XYSP	Xylitol selling price

# 1. Introduction

## 1.1 Background

The South African (SA) sugar milling industry is currently experiencing diminishing profitability due to a number of external factors, including lower crop yields due to recent severe drought conditions (as a symptom of climate change), increased labour costs, aging infrastructure and a decrease in global sugar prices (USDA, Economic Research Service, 2017; AgriSA, 2016; Zhao & Li, 2015). In addition, announcements have recently been made to enact a new sugar tax in SA, aimed at reducing non-communicable diseases (NCDs) that have been linked to excess sugar consumption, which would consequently reduce demand for sugar-based products and further threaten profit margins made by the SA sugar industry (Singh, et al., 2015; SA National Department of Health, 2013).

Lignocellulose is one of the most promising sources of renewable biomass for the emerging global bio-economy (a second generation or 2G biomass source), due to its relative abundance and lower cost, and reduced competition for land use, in comparison to first generation (1G) feedstocks typically grown on land allocated for food production (Farzad, et al., 2017; Quintero, et al., 2013). Sugar milling produces large amounts of agro-processing waste in the form of bagasse, a fibrous pulp produced after extracting the juice from sugarcane (Lavarack, et al., 2002), as well as tops and leaves (termed trash), which are left in the sugarcane fields after harvest. To mitigate the costs associated with collecting and transporting the bulky residues to the sugar mill, the trash is typically left in the field or burnt before harvest (Smithers, 2014). This practice releases large amounts of gaseous emissions and soot into the air that can enter the lungs of labourers and neighbouring communities in the area and lead to respiratory health problems (Leal, et al., 2013).

One approach that the sugar milling industry has currently adopted to lower production costs and extract value from wastes, is to burn sugarcane bagasse and trash in low pressure boilers, in the place of fossil fuels for electricity and steam production. This is used to power and heat the various process operations utilised during sugar production (Dias, et al., 2011). An alternative direction by which sugar mills may sustain future economic viability, is by expanding their product range to include the production of alternative fuels or chemical products from the valorisation of lignocellulosic waste. Lignocellulose has been studied extensively in the past for the production of bio based fuels such as ethanol and butanol (Naleli, 2016; Bensah & Mensah, 2013; Canilha, et al., 2012; Chandel, et al., 2011; Humbird, et al., 2011; Sun & Cheng, 2002). However, recent studies have increasingly been exploring the conversion of lignocellulosic feedstocks to value-added bioproducts (Farzad, et al., 2017; Zhou, et al., 2017; Mountraki, et al., 2017; Yamaguchi, et al., 2016; Görgens, et al., 2015).

Xylitol, citric acid and glutamic acid bioproducts have previously been selected based on potential application, demand and technological production maturity (Kapanji, 2016; Davis, et al., 2015; Werpy, et al., 2004). The sugar-alcohol, xylitol, was listed as a promising chemical product candidate due to its unique health benefits and numerous applications within the food and pharmaceutical sector, as well as being a chemical building block for various other potential products (Grumezescu & Holban, 2017; Mohamad, et al., 2015; Werpy, et al., 2004). Therefore, a strong demand for xylitol exists with world demand estimated by some sources to be 125 000 tons per annum with a bulk market price ranging from \$4-5 per kg (Grumezescu & Holban, 2017). However, recent reports estimate a global market of 190 900 tons per annum, with market demand expected to grow annually by 5.7% and reach 266 500 tons per annum by 2022 (Industry Experts, 2017). Citric acid (CA) has also been selected as a product candidate and finds mass use across different sectors, including the food, pharmaceutical, cosmetic, detergent and cleaning industry (Ciriminna, et al., 2017; Angumeenal & Venkappayya, 2013; Soccol, et al., 2006; Petrides, 2000; Vandenberghe, et al., 1999). The current world demand of CA is estimated to be 2-2.3 million tons per annum with a 4% annual growth (Ciriminna, et al., 2017; Expert Market Research, 2017). Future demand is expected to reach 2.8 million tons per annum by 2022 (Expert Market Research, 2017). Glutamic acid (GA) is another candidate product that can be valorised from sugars liberated from lignocellulosic feedstock, and has applications within the food and pharmaceutical industry (Kumar, et al., 2014; Vishwanath, 2006; Heys & Ashkanani, 1999). It has a current estimated global market size of 2.9 million tons per annum with an annual growth of 7.5% and is estimated to reach a global demand of 4 million tons per annum by 2023 (Global Market Insights, Inc., 2016). The next section details the scope of this research project.

## 1.2 Project Scope

In order to ensure future economic competitiveness and environmental sustainability within the SA sugar milling industry, it is proposed that lignocellulosic waste and crop residues (sugarcane bagasse and trash) be valorised to produce bioproducts. This is achieved by annexing integrated product biorefineries with electricity co-generation to existing sugar mills.

The aim of this study is to develop techno-economic analyses for three different bioproduct and electricity co-generation scenarios that are annexed to adjacent sugar mills. Product scenarios investigated, include catalytic xylitol production, fermentative citric acid (CA) production and glutamic acid (GA) production from waste sugarcane bagasse and trash, all of which have some amount of electricity co-production and sale (Figure 1). The profitability of each product biorefinery is compared to a combined heat and power (CHP) plant baseline scenario, where 100% of the biomass is utilised for electricity and steam production in a newer, more efficient high pressure boiler.(Figure 1).

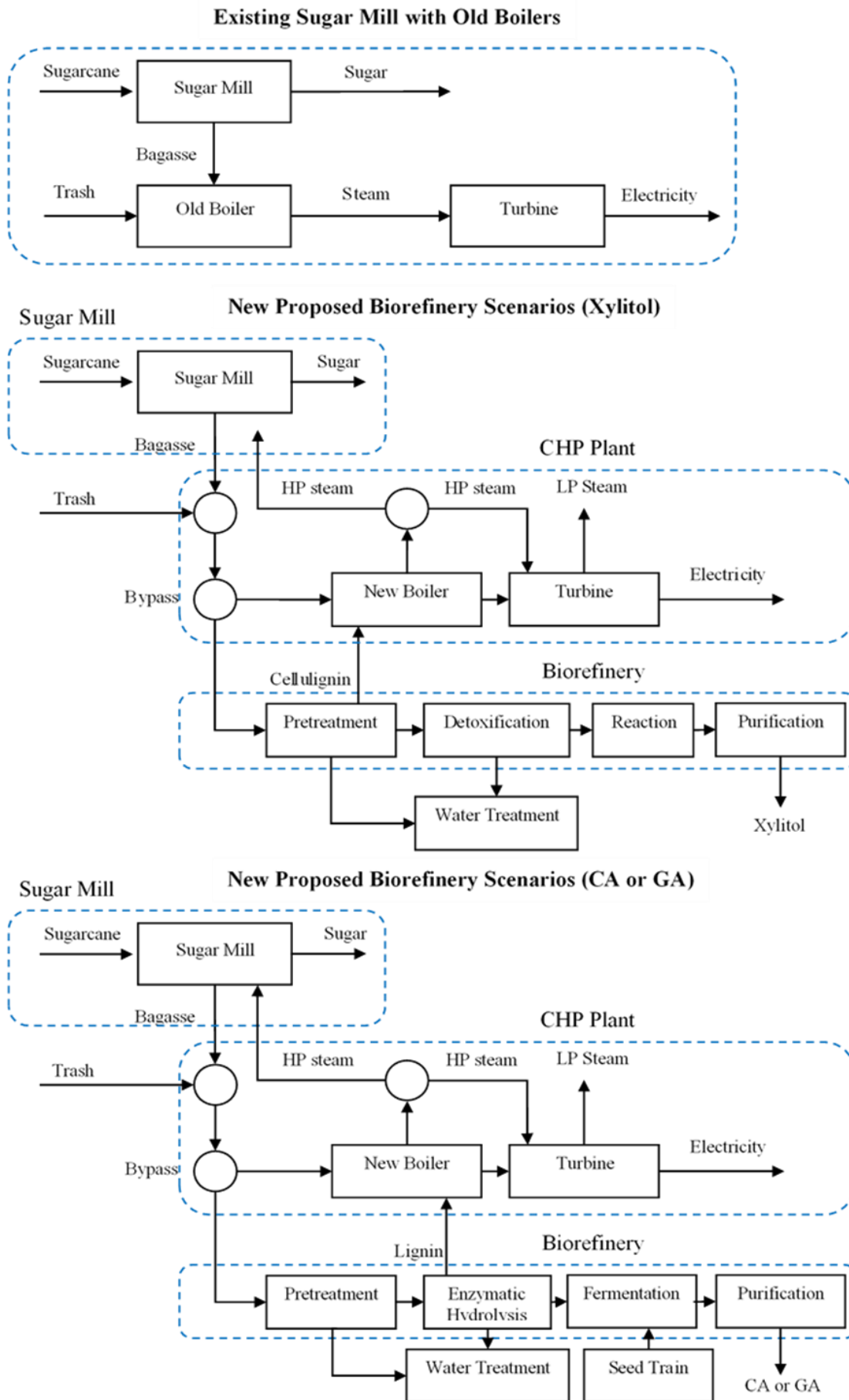


Figure 1 – Investigated Biorefinery Scenarios Annexed to Existing Sugar Mills (Xylitol, CA and GA with Electricity Co-production)

Detailed process models for each product scenario were developed using data reported in literature and simulated in Aspen Plus®. Simulation results were used to define equipment, processes and utility requirements and determine the total operating costs (TOC) and capital investment (TCI) for discounted cash flow analysis. Profitability is assessed in terms of financial parameters such as the internal rate of return (IRR), net present value (NPV) and pay-back-period (PBP).

This work is investigated to determine the most financially viable product scenarios and investigate whether it is more economically profitable to burn sugarcane lignocellulose for electricity-only revenue or valorise it to chemical products in product biorefineries with electricity co-generation.

### 1.3 Novelty of Study

Simplified economic analyses on fermentative xylitol production, have previously been reported for biorefineries utilising beer spent grain (BSG) and olive stone lignocellulosic feedstock. Basic profitability results were reported in terms of the cost of xylitol production and total annual operating costs (TOC) (Hernández, et al., 2014; Mussatto, et al., 2013). However, these studies did not conduct a detailed discounted cash flow (DCF) analysis and neglect to report profitability in terms of comparative economic indicators (IRR, NPV and PBP). A more detailed DCF analysis for a standalone model (no integration into an existing mill) for fermentative xylitol production was undertaken for sugarcane bagasse feedstock, but results were limited to TCI and IRR and did not report the NPV or TOCs achieved over the project's short 5 year life-span (Clauser, et al., 2016). A recent techno-economic study rigorously modelled and compared the economic performance of xylitol biorefineries utilising biotechnological and chemical catalytic routes, using hydrolysed C5 sugar syrup as feedstock. Profitability results were reported in terms of TCI, NPV and PBP, but did not report the IRR (Mountraki, et al., 2017). Work presented in the current study uses rigorous DCF analysis to evaluate the profitability (in terms of TCI, NPV, IRR and PBP) of a catalytic xylitol biorefinery with electricity co-production, utilising second generation lignocellulosic sugarcane bagasse and trash annexed to an existing sugar mill.

To the best of the author's knowledge, no current literature sources have reported techno-economic analyses on CA and GA integrated product biorefineries with electricity co-generation utilising second generation lignocellulosic feedstock (sugarcane bagasse and trash). Past studies have assessed the economic performance of GA production utilising first generation renewable feedstock (sugarcane juice) in membrane-integrated hybrid reactors (Pal, et al., 2016; Pal, et al., 2015). However, economic profitability in these studies is often reported in a simplistic or obscure manner in terms of operating costs per mass of product or in terms of return on investment (ROI). Process modelling with rigorous DCF analysis has not been undertaken for these product biorefineries (CA and GA). The CA industry is specifically a highly competitive and secretive



industry, with proprietary registered organisms and processes and therefore studies on economic profitability are limited or currently unavailable in academic literature. Solvent extraction is considered a promising alternative purification process in comparison to classical precipitative methods and has not been investigated for CA production and is covered in the present work.

#### 1.4 Study Limitations

Simulating product biorefineries in Aspen Plus® provides the advantage of being able to rapidly generate detailed technical descriptions for process unit operations and adjust parameters for process optimisation. Simulations for this work have been generated using data collected from literature. However, in some instances detailed data is often restricted for certain individual process units (e.g. purification and recovery equipment) and is instead reported for an overall process. Therefore, a number of technical assumptions must often be made to provide a basis on which the simulations can be developed. In addition, lignocellulosic biomass is a complex feedstock, that is composed of solid and liquid fractions, as well as numerous chemical polymers, some of which need to be represented using alternative compounds with similar physical properties in simulations (Görgens, et al., 2015).

In terms of economic analyses, it can be challenging acquiring updated cost data for certain chemicals and uncommon types of process equipment. Equipment vendors typically do not disclose prices to individuals from academic institutions and past literature sources may not fully characterise the true cost of the process equipment. All these combined factors can put limitations on the models ability to accurately represent certain production steps in real-world biorefineries.

Chapters presented further will elaborate on current established literature necessary to develop a biorefinery solution for each of the product scenarios investigated. Production methods from pre-treatment to purification are reviewed to establish a design basis for each biorefinery scenario (xylitol, citric acid, glutamic acid) (see chapter 2). Concise problem statements and objectives are provided along with an explanation of the tools and methods required to simulate scenarios in Aspen Plus® and ultimately produce economic analyses for each product scenario investigated (chapters 3 and 4). Thereafter, economic results are presented and discussed, along with a conclusion on the profitability of each biorefinery scenario relative to the CHP plant baseline scenario (chapters 5 and 6).

## 2. Literature Review

In order to model product biorefineries and determine their respective economic feasibility within an acceptable level of accuracy, a thorough review and understanding of the lignocellulosic feedstock, pre-treatment methods, detoxification methods and production steps is covered in this chapter.

### 2.1 Lignocellulosic Feedstock from Sugarcane

Sugarcane bagasse and trash are two abundant sources of lignocellulosic biomass produced during sugar production and harvesting respectively (Canilha, et al., 2012; Lavarack, et al., 2002). Lignocellulose is mainly composed of the polysaccharides cellulose and hemicellulose, as well as a binding substance that gives rigid structure to the cell walls in the bagasse and leaves, a polyalcohol named lignin (Bhayani & Ramarao, 2013; Mussatto & Teixeira, 2010) (Figure 2). The ratio of these different polysaccharides and components are dependent on the chosen feedstock.

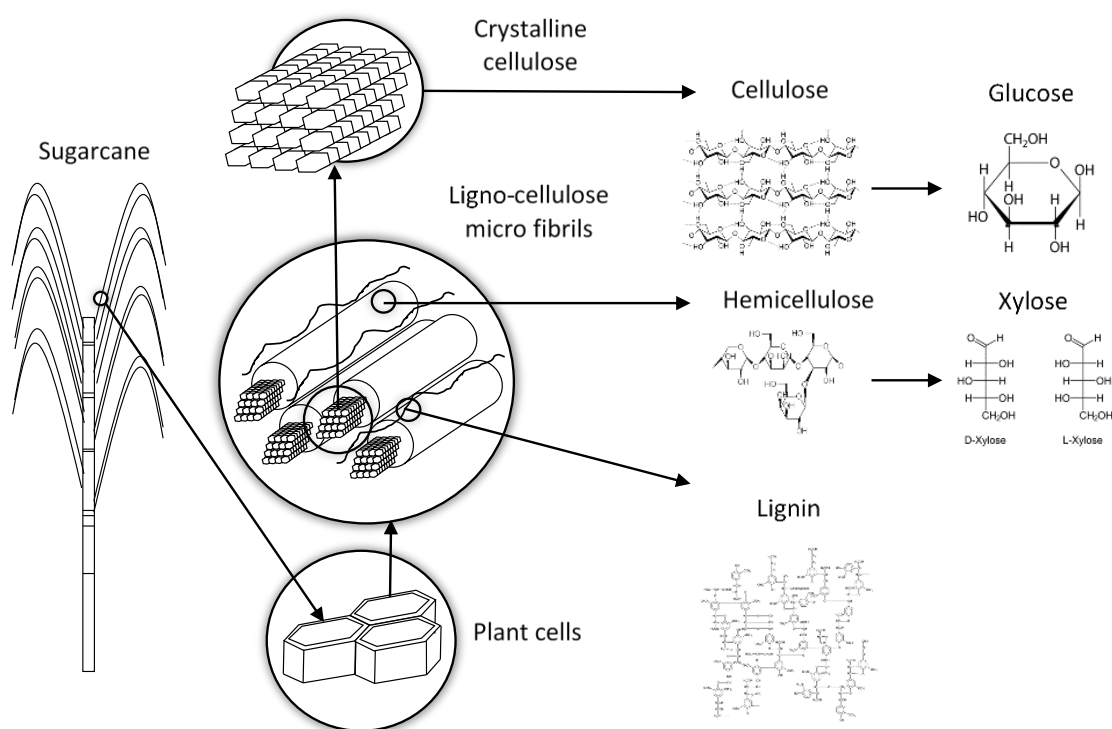


Figure 2 - Sugarcane Lignocellulosic Structural Breakdown from Lignocellulose to Sugars redrawn from (Wyman & Yang, 2009); (Mussatto & Teixeira, 2010); (Bhayani & Ramarao, 2013))

After milling, bagasse typically retains 50 % (w/w) moisture (Petersen, et al., 2014). On a dry mass basis, sugarcane bagasse has cellulose, hemicellulose and lignin contents typically ranging from 39-43%, 25-27% and 20-25% respectively (Carvalho, 2009). Other studies have reported values of 40%, 24% and 25% for cellulose, hemicellulose and lignin respectively (Lee, 2005). The average distribution of cellulose, hemicellulose and lignin by mass for different SA varieties of sugarcane was reported to be 41.1%, 26.4% and 21.7 % respectively (Görgens, et al., 2015; Petersen, et al., 2014). The remaining

fractions of bagasse and trash also consist of small quantities of ash and extractives (4 % (w/w) and 6.8% (w/w) respectively) as reported by Petersen et al. (2014). Flow values provided in Table 1 are reported for a plant feed capacity of 113.5 tons/h of sugarcane lignocellulose (Petersen, et al., 2014).

Table 1 – Average Polysaccharide Distribution for Different Varieties of SA Bagasse and Trash (Görgens, et al., 2015; Petersen, et al., 2014)

Component	Bagasse*		Harvesting Residues*		Mixture (Feedstock)	
	Fraction % (w/w)	Flow (ton/h)	Fraction % (w/w)	Flow (ton/h)	Fraction % (w/w)	Flow (ton/h)
Cellulose	41.1	18.495	39.8	7.96	40.7	26.455
Hemicellulose	26.4	11.8	28.6	5.72	27.1	17.6
Lignin	21.7	9.765	22.5	4.5	21.9	14.265
Ash	4	1.8	2.4	0.48	3.5	2.28
Extractive	6.8	3.06	6.7	1.34	6.7	4.4
<b>Sum DM</b>	<b>100</b>	<b>45</b>	<b>100</b>	<b>20</b>	<b>100</b>	<b>65</b>
<b>Water</b>		<b>45</b>		<b>3.5</b>		<b>48.53</b>
<b>Total (Liq + Solid)</b>		<b>90</b>		<b>23.53</b>		<b>113.53</b>

\*Values based on an average for South African bagasse (Petersen, et al., 2014); DM = Dry Mass Basis

### 2.1.1 Cellulose

Cellulose is a polysaccharide consisting of glucose monomer units ( $C_6H_{10}O_5$ )<sub>n</sub> linked together via  $\beta$ -(1-4) - glycosidic bonds (Agbor, et al., 2011). It has a very high degree of polymerisation, consisting of long polymer chains of between 2000 to 27 000 cellobiose units (dimer of glucose monomers) (Heinonen, 2013), packed together in a crystalline form that takes the shape of long strands called micro fibrils (Bhayani & Ramarao, 2013; Carvalho, 2009). Micro-fibrils are further bonded together in a matrix of hemicellulose and lignin via hydrogen bonds and covalent bonds respectively, in an orderly manner to form larger macro-fibrils that make up the cell walls of plant cells in sugarcane (Carvalho, 2009) (refer to Figure 2).

Owing to celluloses' rigid crystalline structure, high degree of polymerization and hydrophobic properties, it is highly resistant to pre-treatment methods and insoluble in inorganic solvents (Carvalho, 2009). In addition, water layers get trapped at the cell wall interface, preventing the diffusion of enzymes and other degradation products (Bhayani & Ramarao, 2013; Wyman & Yang, 2009; Carvalho, 2009). Thus, pre-treatment is an essential process step exposing the tight crystalline structure of the cellulose, hemicellulose and lignin to further down-stream processing steps in a biochemical conversion consisting of enzymatic hydrolysis, possibly followed by fermentation.

### 2.1.2 Hemicellulose

Hemicellulose is a hetero-polysaccharide that consists of units of five and six carbon sugars and other acidic non-sugar groups (Mamman, et al., 2008; Sun, 2008). It encases the cellulose in a non-crystalline sheath form together with lignin (Bhayani & Ramarao, 2013). Hemicellulose polymers are branched and amorphous, which unlike cellulose, have a low degree of polymerisation in the range of between 80 to 200 units (Sun, 2008). Therefore, owing to hemicellulose's branched structure, it is more susceptible to pre-treatment attack and hydrolysis process steps than cellulose and lignin. Often pre-treatment steps will solubilise hemicellulose polymers into shorter chain oligomers and smaller monomer sugar units, which can be physically separated from an insoluble cellulose and lignin (cellulignin) fraction (Mamman, et al., 2008).

Hemicellulose monomer units have the general formula  $(C_5H_8O_4)_n$  for pentosans and  $(C_6H_{10}O_5)_n$  for hexosan units. These consist primarily of the following sugars and acidic monomers: D-xylose, L-arabinose (pentosans), D-glucose, D-galactose, D-mannose (hexosans), D-glucuronic acid, 4-O-methyl-D-glucuronic acid and D-galacturonic acid (acids) (Agbor, et al., 2011; Hendriks & Zeeman, 2009; Sun, 2008; Mamman, et al., 2008). For the purpose of this study, xylose contained within the hemicellulose polymer (xylan hemicellulose) and glucose contained within cellulose and hemicellulose (glucan hemicellulose), will specifically receive focus to be chemically converted to value-added products.

### 2.1.3 Lignin

Lignin is a cross-linked polymer consisting of numerous phenolic compounds (Bhayani & Ramarao, 2013). Lignin binds the cellulose crystals and hemicellulose sheath into a matrix that forms the structure of the plant cell walls in the bagasse and leaves (refer to Figure 2). Lignin is constructed in a three-dimensional randomised structure inside the cell walls from 3 major  $C_6-C_3$  monomers (as seen in Figure 3 below), through oxidative coupling (Fisher & Fong, 2014; Mamman, et al., 2008). These monomers include sinapyl alcohol, coniferyl alcohol and p-coumaryl alcohol monolignol oligomer units (Fisher & Fong, 2014; Mamman, et al., 2008). These are present within lignin in the form of phenylpropanoids as syringyl (S-unit), guaiacyl (G-unit) and p-hydroxyphenyl (H-unit) oligomer units (Boerjan, et al., 2003).

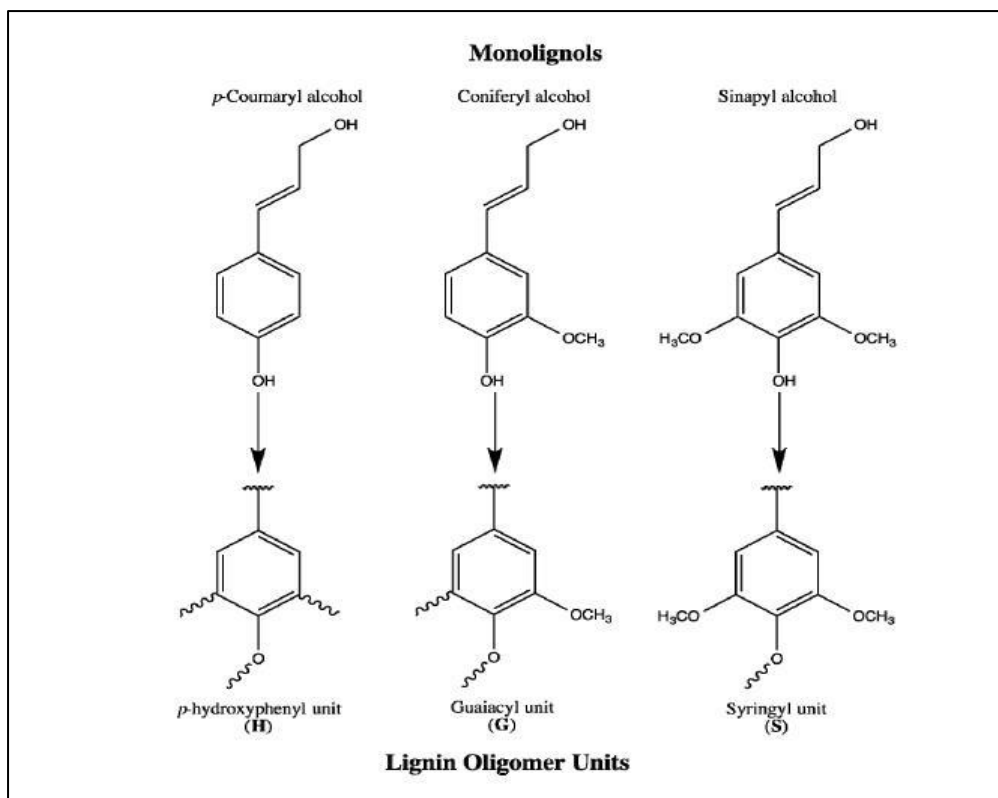


Figure 3 - Three Major Lignin Polymer Components (permission obtained from (Fisher & Fong, 2014) under Creative Commons Attribution License)

As research for lignin valorisation is ongoing and chemical component separations in lignin remain a large obstacle in terms of cost, biorefineries typically combust the insoluble lignin portion of the lignocellulose in boilers as fuel for energy (Görgens, et al., 2015). Lignin combustion is capable of producing heat and energy that power downstream production processes as well as existing annexed processes such as sugar mills. Heating values of 21.45 and 23.5 MJ/kg for lignin in have been provided for hardwoods and softwoods respectively (Blunk & Jenkins, 2000).

## 2.2 Pre-treatment Technologies

Pre-treatment aims to cleave inter-chain bonds in cellulose and hemicellulose to form commercially viable quantities of monomeric sugars (xylose, arabinose and glucose) for down-stream reactions to bioproducts (Harrison, et al., 2013; Lavarack, et al., 2002). In terms of glucose utilization to bioproducts, pre-treatment allows hydrolytic enzyme access to the crystalline portion of the cellulose chains and the formation of soluble glucose from amorphous cellulose (Jain, et al., 2016; Diedericks, et al., 2013). Pre-treatment can fall into one of four categories, namely, physical pre-treatment, chemical pre-treatment, physio-chemical pre-treatment and biological treatment (Harrison, et al., 2013; Carvalho, 2009).

## 2.2.1 Physical pre-treatment

### 2.2.1.1 Ultrasonic

Ultrasonic pre-treatment of lignocellulosic biomass involves subjecting the biomass to mechanical vibrations at an ultrasonic frequency (greater than 20 kHz) (Liyakathali, 2014). The ultrasonic waves are produced by a transducer made from a piezoelectric material that is supplied with an alternating current (Liyakathali, 2014; Ushakov, 2005). The action of the ultrasonic waves enhances other chemical and physical treatments by producing pressure differences in the solution medium (Mason & Lorimer, 2003). Two primary effects are produced by ultrasonic energy, namely, cavitation and heating (Yachmenev, et al., 2009). Cavitation occurs at lower frequencies of energy (16-100 kHz), whereas heating effects only occur from frequencies exceeding 100 kHz (Liyakathali, 2014). In terms of cavitation effects, high pressure compression and low pressure rarefaction produce expanding and contracting cavitation bubbles respectively. Cavitation bubbles transport additional molecules and catalyst into the process (Liyakathali, 2014). When the bubbles collapse near the substrate surface, they produce high localised pressures and temperatures that result in the formation of shearing forces around the bubble and consequently a strong mechanical action that disrupts the lignocellulosic structure, exposing it to enzymatic hydrolysis (Yachmenev, et al., 2009). The amount of cavitation bubbles produced, their lifetime and their implosion pressure are affected by the power of the ultrasound. Higher power, in terms of transducers translates to higher acceleration intensity and ultrasonic wave amplitude (Liyakathali, 2014). Research has reported an optimum temperature of 50°C at which maximum cavitation effects occur for enzymatic action (Yachmenev, et al., 2009).

### 2.2.1.2 Milling

Milling is the mechanical breakdown of lignocellulosic material (Canilha, et al., 2012). The objective of milling is to reduce the particle size and crystallinity of lignocellulosic feedstock by subjecting it to a chipping, serrating or grinding action (Canilha, et al., 2012; Mamman, et al., 2008). This increases the feedstock's surface area and reduces its degree of polymerization (Alvira, et al., 2010). Ball milling is one method that involves pulverising the biomass into finer particles using spheres in a rotating cylindrical drum. The particles produced by ball-milling are fine and small in size, with a large amount of cellulose crystallinity reduced during the process. This results in a feed more susceptible to enzymatic hydrolysis (Silva, et al., 2010). Wet disk milling is another method that uses non-porous disk grinders (typically made of ceramic materials) to decrease lignocellulosic feed particle size (Silva, et al., 2010). The material is ground between upper and lower rotating disks with a very low clearance. Clearance between disks range from 20-40 µm with a rotation speed of 1800 rpm (Silva, et al., 2010). This type of milling has gained acceptance as a potential attractive method for pre-treating rice straw in Japan, as a means to improve enzymatic digestibility (Hideno, et al., 2009). This milling technique is

seen as an attractive method in comparison to conventional ball milling and hot-compressed water pretreatment methods due to its lower energy consumption (Hideno, et al., 2009). Additional milling techniques include two-roll, hammer, colloid and vibro-energy milling to name a few, all of which can be used to enhance downstream pretreatment steps and enzymatic hydrolysis of cellulose (Alvira, et al., 2010).

## *2.2.2 Chemical pretreatment*

### *2.2.2.1 Acid Hydrolysis*

Acid hydrolysis can be categorised into two categories, based on the concentration of the acid applied, the severity of the reaction conditions and the end application thereof. Concentrated acid hydrolysis (concentrations between 30-70 wt. %) is applied to enable the depolymerisation of both the hemicellulose and cellulose fractions (Gírio, et al., 2010). Concentrated acid treatment is less favourable than dilute acid (DA) pre-treatment due to increased operational costs, as well as higher formation of degradation products, when exposed to raised temperatures (Alvira, et al., 2010; Hendriks & Zeeman, 2009). Additional disadvantages of concentrated acid treatment also include high acid recovery requirements and costly corrosion resistant equipment (Gírio, et al., 2010; Zheng, et al., 2009). The formation of inhibitors for both types of hydrolysis may require additional detoxification steps to improve the fermentability of hydrolysates (more so for concentrated acid hydrolysis) (Canilha, et al., 2012; Palmqvist & Hahn-Hagerdal, 2000).

DA pre-treatment is one of the most widely understood and researched pre-treatment methods (Chandel, et al., 2012; Agbor, et al., 2011; Kazi, et al., 2010). It uses low concentrations of acids at elevated temperatures to solubilize the hemicellulose fraction in the lignocellulosic feedstock to liberate pentose sugars and allow the cellulose portion to become more accessible to down-stream enzymatic hydrolysis (Canilha, et al., 2012; Alvira, et al., 2010; Hendriks & Zeeman, 2009; Palmqvist & Hahn-Hagerdal, 2000).

Treatment conditions for DA range from 140 – 215 °C for temperatures and a few minutes for residence times (12- 22 min) (Diedericks, et al., 2013; Benjamin, et al., 2013; Agbor, et al., 2011). Other sources provide shorter temperature ranges of 121-160 °C depending ultimately upon the feedstock being treated as well as the desired downstream products (Gírio, et al., 2010). The types of inorganic acids typically utilized during DA pre-treatment include hydrochloric, nitric, phosphoric and sulphuric acids. However, sulphuric acid is the most common choice of acid (Mosier, et al., 2005). Acid

concentrations vary between 0.5-1.5 % (w/w) for most aqueous solutions containing substrate (Agbor, et al., 2011; Gírio, et al., 2010).

During hydrolysis, water is used to cleave the inter-chain bonds in cellulose and hemicellulose. Bonds between the hemicellulose polymers are weaker than those for cellulose and consequently break down into oligomers as well as individual C5 and C6 monomer sugars and acids (mainly arabinose, xylose, mannose, glucose and side chain acetyl groups forming acetic acid) (Chandel, et al., 2012; Hendriks & Zeeman, 2009; Mosier, et al., 2005). Consequently, a soluble liquid phase containing hemicellulosic oligomers and monomers and a solid phase forms. The hydrolysate predominantly consists of xylose, whereas the solid phase primarily contains cellulose and lignin fractions (cellulignin). The formation of two distinct phases after DA pre-treatment allows the physical separation of the two to take place (Dussán, et al., 2014; Chandel, et al., 2012). Table 2 below provides a summary of the composition of hemicellulosic hydrolysate using DA for various agro-residues, including sugarcane leaves and bagasse (Chandel, et al., 2012).

Table 2 - Hemicellulose hydrolysate composition for different feedstock using dilute acid pre-treatment, adapted from (Chandel, et al., 2012)

Component	Rice straw	Wheat straw	Corn Stover	Sorghum straw	Sugarcane leaves	Sugarcane bagasse
<b>Sugar Concentrations (g/l)</b>						
Xylose	17.2	12.9	9.09	54.2		
Arabinose	3.3	2.71	1.01	12	30.29*	56.5*
Glucose	4.3	1.85	2.13	13.5		
<b>Inhibitors Concentrations (g/l)</b>						
Acetic acid	1.43	2.85	1.48	0	3.19	5.45
HMF	0.15	-	-	-	0.15	-
Furfural	0.25	0.16	0.56	0.2	0.56	1.89
Phenolic	-	-	0.08	-	-	2.75
<b>Process Conditions</b>	1.5 % H <sub>2</sub> SO <sub>4</sub> , 130 C, 30 min, S: L = 1:10	1.85 % (w/v) H <sub>2</sub> SO <sub>4</sub> , 90 C, 18 h, S: L = 1:20	2.13 % H <sub>2</sub> SO <sub>4</sub> , 180 min, 121 C, S: L = 1:10	2.13 % H <sub>2</sub> SO <sub>4</sub> , 180 min, 121 C, S: L = 1:10	130 C, 2.9 % H <sub>2</sub> SO <sub>4</sub> w/v, S:L = 1:4, 30 min	2.5 % (v/v) HCl, 140 C, 30 min, S: L = 1:10
	<b>Reference</b>	(Baek & Kwon, 2007)	(Canilha, et al., 2008)	(Cao, et al., 2009)	(Sepúlveda- Huerta E, 2006)	(Moutta, et al., 2011)

\* combined total sugars (xylose, arabinose and glucose)



From the table above, sugarcane bagasse and leaves are seen having a total sugar content of 56.5 and 30.29 g/l in the hydrolysate respectively, after having undergone DA pretreatment for the conditions listed above (Moutta, et al., 2011; Chandel, et al., 2007). Benjamin, et al., (2014) employed a combined severity factor (CSF,  $\log R_0'$ ) to relate the DA residence time (t, min), the acidity (in terms of pH) and the reaction temperature ( $T_H$ , C°) to the xylose and glucose recoveries in the liquid hydrolysate and solid cellulignin fractions of 4 different sugarcane cultivars after DA pretreatment. In addition, the CSF can be used to predict the amount of furfural produced during the pretreatment process, whereas other inhibitors present are not reported to be modelled (Benjamin, et al., 2014) (Refer to Equation 1).

$$\log R_0' = \log \left( t \cdot \exp \left[ \frac{T_H - 100}{14.75} \right] \right) - pH_{out} \quad [\text{Eq. 1}]$$

Diedericks, et al., (2013) determined a statistical relationship between the reaction time, reaction temperature, as well as the acid loading (mmol/l) to the xylose, glucose and total sugar yield (g monomers and oligomers/kg biomass) as well as the furfural production (g furfural/kg biomass). An optimum point for maximum xylose and minimum furfural production was listed as 182 g xylose/kg biomass and 6 g furfural/kg biomass for a reaction time of 8.6 min, a reaction temperature of 433.2K (160 °C) and an acid loading of 45 mmol H<sub>2</sub>SO<sub>4</sub>/l. Longer residence times and higher temperatures result in unfavourable furfural formation (Diedericks, et al., 2013).

#### 2.2.2.2 Alkaline pretreatment

Alkaline pretreatment is a delignification process. Depending on severity conditions, alkaline pretreatment solubilises lignin with significant portions of hemicellulose (under higher severity) (Chandel & Silva, 2013; Canilha, et al., 2012). It exposes the remaining cellulose and hemicellulose fractions in pre-treated solids to further enzymatic reactions (Agbor, et al., 2011; Carvalho, et al., 2008). The mechanism behind alkaline hydrolysis is the solvation and saponification of intermolecular ester bonds that crosslink xylan hemicelluloses and lignin (Chandel & Silva, 2013; Sarkar, et al., 2012; Sun & Cheng, 2002), causing the lignocellulosic structure to swell (Zheng, et al., 2009).

Alkaline pretreatment typically utilises bases such as potassium and sodium hydroxide (KOH & NaOH), lime (CaOH<sub>2</sub>), sodium carbonate (Na<sub>2</sub>CO<sub>3</sub>) (Carvalho, 2009), hydrazine, ammonia or ammonium hydroxide in combination with hydrogen peroxide (Chandel & Silva, 2013; Agbor, et al., 2011; Alvira, et al., 2010). Conditions utilized during alkaline pretreatment include, low alkali concentrations (less than 4 wt. %) and mild temperatures (25-60 °C) (Heinonen, 2013; Tutt, et al., 2012).

An advantage of alkaline pretreatment is that it utilises lower temperature and pressure ranges in comparison to other pretreatment methods and is therefore from an operational perspective, more economical (Zheng, et al., 2009). However, a disadvantage is that biomass requires exposure times typically in the order of hours or even days making the treatment slow in comparison to other pretreatment methods (Zheng, et al., 2009). In addition, alkali solutions are converted to irrecoverable salts that become incorporated into the lignocellulose structure during the pretreatment step, which makes them challenging to treat before undergoing subsequent fermentation and purification steps, increasing production costs (Zheng, et al., 2009).

Alkali pretreatment is seen as a suitable pretreatment technology for biomass feeds consisting of low lignin content such as agricultural residues rather than high lignin-containing biomass such as soft and hardwoods (Chandel & Silva, 2013; Agbor, et al., 2011; Zheng, et al., 2009).

#### *2.2.2.3 Wet oxidation*

Wet Oxidation (WO) is a pretreatment method that exposes lignocellulosic biomass to water and oxygen at temperatures greater than 120 °C during a two-step reaction. The first step is a low temperature hydrolytic reaction and the second step is a high temperature oxidative reaction (Martin & Thomsen, 2007; McGinnis, et al., 1983). In terms of conversions, Martin & Thomsen, (2007) found that 56.5% of cellulose contained within raw sugarcane bagasse, was converted to glucose after WO pretreatment at 195°C for 10 min under pressures ranging from 3-12 bars followed by enzymatic hydrolysis.

#### *2.2.2.4 Ozonolysis*

Ozonolysis is a pretreatment method that employs the strong oxidising agent ozone to selectively degrade lignin over carbohydrates in crop residue feedstock. This is achieved by cleaving olefinic and activated aromatic bonds (Panneerselvam, 2013). The ozone specifically oxidises and reacts with conjugated double bonds such as those most prevalent in lignin (Kumar, et al., 2009), thereby solubilizing the lignin inside the lignocellulose structure (Travaini, et al., 2016). Treatment under normal room temperatures and pressures is one of the main advantages of this pretreatment method (Mussatto & Teixeira, 2010). However, costs associated with this treatment are higher than conventional alkali and acid pretreatment due to the high quantities of ozone required (Kumar & Sharma, 2017; Bensah & Mensah, 2013).

#### *2.2.2.5 Solvent Extraction*

This pretreatment method uses a mixture of organic solvents (e.g. ethanol, methanol, acetone and ethylene to name a few) and water together with or without acid catalysts to hydrolyse the bonds between lignin and holocellulose (cellulose and hemicellulose) (Mussatto & Teixeira, 2010). This solubilises the lignin and partially solubilises the hemicellulose as well (Mussatto & Teixeira, 2010).

#### *2.2.3 Physio-chemical pretreatment*

##### *2.2.3.1 Steam Explosion (Auto hydrolysis)*

Steam explosion (SE) is one of the most widely studied and documented pretreatment methods and one of the few technologies that are applied on industrial scale for the solubilisation of hemicellulose (Canilha, et al., 2012; Agbor, et al., 2011; Alvira, et al., 2010; Hendriks & Zeeman, 2009). In comparison to other pretreatment methods, steam explosion is a favourable method due to its low environmental impact and capital cost to implement as result of not requiring the addition of chemicals (Avellar & Glasser, 1998). One of its notable disadvantages is the formation of large quantities of inhibitors after pretreatment as a result of severe temperatures and pressures (Hendriks & Zeeman, 2009). During this pretreatment, biomass is subjected to high pressure saturated steam for short time periods from a few seconds to minutes (Agbor, et al., 2011; Alvira, et al., 2010). The sudden depressurisation combined with high temperatures result in the physical exposure of the cellulose structure and the chemical hydrolysis of the hemicellulose for the formation oligomers and monomeric sugars (Canilha, et al., 2012).

Auto-hydrolysis is synonymous with steam explosion (Agbor, et al., 2011). Hydrolysis of acetyl groups on the hemicellulosic portion of biomass cause the formation of acetic acid that further catalyse the hydrolysis of hemicellulose to xylose and glucose monomers (Alvira, et al., 2010). In addition, at high temperatures water exhibits acidic properties, further catalysing the breakdown of the lignocellulose structure (Alvira, et al., 2010). Key factors affecting the performance SE include, operating temperature, particle size and exposure time. Typical conditions used during SE include temperatures and pressures ranging between 160 - 240 °C and 0.7 - 4.8 MPa respectively (Canilha, et al., 2012; Agbor, et al., 2011). Furthermore, steam explosion can be catalysed with the addition of acids, most commonly H<sub>2</sub>SO<sub>4</sub> or SO<sub>2</sub>, CO<sub>2</sub> gases (Mosier, et al., 2005).

##### *2.2.3.2 Carbon Dioxide Explosion*

Carbon dioxide explosion, a variation of steam explosion, increases cellulosic fibre surface area by exposing lignocellulose, after a rapid increase in pressure, to supercritical CO<sub>2</sub> gas, followed by a rapid decrease in pressure in an explosive manner. High pressures cause the CO<sub>2</sub> to dissolve within the water

to form carbonic acid, which further encourages hydrolysis (Larissa Canilha, 2012). This disrupts the micro-fibril crystalline structure of cellulose and increases exposure surface area of polymers so that effective hydrolysis reactions can convert cellulose and hemicellulose into sugars. The distinction between operating conditions for steam and carbon dioxide explosion variations is that CO<sub>2</sub> explosion exposes biomass to far lower temperature ranges that do not cause the degradation of xylose in hemicellulose to form inhibitory products (Zheng, et al., 1998). Temperature conditions are far lower at 31°C, whereas pressures are high, reaching as high as 73 atm (Harmsen, et al., 2010; Zheng, et al., 1998).

#### *2.2.3.3 Ammonia Fibre Explosion (AFEX)*

In a similar fashion to steam and carbon dioxide explosion pretreatment techniques, lignocellulosic material is treated with liquid ammonia at high temperatures (typically 90 °C) and high pressures for a short duration (30 min, 1-2 kg NH<sub>3</sub>/kg dry biomass) (Harmsen, et al., 2010). The pressure is thereafter, reduced rapidly. This disrupts the cell wall structure of the lignocellulose, reducing cellulose crystallinity. In addition the ammonia dissolves parts of the hemicellulose and lignin exposing the cellulose structure to downstream enzymatic hydrolysis (Harmsen, et al., 2010).

#### *2.2.4 Comparative Summary of Pretreatment Technologies*

The choice of pretreatment for the production of bioproducts ultimately depends on the production process and the fractions of lignocellulose that are to be utilized for further down-stream reactions. Table 3 compares the effectiveness of different pretreatment technologies in terms digestibility and solubilisation criteria (Mosier, et al., 2005).

Table 3 - Effect of different pretreatment technologies on the structure of lignocellulose adapted from (Mosier, et al., 2005); (Alvira, et al., 2010))

Pretreatment Criteria	Milling	Steam Explosion	LHW	Acid	Alkaline	Oxidative	AFEX	ARP	Lime	CO <sub>2</sub> Explosion
<b>Increases accessible surface area</b>	H	H	H	H	H	H	H	H	H	H
<b>Cellulose Decrystallisation</b>	H	-	n.d.	-	-	n.d.	H	H	n.d.	-
<b>Hemicellulose solubilisation</b>	-	H	H	H	L	-	M	M	M	H
<b>Lignin solubilisation</b>	-	M	L	M	M	M	H	H	H	-
<b>Generation of toxic inhibitors</b>	-	H	L	H	L	L	L	M	M	-
<b>Lignin structure alteration</b>	-	H	M	H	H	H	H	H	H	-

H: High effect, M: Medium effect, L: Low effect, n.d: not determined

Table 4 below provides a comparison between 5 different pretreatment methods, in terms of fixed capital investment and product formation (litres of product formed, with bioethanol being a widely studied biofuel), using a no-pretreatment criteria as a baseline (Eggeman & Elander, 2005). It can be seen that alkaline pretreatment with lime requires the lowest total fixed capital investment out of the five technologies evaluated, but has the second smallest product formation, whereas DA and AFEX require the highest capital investment but contribute to higher revenues in terms of bioethanol volume produced and therefore require lower total fixed capital per litre of annual capacity.

Table 4 - Capital cost of five pretreatment processes adapted from (Eggeman &amp; Elander, 2005)

Method	Pretreatment direct fixed capital, \$MM	Pretreatment breakdown % reactor/other	Total fixed capital, \$MM	Ethanol production, MI/year	Total fixed capital, \$/l annual capacity
DA	25	64/36	209	212	14.1
Hot water	4.5	100/0	201	167	17.3
AFEX	25.7	26/74	212	215	14.1
ARP	28.3	25/75	211	175	17.3
Lime	22.3	19/18	164	185	12.7
No pretreatment	0	N/A	200	34	84.3

Table 5 provides a summary of sugar and bioethanol conversions and yields respectively, for different pretreatment technologies utilising switch grass as the lignocellulosic feed (Tao, et al., 2011). SO<sub>2</sub> catalysed steam explosion, DA and AFEX show the highest total monomer sugar yield as well as ethanol production in comparison to other pretreatment methods listed.

Table 5 - Summary of conversion data for six different pretreatment technologies using switch grass as the lignocellulosic feed adapted from (Tao, et al., 2011)

Method	Monomer sugar yield (%)	Glucan → glucose yield (%)	Xylan → xylose yield (%)	Ethanol production (MM gal/year)	Ethanol yield (gal/dry US ton feedstock)	% Theoretical ethanol yield
AFEX	76	76	76.1	50.8	65.8	58
DA	76	74.5	78.4	49.9	64.6	56.9
Lime	70	80.9	52.7	46.8	60.5	53.3
LHW	61.4	78.8	33	40.1	52	45.8
SAA	52	65.5	30.8	34.3	44.4	39.1
SO <sub>2</sub>	79.2	80.5	77.1	52.2	67.7	59.6

### 2.3 Enzymatic Hydrolysis

Enzymatic hydrolysis usually follows after preceding pretreatment methods (e.g. steam explosion or DA). During this process, enzymes such as cellulases are utilized to catalyse the break-down of the solid recalcitrant cellulose stream to liberate glucose monomers for further down-stream fermentation. Cellulases can be produced from genetically engineered microorganisms such as *Trichoderma reesei*, which can be produced on-site at bio-refineries, such as those used for bioethanol production or bought off-site (Humbird, et al., 2011). Cellulases can further be divided into 3 main types (Quiroz-Castañeda & Folch-Mallol, 2013; Harrison, et al., 2013; Mussatto & Teixeira, 2010)) as seen in Figure 4:

- Endo-glucanases which cleave the internal section of cellulose polymer chains
- Exo-glucanases which cleave cellobiose from the end of cellulose chains
- β-glucosidases which cleave the dimer cellobiose into glucose

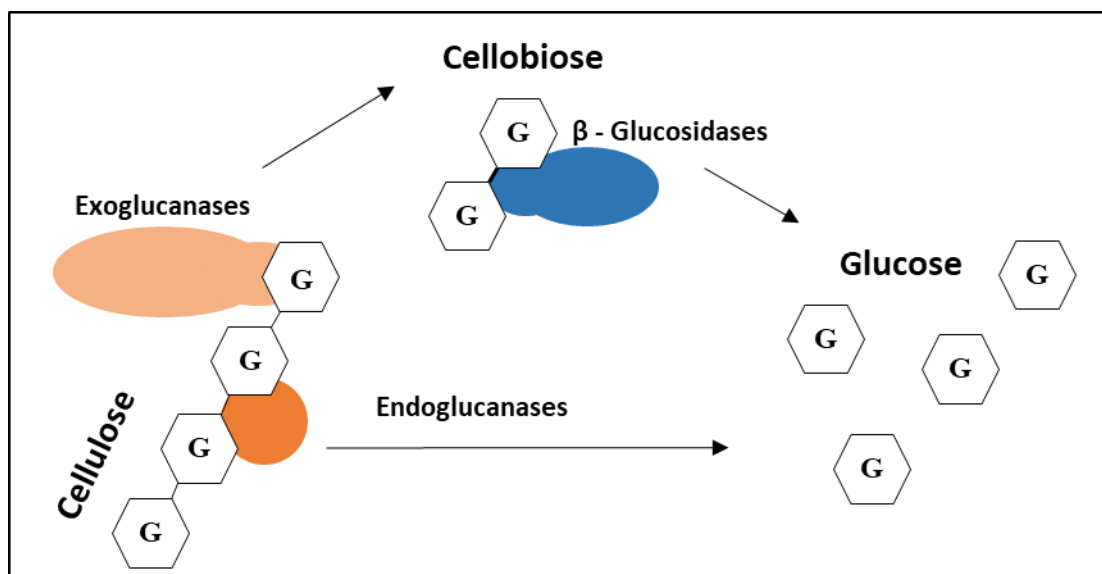


Figure 4- Hydrolysis of cellulose onto cellulase active site into monomeric glucose redrawn from (Quiroz-Castañeda & Folch-Mallol, 2013; Mussatto & Teixeira, 2010)

During enzymatic hydrolysis, cellulose attaches to an amino acid active site on the cellulase, where the long chain cellulose polymer is cleaved by water into smaller oligomers, mainly cellobiose dimers (via active sites on exo-glucanases) and further into glucose monomers via  $\beta$ -Glucosidases from (Quiroz-Castañeda & Folch-Mallol, 2013; Mussatto & Teixeira, 2010). Cellulase loading for bioethanol production has previously been reported to be 20 mg enzyme per gram cellulose (Humbird, et al., 2011; Roche, et al., 2009). Sugars liberated via enzymatic hydrolysis are then ready for fermentative or chemical conversion into useful bioproducts.

Advantages over concentrated acid hydrolysis is that the process is more selective, producing minimal by-products. Enzymatic hydrolysis is also less energy intensive and provides higher sugar yield (Bensah & Mensah, 2013). In addition, little to no severe chemicals are required. A disadvantage is that enzymatic hydrolysis is a slower and more expensive method in comparison to concentrated acid hydrolysis (Heinonen, 2013). Table 6 provides enzymatic hydrolysis reactions with their respective conversions (Humbird, et al., 2011).

Table 6 - Enzymatic Hydrolysis Reactions and Conversions for Cellulosic material adapted from (Humbird, et al., 2011)

Reaction	Reactant	% Conversion
$(Glucan)_n \rightarrow n \text{ Glucose Oligomer}$	Glucan	4
$(Glucan)_n + \frac{1}{2}n H_2O \rightarrow \frac{1}{2}n \text{ Cellobiose}$	Glucan	1.2
$(Glucan)_n + n H_2O \rightarrow n \text{ Glucose}$	Glucan	90
$Cellobiose + H_2O \rightarrow 2 \text{ Glucose}$	Cellobiose	100

## 2.4 Purification and Detoxification Technologies

The use of purification or detoxification technologies is required to remove three main groups of inhibiting compounds from hydrolysates that can potentially inhibit reactions during fermentation or catalytic conversion stages. These are namely, furans (5-HMF and furfural), aliphatic acids (levulinic and formic acid), acetic acid and phenolic compounds produced during the hydrolysis of the lignocellulosic structure, during pretreatment stages (Bao, et al., 2014; Jönsson, et al., 2013; Chandel & Silva, 2013; Palmqvist & Hahn-Hagerdal, 2000; Larsson, et al., 1999). Figure 5 shows the different types of inhibitors formed during pretreatment stages from their respective sugars and organic molecules under severe temperature and acid concentration (Jönsson, et al., 2013).

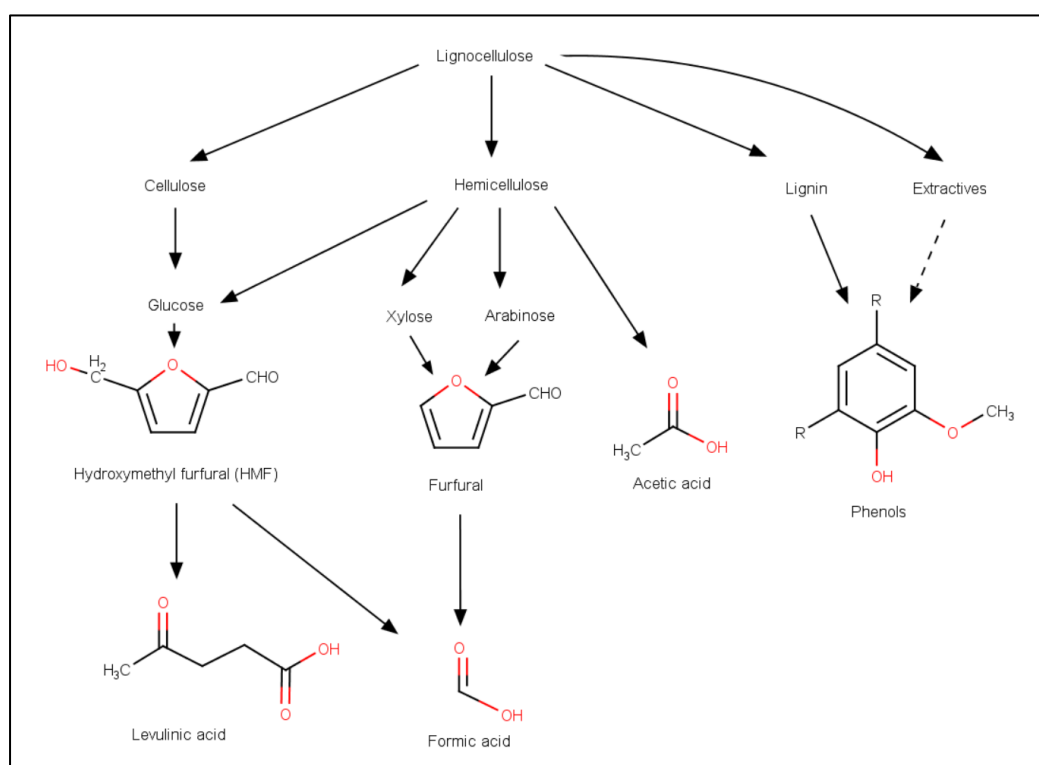


Figure 5 - Different Types of Inhibitors Formed during Pretreatment (redrawn with permission from (Jönsson, et al., 2013))

The objectives of detoxification are to remove as much inhibiting components from the liquid hydrolysate as possible while simultaneously minimising xylose and glucose sugar losses and operating costs. A few detoxification technologies that are utilized in biorefineries are covered within the subsequent sections below.

### 2.4.1 Activated carbon adsorption

Activated carbon adsorption is widely used as a detoxification and purification method due to its high adsorption capacity, low cost and simple operation (Liu, et al., 2014; Chandel & Silva, 2013). It is typically used to remove non-saccharide components from hemicellulosic hydrolysate such as furans,



acids and phenolic components (Zhaojiang Wang, 2016; Palmqvist & Hahn-Hagerdal, 2000). Important variables that influence the efficacy of the method, include the pH of the hydrolysate feed, the adsorption exposure time and the concentration of activated carbon applied to the hydrolysate. Variations in activated charcoal concentrations and exposure times on inhibitor removal (furans and phenolic components) from sago trunk hydrolysate were investigated by Kamal et al. (2011) for xylitol production. Optimal conditions were established for 60 min of charcoal adsorption exposure time and 2.5 % (w/v) charcoal dosing, which resulted in the highest reduction of furfural and phenolic components of 53% and 78% respectively, compromising a 7% total loss in xylose and glucose sugars (Kamal, et al., 2011). *Eucalyptus* hydrolysate at a pH of 5.5 exposed to 5% (w/v) of activated charcoal for 60 min exposure time, was found to be the only condition that minimised sugar losses (7.7 % total sugar loss) and simultaneously allowed 100% removal of HMF and furfural (Villarreal, et al., 2006). In addition, a 79.1% reduction in phenolic compounds were reported for the same conditions. Activated carbon adsorption (10 % w/v, 1h) under a pH condition of 5.5, was found to remove 92 % of furfural, 68 % of 5-HMF, 17 % acetic acid, 11% formic and all levulinic acid from beer spent grain (BSG) hydrolysate respectively (Carvalho, et al., 2005). In addition, a 58 % reduction in phenolic components by (Carvalho, et al., 2005) was also reported.

#### 2.4.2 Evaporation

Evaporation is used as a physical detoxification method to increase monomer sugar concentrations by subjecting solutions to low vacuum pressures and mild temperatures (Canilha, et al., 2012; Mala & Anish, 2008). Evaporation can be used as a detoxification method for volatile inhibitor components in the hydrolysate that may be toxic for microorganisms involved during the fermentation step such as acetic acid, furfural and vanillin (Chandel & Silva, 2013). Consequently, it also concentrates toxic components that are non-volatile which is one of the main drawbacks of this technology. Larsson, et al., (1999) reported a furfural, 5-HMF, formic acid and acetic acid reduction of 100%, 4%, 65 and 74 % respectively, when spruce hemicellulosic hydrolysate was subjected to vacuum evaporation conditions resulting in a 90 % reduction in initial hydrolysate volume. Vacuum evaporation applied to *Eucalyptus* hemicellulosic hydrolysate at 70 °C, has been demonstrated to remove 97.7% of furfural (Villarreal, et al., 2006).

#### 2.4.3 Ion-exchange resins

The use of ion-exchange resins are a widely studied detoxification method for hydrolysates derived from a variety of feedstock (Chandel, et al., 2007; Villarreal, et al., 2006; Carvalho, et al., 2005; Larsson, et al., 1999). This method utilizes polymer resins with charged functional groups to remove

certain ionized toxic components from hydrolysates derived from the hydrolysis of lignocellulose (Canilha, et al., 2012). A negatively charged resin is used for cationic exchange, whereas positively charged resins are used during anionic exchange.

For large scale separations of products and starting sugars, columns packed with resins such as divinyl benzenes cross-linked with sulfonated polystyrene in alkali earth metal form, have been reported to successfully purify xylitol rich fractions originating from fermentation of xylose (Heikkila, et al., 1992). Notable advantages of ion-exchange technology include resin regeneration, once saturated, for reuse in the system and the ability to remove phenolic components, as well as furans and acidic groups (Canilha, et al., 2012). Disadvantages include the high loss of sugars and long processing times (Canilha, et al. 2012; Chandel, et al. 2011; 2007).

Carvalho, et al. (2005), investigated the use of various detoxification methods, including weak anionic and strong cationic exchangers at varying pH levels to reduce inhibitor concentration in BSG hydrolysate for the production xylitol. It was reported that during anion exchange at pH conditions at both 5.5 and 0.77, phenolic components in the hydrolysate were reduced by 83 % (Carvalho, et al., 2005). Both detoxification conditions also resulted in 100% furfural and levulinic acid removal. However total monosaccharides in the hydrolysate were reduced by 3 % in 5.5 pH conditions as opposed to 5% monosaccharide for 0.7 pH conditions (Carvalho, et al., 2005). Sequential treatment of sugarcane bagasse hydrolysate with anion-exchange followed by cation exchange resins was also previously reported to reduce acetic acid concentration by more than 40 % (Watson, et al., 1984). Other research highlighted an 84 % removal of acetic acid, employing anion exchange resins, which improved ethanol fermentation product yield from 0.27 to 0.37 (g/g) (van Zyl, et al., 1991). Inhibitors present in a *Eucalyptus* hydrolysate were reduced in a 4 resin system ion exchange column, operating at a pH of 1.8 and a down flow of 10 ml hydrolysate/min with alternating cation and anion stages, resulting in a maximum reduction in phenolic components of 93.4 % (Villarreal, et al., 2006).

#### 2.4.4 Centrifugation

Centrifugation is a physical separation technique used to amplify the force of gravity in order to separate solid and liquid or 2 liquid phases of differing densities from one another (Todaro, 1996). Centrifuges consist of a rotor that rotates and holds the product, a solids discharge, a drive system (shaft, bearings and motor), support frame and rotor enclosure (Todaro, 1996). In terms of solid-liquid phase separations, performance is dependent on a number of factors such as particle size distribution, concentration and shape. Although a wide variety of centrifuges exist, (sedimentation centrifuges,

tubular bowl, continuous decanter centrifuges, disk centrifuges, filtering centrifuges, vertical basket) all centrifuges operate on the basic principle that a mass spinning around a central axis at a fixed distance experiences a force equivalent to the mass of the object multiplied by its directional acceleration (Todaro, 1996). Centrifugal and centripetal forces keep the spinning object on its orbit around the axis of rotation.

(Linzmeyera, et al., 2014), compared centrifugation, microfiltration and vacuum filtration separation methods for solids removal in ethanol fermentation broth from banana culture waste. It was found that centrifugation yielded the lowest removal of solids as expressed by the high total solids present in the liquid fraction (Linzmeyera, et al., 2014).

#### *2.4.5 Membrane Extraction Filtration*

Membrane adsorption is used to separate the hydrolysate aqueous phase from the organic toxic phase containing the inhibitor components (Grzenia, et al., 2012; Canilha, et al., 2012). Membrane internal pores contain attached functional groups that allow the removal of toxic components (Grzenia, et al., 2012). An 80% reduction in acetic acid from corn stover hydrolysate, extracted using Alamine<sup>®</sup> 336 as the organic phase, was achieved in non-dispersive hollow fibre membrane extraction (Grzenia, et al., 2008).

#### *2.4.6 Neutralisation*

After lignocellulose has undergone pretreatment, hemicellulosic hydrolysate can undergo neutralisation in order to match pH levels during fermentation (Canilha, et al., 2012; Chandel, et al., 2011). Chemicals typically utilized during treatment include calcium hydroxide (over-liming) and sodium hydroxide as well as other mineral bases (Chandel, et al., 2011). During neutralization inhibitors, phenolic components and furfural are removed by precipitation (Chandel, et al., 2011). The addition of lime generates large quantities of calcium sulphate precipitate ( $\text{CaSO}_4$ ) that need to be removed from the system before subsequent fermentation, requiring additional separation steps that add cost to detoxification (Palmqvist & Hahn-Hagerdal, 2000).

#### *2.4.7 Summary of Detoxification Technology Applications*

Table 7 below provides a summary of the mechanisms, advantages and disadvantages of different detoxification technologies applied to hemicellulosic hydrolysate containing inhibitor compounds (Canilha, et al., 2012).

Table 7 - Advantages and disadvantages of different detoxification methods of hemicellulosic hydrolysate adapted from (Canilha, et al., 2012)

Detoxification	Process/Mechanism	Advantages	Disadvantages
<b>Evaporation</b>	Removes toxic compounds by evaporation in a vacuum concentrator based on the volatility	<ul style="list-style-type: none"> <li>• Reduces volatile compounds as acetic acid, furfural, and vanillin</li> </ul>	<ul style="list-style-type: none"> <li>• Increases the non-volatile toxic compounds as extractives</li> </ul>
<b>Membrane</b>	Membranes have surface functional groups attached to their internal pores, which may eliminate metabolic inhibitors	<ul style="list-style-type: none"> <li>• Avoids the need to disperse one phase</li> <li>• Minimize the entrainment of small amounts of organic phase</li> </ul>	<ul style="list-style-type: none"> <li>• High cost</li> <li>• selective removal of inhibitors</li> </ul>
<b>Ion Exchange</b>	Resins change undesirable ions of the liquid phase to be purified by saturating of functional groups of resins	<ul style="list-style-type: none"> <li>• Can be regenerated and reused</li> <li>• Does not cause high sugars loss</li> <li>• Remove lignin-derived inhibitors</li> </ul>	<ul style="list-style-type: none"> <li>• High pressure</li> <li>• Difficult to scale-up</li> <li>• Possible degradation of fragile biological product molecules</li> <li>• Long processing time</li> </ul>
<b>Over liming</b>	Increase of the pH followed by reduction	<ul style="list-style-type: none"> <li>• Precipitate toxic compounds</li> </ul>	<ul style="list-style-type: none"> <li>• High sugars loss</li> <li>• Filtration complexity</li> </ul>
<b>Activated Charcoal</b>	Adsorption of toxic compounds by charcoal which is activated to increase the contact surface	<ul style="list-style-type: none"> <li>• Low cost</li> <li>• Minimizes loss of sugars</li> <li>• Removes phenolic and furans</li> </ul>	<ul style="list-style-type: none"> <li>• Filtration complexity</li> </ul>
<b>Extraction with Organic Solvents</b>	Mix of liquid phase to be purified and an organic solvent. The liquid phase is recovered by separation of two phases (organic and aqueous)	<ul style="list-style-type: none"> <li>• Recycling of solvents for consequent cycles</li> <li>• Removes acetic acid, furfural, vanillin, 4-hydroxybenzoic acid and low molecular weight phenolic</li> </ul>	<ul style="list-style-type: none"> <li>• Long processing time</li> </ul>

## 2.5 Products Valorised from Sugarcane Lignocellulose

The following section highlights the chemical structure, applications, historical discovery and current industrial production methods for xylitol, citric acid and glutamic acid.

### 2.5.1 Xylitol

#### 2.5.1.1 Chemical Structure

Xylitol (pentahydroxypentane), is a 5-carbon sugar alcohol with sweetness comparable to sucrose and is found naturally in a number of edible plants and mushrooms (Mohamad, et al., 2015; Peterson, 2013; Rafiqul & Sakinah, 2013). It is also an important intermediary in the gluconate xylulose cycle for carbohydrate metabolism in humans (Makinen, 2016). It is derived from xylose through the reduction of the carbonyl group (Chandel & Silva, 2013). It has the chemical formula  $\text{CH}_2\text{OH}(\text{CHOH})_3\text{CH}_2\text{OH}$  and is structurally represented in Figure 6 with five carbon atoms in the main chain and five hydroxyl groups.

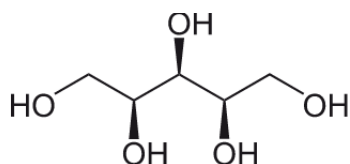


Figure 6 - Xylitol structural formula

#### 2.5.1.2 Industrial and Household Applications

Xylitol has a wide variety of applications in the food, dental and pharmaceutical industry due to its numerous beneficial properties. Xylitol has a high endothermic heat of solution, which helps induce a cooling effect in the palate and has sweetness comparable to sucrose with a lower calorie content. Therefore, xylitol aids as a healthier sweetening alternative in confectionary treats, soft drinks and chewing gum as well as in the dental and pharmaceutical industry in the place of other sugars (sucrose, glucose and fructose), which are harmful to individuals such as diabetics who are sensitive to differences in blood sugar levels, due to a lack of insulin production (Grumezescu & Holban, 2017; Kannan & Agastian, 2016; Rafiqul & Sakinah, 2013; Villarreal, et al., 2006). Another unique characteristic of xylitol compared to conventional sugars, is that it does not contain ketone or aldehyde groups that would otherwise darken food products when reacting with protein during sugar caramelization (Grumezescu & Holban, 2017). It is therefore used in products where the visual colour is required to remain unchanged. In recent work, xylitol has been discussed to have potential application within the polymer industry by functioning as a plasticizer of thermoplastic starch. This has been shown to provide a number of physical enhancements to the thermoplastic starch such as improved tensile properties, water sensitivity and reduced embrittlement (Muscat, et al., 2012). In the explosives industry, xylitol (including other sugar alcohols) can be nitrated through the addition of a combined cooled mixture of acetic anhydride and fuming nitric acid to form xylitol pentanitrate (Ostrinskaya, et al., 2016; Wright & Hayward, 1960). Xylitol may also potentially be used as a building block chemical for xylaric acid, propylene glycol, ethylene glycol and glycerol (Werpy, et al., 2004).

#### 2.5.1.3 Historical Discovery and Production

Xylitol was discovered in the 1890s by German researchers Fischer and Stahe in Germany and independently by French researcher Bertrand (2010). Finnish engineers and chemists achieved the first successful crystallisation of unstable xylitol from xylose reduction during World War II. The war had created a huge raw sugar material shortage in the country, resulting in the need to develop a sugar alternative. Thus the first industrial xylitol production facilities were established by the former Finnish Sugar Company Ltd. After the Second World War, the sugar shortage ceased and research into xylitol production stagnated for a while (Makinen, 2016). It was not until 1972 that xylitol once again

began receiving attention. The famous Turku Sugar Studies began in this year (2010). The studies leading to 1975 were concluded, with the outcome showing the positive health benefits of xylitol on the human body, namely its inhibitory effect on dental caries (Cooke, 2013). Within the same year, the first largescale xylitol production facility was started in Kotka, Finland with birch chips used as feedstock (Makinen, 2016; Cooke, 2013).

#### 2.5.1.4 Catalytic Hydrogenation of Xylose to Xylitol

Xylitol is currently produced on an industrial scale through the catalytic hydrogenation of xylose derived mainly from wood hydrolysates under high pressures using metal catalysts such as the Raney nickel catalysts (Mikkola, et al., 2003; Melaja & Hamalainen, 1977) (refer to Figure 7 for chemical reaction steps).

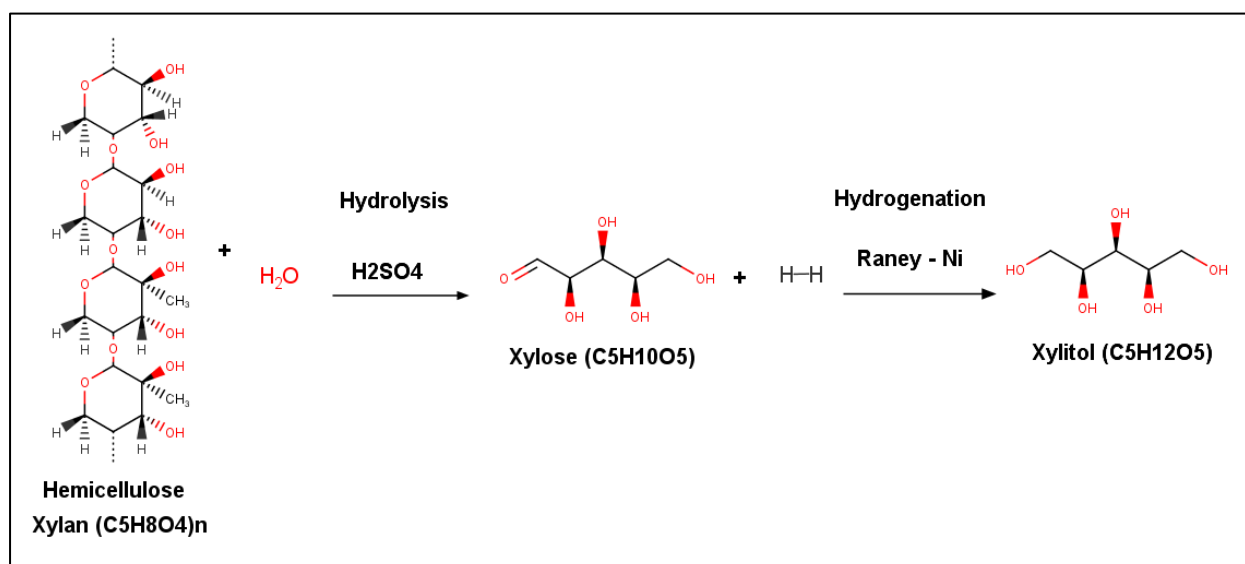


Figure 7 - Hydrolysis and Hydrogenation Steps in Xylitol Production, redrawn from (Tomishige, et al., 2016)

The main by-products produced during hydrogenation with Raney nickel catalysts include D-arabinitol (arabitol), D-xylulose, furfural and D-xylonic acid (Mikkola, et al., 1999). Their formation is primarily influenced by the reaction temperature, pH of the reaction media and hydrogen mass transfer (Yadav, et al., 2011; Mikkola, et al., 2003). Main and side reactions are provided in Table 8.

Table 8 - Main and side reactions for the hydrogenation of xylose adapted from ( (Mikkola, et al., 2003); (Yadav, et al., 2011))

Reaction	Reactant	Product
$C_5H_{10}O_5 + H_2 \rightarrow C_5H_{12}O_5$	Xylose	Xylitol
$Xylose (C_5H_{10}O_5) \rightarrow xylulose (C_5H_{10}O_5)$	Xylose	D-Xylulose
$C_5H_{10}O_5 - 3H_2O \rightarrow C_5H_4O_2$	Xylose	Furfural
$C_5H_{10}O_5 + H_2 \rightarrow C_5H_{12}O_5$	D-Xylulose	D-Arabinitol
$2C_5H_{10}O_5 + O_2 \rightarrow 2C_5H_{10}O_6$	Xylose	D-Xyloic acid

A variety of other metallic catalysts employed in current research include platinum supported on carbon, ruthenium supported on carbon, TiO<sub>2</sub>, SiO<sub>2</sub> or alumina modified with other metallic groups such as NiO or rhenium (Tomishige, et al., 2016; Yamaguchi, et al., 2016; Yadav, et al., 2011; Baudel, et al., 2005). More recently, the use of an iridium catalyst modified with rhenium supported on SiO<sub>2</sub> has been reported to be capable of simultaneously hydrolysing and hydrogenating hemicellulose to xylitol, pentanols and n-pentane derived from xylitol (Tomishige, et al., 2016). The main advantages of this new research is the ability to reduce detoxification and purification process steps, eliminating significant capital costs for technological implementation (Tomishige, et al., 2016). A disadvantage of this system is the higher expense of the catalyst due to the presence of iridium.

The raw feedstock studied in this context is lignocellulosic biomass and therefore initially requires pretreatment to release the xylose from the hemicellulosic portion of bagasse and trash, which is subsequently converted to xylitol. DA pre-treatment is a commonly applied method for the hydrolysis of the hemicellulose content in hardwoods or corncobs for the production of xylitol (Rafiqul & Sakinah, 2013; Baudel, et al., 2005; Melaja & Hamalainen, 1977). Various patents report using a xylose-rich solution, obtained through the hydrolysis of birch wood chips, with a solids content ranging between 25 and 50 wt. %, that consists of a xylose feed composition ranging from 75 to 85 wt.% (on a dry basis), after undergoing upstream ion exchange and activated carbon process steps (Melaja & Hamalainen, 1977). Whereas, other sources report feeding 50% (w/w) xylose in water for hydrogenation when using a Raney nickel catalyst (Mikkola, et al., 2003).

#### 2.5.1.5 Current Industrial Production Methods

Production of xylitol can be achieved via two main routes namely chemical production or fermentative production. Fermentative production involves the use of microorganisms to convert xylose to xylitol.

Yeasts such as *Candida tropicalis* and *Candida guilliermondii* are typically used (Rafiqul & Sakinah, 2013). This study will not be dealing with fermentative production and will instead pursue a simulation and economic analysis of a chemical synthesis route. Therefore, an in-depth discussion of fermentative technologies for xylitol production will not be presented. Chemical production is achieved through the hydrogenation of xylose in the presence of metal catalysts such as the Raney-nickel catalysts as well as Ruthenium metal catalysts supported on numerous materials as mentioned in previous studies by (Mikkola, et al., 1999) (Refer to Figure 7 above).

Chemical synthesis holds a few notable advantages over fermentative methods, namely high xylose conversions and low by-product selectivity, excluding the need for sterilization and additional cell-product separation and purification steps and is thus the method of focus for this study (Yadav, et al., 2011; Heikkila, et al., 1992; Melaja & Hamalainen, 1977).

## 2.5.2 Citric Acid

### 2.5.2.1 Chemical Structure

Citric acid, also chemically known as 2-hydroxy-propane-1,2,3-tricarboxylic acid (Papagianni, 2007) is a weak, non-toxic tri-carboxylic acid naturally found in citrus fruits and has the chemical formula in monohydrate form as  $C_6H_8O_7 \cdot H_2O$  (Angumeenal & Venkappayya, 2013). Its structural formula can be seen in Figure 8 below.

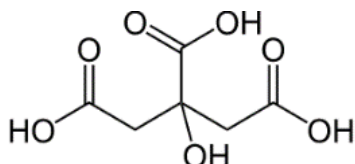


Figure 8 - Citric Acid Anhydrous Structural Formula

### 2.5.2.2 Industrial and Household Applications

Citric acid finds a number of applications within the food & beverage industry (70% global use) and as well as in the pharmaceutical and cosmetics industries (12 % global use) (Vandenberghe, et al., 1999). The list below provides a summary of its many applications:

- Natural preservative and flavour enhancer in food and cold drinks (Angumeenal & Venkappayya, 2013; Sian, 2013; Heinzle, et al., 2006)
- Emulsifier in ice creams and cheese (Vandenberghe, et al., 1999)
- Natural antioxidant, anticoagulant and effervescent in medication (Sian, 2013; Angumeenal & Venkappayya, 2013; Vandenberghe, et al., 1999)



- Buffering agent and replacement of polyphosphates in detergents (Angumeenal & Venkappayya, 2013; Soccol, et al., 2006; Petrides, 2000; Vandenberghe, et al., 1999)
- Descaling agent for membranes used in wastewater treatment (Ciriminna, et al., 2017)
- Small quantities also find applications in dyeing (Angumeenal & Venkappayya, 2013)
- Recently CA has been used as a cross-linking agent in the production of hard thermoset plastics that adhere to many materials, are non-toxic and fully bio-degradable. This is achieved through a Fischer-esterification process by dissolving CA in glycerol under ambient pressures below 130°C (Alberts & Rothenberg, 2012)
- Extraction agent when dissolved in water for the removal of pectin from apple pomace and flour (Canteri-Schemin, et al., 2005)

#### 2.5.2.3 Historical Discovery and Production

In 1784, the Swedish chemist Carl Wilhelm Scheele discovered citric acid (Papagianni, 2007). Lemon juice was treated with calcium hydroxide to form the salt calcium citrate and thereafter citric acid was isolated through the reaction with sulphuric acid ( $H_2SO_4$ ) (Kristiansen, et al., 2002; Vandenberghe, et al., 1999). Commercial production of citric acid began later, in 1826, in England using lemon juice (content of 7-9 wt. %) imported from Italy (Max, et al., 2010; Vandenberghe, et al., 1999). Italian citrus growers held a monopoly over the supply of citric acid until the early part of World War I (Vandenberghe, et al., 1999). In 1919, industrial production methods changed following the work conducted by (Currie, 1917) who found that certain strains of *Aspergillus niger* (*A. niger*) were capable of excreting large amounts of citric acid while growing in salt and sugar based mediums (Vandenberghe, et al., 1999). This work established the use of microbial fermentation for the production of citric acid for subsequent years (surface fermentation) (Max, et al., 2010; Kristiansen, et al., 2002). In the 1950's, the glycolytic pathway and tricarboxylic acid cycle (TCA) mechanism in microorganisms for citric acid production was discovered. This laid the foundation for the development of a new improved submerged microbial fermentation processes currently in use today (Vandenberghe, et al., 1999).

#### 2.5.2.4 Glucose Fermentation to Citric Acid

Citric acid (CA)-producing microorganisms (typically filamentous *A. niger* or yeasts) utilise molasses, sucrose or glucose liberated from starch or lignocellulosic sources as substrate (Ciriminna, et al., 2017; Heinzle, et al., 2006). When utilising second generation feedstock such as lignocellulose for CA production, biomass first undergoes pre-treatment to liberate glucose from hemicellulose (glucan) as well as a small portion of the cellulose. The remaining glucose left over from hemicellulose and the majority in cellulose is freed using enzymatic hydrolysis. Stoichiometry for enzymatic hydrolysis and

fermentation reactions (reactions 1 - 4) for CA production are provided in Table 9 (Khosravi-Darani & Zoghi, 2008; Heinzle, et al., 2006).

Table 9 - Main stoichiometric reactions for citric acid production via fermentation adapted from (Heinzle, et al., 2006)

No.	Reaction	Description	Yield % (w/w)	Reference
1	$(C_6H_{10}O_5)_5 H_2O + 4 H_2O \rightarrow 5C_6H_{12}O_6$	Enzymatic hydrolysis of glucan	92	(Heinzle, et al., 2006)
2	$C_6H_{12}O_6 + 0.28NH_4NO_3 + 0.012KH_2PO_4 \rightarrow 6CH_{1.72}O_{0.55}N_{0.09}P_{0.002} + 1.412H_2O + 1.088O_2 + 0.012K^+$	Biomass growth* with $NH_4NO_3$ as nitrogen source	Not specified	(Heinzle, et al., 2006)
3	$C_6H_{12}O_6 + 1.5 O_2 \rightarrow C_6H_8O_7 \cdot H_2O + H_2O$	Fermentation	97	(Khosravi-Darani & Zoghi, 2008)
4	$C_6H_{12}O_6 + 6 O_2 \rightarrow 6 CO_2 + 6H_2O$	Fermentation	Not specified	(Khosravi-Darani & Zoghi, 2008)

\* (Assuming *A. niger* biomass composition from (Heinzle, et al., 2006) as  $CH_{1.72}O_{0.55}N_{0.09}P_{0.002}$ )

During aerobic fermentation, glucose is metabolised by the microorganism and converted into CA and carbon dioxide (reactions 3 - 4) through a series of reactions within the tricarboxylic acid cycle (TCA) (Figure 9). A portion of the carbon source (glucose) along with a nitrogen source ( $NH_4NO_3$  as is the case with *A. niger*) and other micro nutrient salts are incorporated in the cell for growth (reaction 2) (Heinzle, et al., 2006). In a simplified manner, aerobic catabolism of glucose, in terms of CA production, can be summarised into 3 mechanisms (refer to Figure 9), namely acetyl coenzyme A (acetyl CoA) production, during which glucose is fermented to pyruvate through glycolysis and thereafter converted to the acetyl CoA intermediate by action of *pyruvate dehydrogenase complex*, while releasing  $CO_2$  (Shuler & Kargi, 2002). The second mechanism, acetyl CoA oxidation, is where acetyl CoA enters the TCA cycle and catalytically undergoes condensation with oxaloacetate to form citrate in the presence of *citrate synthase* enzyme. The last mechanism is the respiratory formation of ATP by transferring electrons (not shown) from NADH to an electron acceptor, resulting in the reduction of  $O_2$  to  $H_2O$  (Shuler & Kargi, 2002). Manipulating process conditions, short-circuits the cycle, during which alternative carbon-sources such as carbon dioxide are utilised, allowing the accumulation and eventual release of citric acid outside the microorganism.

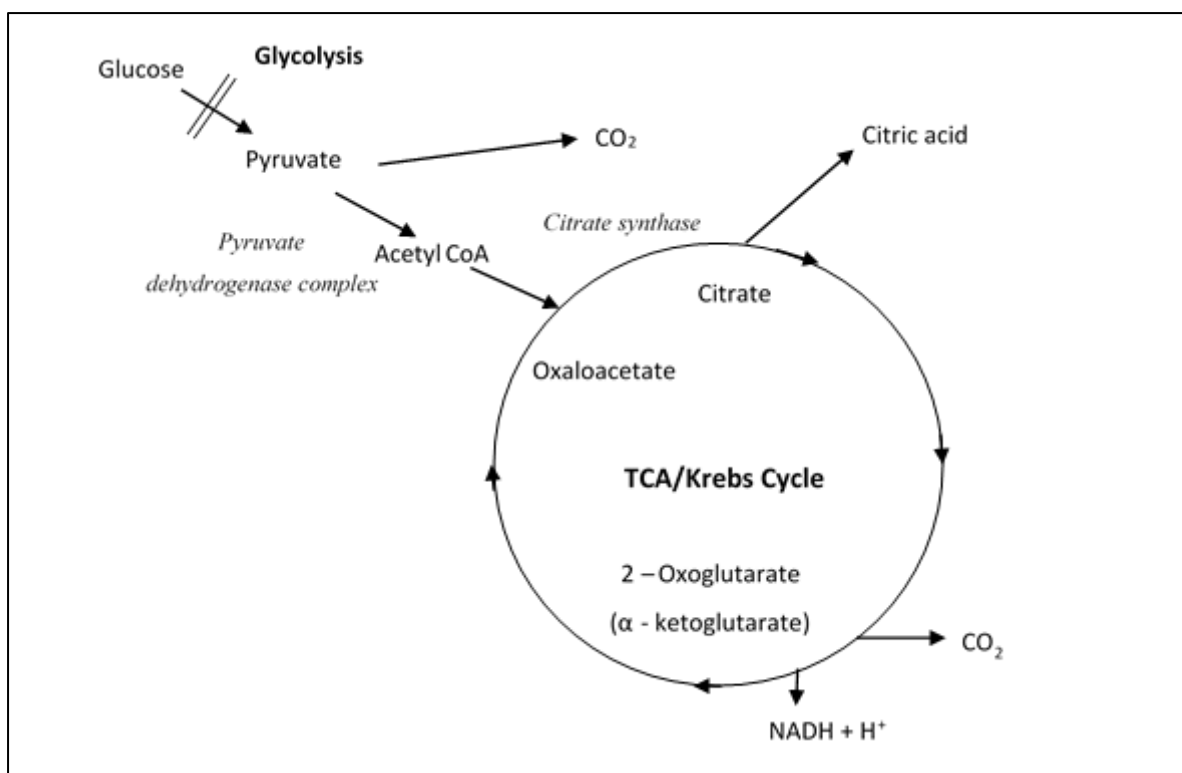


Figure 9 –Simplified Citric Acid Cycle (redrawn from (Shuler & Kargi, 2002); (Ciriminna, et al., 2017))

#### 2.5.2.5 Current Industrial Production Methods

Fermentation of glucose to citric acid utilizing *A. niger* is the current industrial practice, although strains of *Candida* yeasts have also recently seen application in citric acid production (Angumeenal & Venkappayya, 2013). Synthetic production methods do exist, however they are far more costly and impractical due to less direct reaction pathways and more severe processing conditions (Esser & Bennett, 2002).

#### Surface Fermentation

Fermentative production was traditionally achieved by cultivating glucose-consuming microorganisms on growth medium slants. The traditional feed and glucose source has primarily remained consistent over time using either cane or beet molasses for industrial production (Vandenberghe, et al., 1999). Fungal mycelium are grown on a liquid medium as a surface mat in large numbers of shallow trays that are stacked in racks and stored in a sterilized environment. According to Kristiansen, et al. (2002) trays are 50-100 litres in volume, reaching a shallow depth of 5-20 cm with a surface area of 5m<sup>2</sup>. In terms of feed preparation, a 15 % solution of molasses is initially adjusted to a pH within the range of 5-7 and then sterilized with steam. Thereafter, it is pumped into the trays described above. The inoculum spores are introduced into the trays by adding a prepared solution or directly sprayed into the trays with the incoming sterilised air streams (Kristiansen, et al., 2002).

Since the formation of citric acid is an exothermic process, considerable amounts of heat are generated. Therefore, incoming air streams serve a dual purpose of providing oxygen for the aerobic microorganisms to grow and of removing generated heat to retain the fermentation trays between 28-30 °C. Kristiansen et al., (2002) further describes that a flow rate of 10m<sup>3</sup> medium per min is required to remove an average heat generation of 1 KJ/h/m<sup>3</sup> from the fermentation medium. Humid air of between 40-60% moisture is utilised to prevent the removal of moisture from high surface area trays (Kristiansen, et al., 2002).

After a cultivation period of 8-15 days, the surface fermentation method produces approximately 1 kg/m<sup>2</sup>/day of citric acid with a 75% yield (Kristiansen, et al., 2002). The result is a spent fermentation medium and mycelia mat that are separated from each other using filtration and washing steps. The mycelia mat contains 15 % of the total citric acid (Kristiansen, et al., 2002).

Both portions (mycelia mat and washing fluid) are treated with calcium hydroxide (slaked lime) at 90 °C to form an insoluble calcium citrate precipitate in the form of tri-calcium di-citrate tetra-hydrate  $[(Ca_3 (C_6H_5O_7)_2 \cdot (H_2O)_2) \cdot 2H_2O]$ , which is further washed to remove spent medium (Osiewacz, 2002). Thereafter the calcium citrate precipitate is suspended in dilute sulphuric acid solution to form the precipitate gypsum (calcium sulphate) as well as citric acid in solution. Thereafter the solution undergoes further purification (carbon adsorption) and concentrating steps (rotary drum filtration, crystallisation and drying) to produce the final crystalline product (Esser & Bennett, 2002). The surface fermentation method is still utilised around the world in a number of older production facilities, but this method has the main drawback of being more labour intensive and requiring a larger plant footprint than submerged fermentation methods and is therefore a more costly method (Esser & Bennett, 2002).

### ***Submerged Fermentation***

In this form of fermentation the microorganism is grown submerged within the growth medium. Conversion of raw feed is typically achieved in stainless steel vessels configured as a continuous stirred tank or tower reactor where air or pure oxygen is introduced from the bottom of the vessel (Vandenberghe, et al., 1999). Aeration is to be maintained such that the oxygen concentration remains above saturation at 25 %. External water films or jackets cool the outside walls of the fermenter (Vandenberghe, et al., 1999). Since the late 1960's to the present day, glucose consuming *A. niger* strains have been cultivated through submerged fermentation, using batch configuration, producing between 150 to 180 g/l of citric acid within 5-6 days (Anastassiadis & Rehm, 2006). The advantage of

this method is a far higher production rate (more product specific), a less labour intensive process and a smaller plant footprint area in comparison to surface fermentation. This method currently produces 80% of the world supply of citric acid (Yalcin, et al., 2010).

### *Koji Process (Solid State Fermentation)*

This method was developed in Japan where feedstocks consist of solid rice bran and food wastes or other carbohydrate sources (Vandenberghe, et al., 1999). This process specifically uses *A. niger* as the microorganism of choice, with the carbohydrate source of starch for the feed. The feed is initially sterilized through the action of steam, resulting in the formation of a 70 % water content semi-solid paste at a pH of 5.5. The reaction vessel is inoculated with *A. niger* spores through spraying and held at 30 °C to ferment for 4-5 days to yield citric acid (Kristiansen, et al., 2002).

## 2.5.3 Glutamic Acid

### 2.5.3.1 Chemical Structure

Glutamic acid, also known by its name as (2S)-2-aminopentanedioic acid, is a non-essential amino acid that is synthesised within the human body and in L-glutamate salt-form and functions as an important excitatory neurotransmitter (Sapolsky, 2005). In addition, glutamic acid is utilized by the human body to synthesise the amino acid L-glutamine through glutamate ammonium ligase. Glutamine functions as a nitrogen and carbon source for muscle growth and is responsible for maintaining a strong immune system (Heys & Ashkanani, 1999). Glutamine also slows aging, by functioning as a source of micronutrients for the production of firm and healthy skin as well as acting as an acid-base regulator in the kidneys by producing ammonium (Bowtell, et al., 1999; Heys & Ashkanani, 1999).

L-glutamic acid has the chemical formula  $C_5H_9NO_4$ . The chemical structure consists of a side carboxylic group, an alpha-carboxylic acid group and an amino ( $-NH_2$ ) group as seen clearly in Figure 10 below.

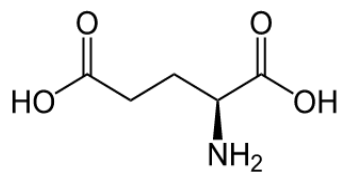


Figure 10 - Glutamic acid chemical structure

### 2.5.3.2 Industrial and Household Applications

- Salts of glutamic acid such as monosodium glutamate (MSG) are used as a flavour enhancer in food (Vishwanath, 2006)

- Glutamic acid acts as a precursor to glutathione, which is an effective antioxidant in the body, neutralising free radicals to protect cells and boost the immune system. It is therefore used in anti-aging supplements (Heys & Ashkanani, 1999)
- Used as a source of protein for food and pharmaceutical supplements for human consumption as well as cattle feed, including glutamine production (Vishwanath, 2006)

#### 2.5.3.3 Historical Discovery and Production

L-Glutamic acid was discovered by Ritthausen in 1866, by treating wheat gluten with sulphuric acid. In 1907, Professor Kikunae Ikeda identified the new umami taste component in Kombu (type of edible kelp) broth as L-glutamic acid crystals (Lakshmi & Mangala, 2011). A year later in 1908, he had discovered a method to isolate and purify the glutamic acid crystals. In 1909, Ikeda and the entrepreneur Saburosuke Suzuki formed the company Ajinomoto and began the first commercial production of the sodium salt of L-glutamic acid, monosodium glutamate (MSG) from wheat gluten protein acid hydrolysate, using hydrochloric acid (HCl) to form L-glutamic acid hydrochloride (Lakshmi & Mangala, 2011; Sano, 2009). Owing to L-glutamic acid hydrochloride salt's low solubility and high selectivity against other amino acids in the HCl hydrolysate solution, L-glutamate was easily extracted during crystallisation (Sano, 2009). Fermentative production methods were developed in 1957, when Kinoshita, (1958) discovered a glutamic acid producing strain of bacteria known as *Micrococcus glutamicus* (also known as *Corynebacterium glutamicum*) that provided a 30 g glutamic acid yield/l using glucose medium feed (Pandey, et al., 2015; Lakshmi & Mangala, 2011). Since then, fermentative production has held a number of advantages in comparison to previous chemical synthesis methods, including higher concentrations of glutamic acid (microorganisms were genetically developed and screened for maximum glutamic acid production), lower production costs and environmental impact (no formation of harmful HCl gas to labour) (Sano, 2009). Thus, most production and research during the 1950's to the present has shifted towards fermentative methods (Sano, 2009).

#### 2.5.3.4 Regulatory Pathway for Glucose Fermentation to Glutamic Acid via TCA cycle

Glutamic acid (GA) is produced from glucose via submerged fermentation under limited oxygen saturation conditions using *Coryneform* bacteria, which include bacteria such as *C. glutamicum*, *M. glutamicus* and *Brevibacterium flavum* all of which are non-motile, non-spore forming, gram-positive rod bacteria and biotin growth dependent (Sano, 2009; Lederberg, 2000). Biotin (C<sub>10</sub>H<sub>16</sub>N<sub>2</sub>O<sub>3</sub>S) is essential for GA-producing microorganism growth, but needs to be limited in order for GA to accumulate in the growth medium. As seen in Figure 11 below, biotin in the presence of the enzyme acetyl-coenzyme A, is converted to oleic acid and thereafter phospholipids (Kumar, et al., 2014). Penicillin or surfactant addition is necessary to overcome biotin limitations (allowing biotin for

abundant cell growth but not limitation of GA), allowing bacteria to still produce high quantities of glutamic acid in the medium outside the cell (Sano, 2009). Tanaka, (1960) and Li, (1965) have both reported an optimum biotin concentration between 2-3 g/l for *M. glutamicus*.

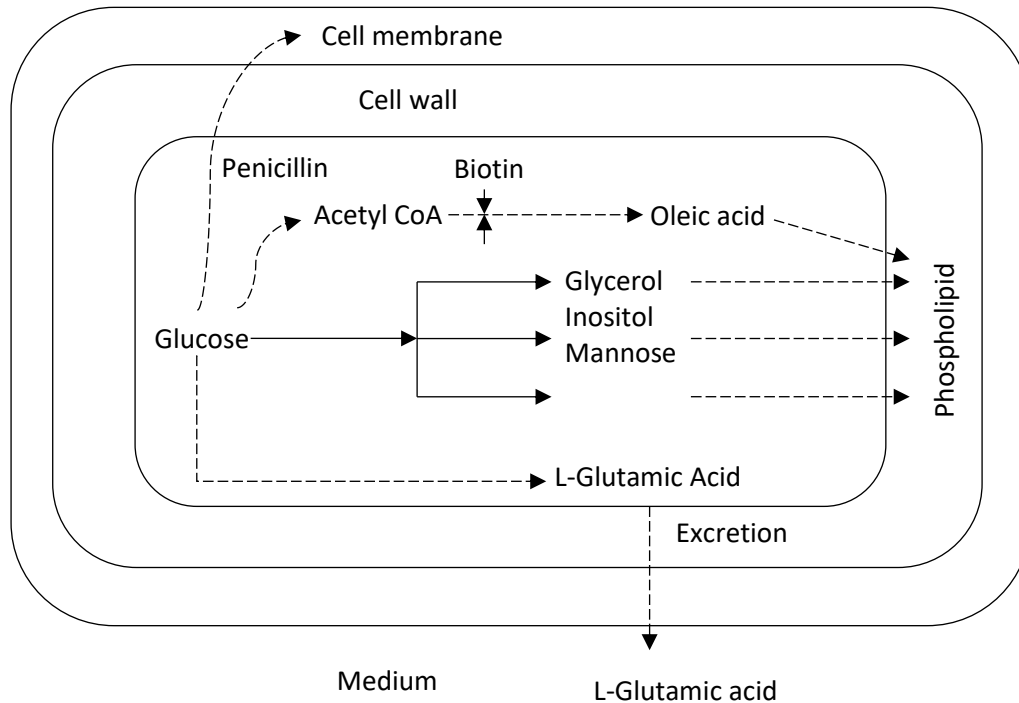
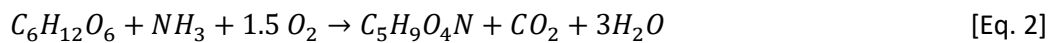


Figure 11 - Cell permeability of glutamic acid in relation to phospholipids in the cell, redrawn from (Kumar, et al., 2014)

Glutamic acid is produced in the cell aerobically through the metabolism of glucose via the tricarboxylic acid cycle (TCA – also known as the citric acid or Krebs cycle) to  $\alpha$ -ketoglutaric acid (2-oxoglutaric acid) (see Figure 12).  $\alpha$ -ketoglutaric acid subsequently reacts with ammonia in the presence of glutamate dehydrogenase (GD) enzyme to form glutamic acid (see red highlighted path in Figure 12 below for extra cellular GA production) (Sano, 2009); (Kumar, et al., 2014)). The overall stoichiometric reaction for the aerobic conversion of glucose (utilizing  $\text{NH}_3$  as a nitrogen source) to glutamic acid in *C. glutamicum* is provided below from (Lederberg, 2000) as follows:



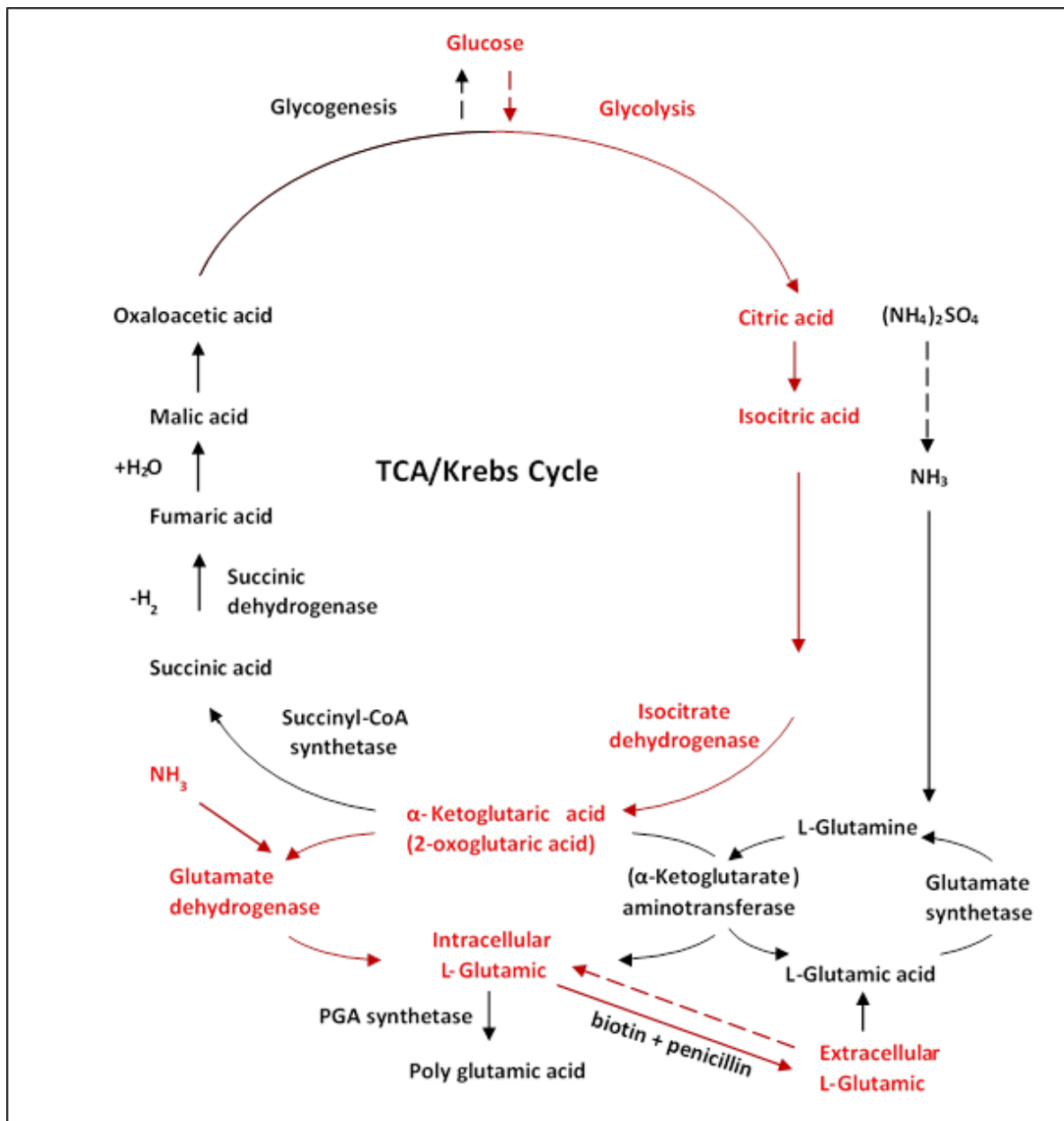


Figure 12 - Regulatory pathway for the biosynthesis of glutamic acid redrawn and adapted from (Kumar, et al., 2014)

An optimum oxygen concentration of 7 mg/l was reported (Kumar, et al., 2014). Insufficient oxygen limits the production of lactic and succinic acid in GA accumulating bacteria (Kumar, et al., 2014; Lederberg, 2000). In terms of temperature, 28-32 °C is the ideal temperature range as studied by (Tanaka, 1960). Lederberg, (2000) reported an industrial temperature range of 30 - 35°C. The addition of biotin (2-3 g/l) and penicillin to the fermentation step with an average growth temperature of 32°C is assumed (Kumar, et al., 2014). An optimal pH range of 7-8 will be maintained within the fermenter as stated by Lederberg, (2000).



## 2.6 Comparison of Microorganism Strains for CA and GA Fermentation

For this study, three main selection criteria are considered when comparing various microorganism strains for citric acid and glutamic acid production. Product yield is one of the selection criteria used to gauge various strain performances. It is predominantly defined as the mass of product produced per mass of sugar substrate consumed (provided as a percentage). In this study, the substrate glucose is liberated from sugarcane lignocellulose after undergoing pre-treatment and enzymatic hydrolysis for both citric acid and glutamic acid production. The higher the product yield, the greater the amount of chemical produced and therefore the greater the amount of revenue that can be generated for respective biorefinery scenarios. The second selection criteria that is important for determining the amount of product purification steps that will be required, is the final product concentration also frequently referred to as titre. A lower product concentration is indicative of higher costs for downstream purifications. The third selection criteria that is used to gauge microorganism strains are volumetric productivity values. This is the amount of product that is produced per volume of the reaction vessel (or in some cases mass of substrate) per unit of time. The higher the productivity, the smaller the size of the fermenters, resulting in decreased capital expenditure. Sections that follow, compare CA and GA microorganism strains based on the selection criteria described. Data that is unavailable for specific selection criteria will be calculated where possible, but will otherwise not be listed for the strain.

### 2.6.1 Citric Acid Microorganism Comparison

Table 10 - Comparison of Various Microorganisms for the Production of Citric Acid

Submerged Fermentation						
Strain	Yield, % (g/g)	Citric Acid concentration (g/l substrate)	Volumetric productivity (g/l/h)	Substrate	Reference	Reactor Type
<i>Candida oleophila</i> ATCC 20177	-	167 <sup>a</sup>	1.5 (4.58 days)	Glucose (336 g/l)	(Anastassiadis & Rehm, 2006)	Repeated fed-batch
<i>Candida oleophila</i> ATCC 20177	-	157.6 <sup>b</sup>	1.21	Glucose	(Anastassiadis & Rehm, 2006)	Repeated fed-batch
<i>Candida oleophila</i> ATCC 20177	>70	127	2.32	Glucose	(Anastassiadis, et al., 1993)	Continuous
<i>Candida oleophila</i> ATCC 20177	75	98 <sup>c</sup>	1.81	Glucose	(Anastassiadis & Rehm, 2006)	Continuous
<i>A. niger</i> GCMC-7	100	113.6	-	Cane Molasses	(Soccol, et al., 2006)	Batch
<i>A. niger</i> ATCC 9142	-	109	-	Beet molasses	(Ogawa, 1976)	-
<i>A. niger</i> ATCC 9142	-	106.65	0.89 (5 days)	Beet molasses	(Madhusudan, et al., 2010)	Batch
<i>Saccharomyces lipolytica</i>	75	95	0.7 (200 h, batch)	Glucose	(Briffaud, 1979)	Batch
<i>S. lipolytica</i>	86	75	1.16 (66 h, batch)	Glucose	(Enzminger, 1986)	-
<i>Yarrow lipolytica</i> <i>Wratislavia</i> AWG7	78	154	1.05	Sucrose	(Moeller, et al., 2013)	Repeated batch
<i>Yarrow lipolytica</i> A101	68.7	54	-	Beet molasses	(Kautola, et al., 1992)	-
<i>A. niger</i> T 55	65	-	0.387 g/g/h (168 h)	Cane Molasses	(Kundu, et al., 1984)	Batch
<i>A. niger</i> IMI-41874	45	27 g/l	0.08 g/l/h (14 days)	Wood Hemicellulose	(Maddox, et al., 1985)	Batch
<i>A. niger</i> CBS733.88	-	21.24 g/l	0.1475 g/l/h (6 days)	Glucose (Sugar cane bagasse)	(Pallares, et al., 1996)	Batch

Solid State Fermentation						
Strain	Yield, % (g/g)	Citric Acid concentration (g/kg substrate)	Productivity (g/kg/h)	Substrate	Reference	Reactor Type
<i>A. niger DS1</i>	55.9	198 g/kg cane bagasse	0.087 g/100 g dry solids/h	Sugarcane bagasse moistened with sucrose medium	(Kumar & Jain, 2008)	Packed Bed Reactor
<i>A. niger CFTRI 30</i>	79	174g /kg dry sugarcane-press mud	0.858 g/kg/h (120 h)	Sugarcane press-mud	(Shankaranand & Lonsane, 1993)	
<i>A. niger ATCC 9142</i>	97 (g CA/g sugar consumed) <sup>d</sup>	94.5 g citric acid/kg dry bagasse	1.1 g/kg/h	Sugarcane bagasse hydrolysate	(Khosravi-Darani & Zoghi, 2008)	Batch fermenter
<i>A. niger</i>	44	29	-	Cellulose hydrolysate from sugarcane	(Manonmani & Sreekantiah, 1987)	-
<i>A. niger IM-155</i>	62	-	-	Corn starch	(Nguyen, et al., 1992)	

a - Continuous mode, fill and drain technique at 80 % air saturation achieved in 4.85 days , 336 g/l glucose  
b - Continuous mode, fill and drain technique at 20 % air saturation achieved in 5.4 days  
c - chemo stat mode at 20 % air saturation  
d - max yield based on sugar consumed after pretreatment

Table 10 provides a comparison between different microorganisms that produce citric acid based on yield, final citric acid concentrations and for some strains listed, volumetric productivities for submerged and solid-state fermentation methods. Strains are arranged in descending order based on citric acid concentration. Productivity data is, for the most part, not explicitly expressed in these literature sources and is calculated by considering the fermentation residence times. In terms of solid-state fermentation methods, it can be seen that *A. niger DS1* studied by (Kumar & Jain, 2008) provides a citric acid yield of 64.5 g citric acid/ 100 g substrate with a final concentration of 198 g citric acid per kg bagasse moistened with sucrose medium. The final concentration is highest for this strain; however, this makes use of a mixed feed of solids and sucrose medium (sucrose, 310 g/l; NH<sub>4</sub>NO<sub>3</sub>, 25 g/l; MgSO<sub>4</sub>·7H<sub>2</sub>O, 2.5 g/l; CuSO<sub>4</sub>, 0.04 g/l). This study is meant to focus on the use of second-generation lignocellulosic feedstocks to produce citric acid in the proposed biorefinery scenario. *A. niger* strain ATCC 9142, however makes use of the preferred feed, sugarcane bagasse hydrolysate, which provides the highest yield of 97 % and a final product concentration of 94.5 g citric acid per kg dry bagasse. In terms of submerged fermentation, the yeast strain *Candida oleophila* ATCC 20177 provides the highest citric acid concentration (for a continuous process) as well as the highest productivity, using glucose

as feed, in comparison to the other microorganism strains listed in Table 10 above (highest conc. and productivity of 167 g/l and 2.2 g/l/h respectively).

## 2.6.2 Glutamic Acid Microorganism Comparison

Table 11 - Comparison of Various Microorganisms for the Production of Glutamic Acid (Concentration, Productivity, Yield)

Strain	Concentration (g/l)	Productivity (g/l/h)	Yield % (g/g)	Substrate	Reference	Production
<i>Corynebacterium glutamicum</i> (NCIM-2168)	175	8.3	95	Sugarcane Juice (89 % sucrose, 7% glucose, 5% fructose)	(Pal, et al., 2015) (Pal, et al., 2016)	3 stage membrane-integrated hybrid reactor system
<i>Corynebacterium glutamicum</i>	150	3.1 (calc. from 500 m <sup>3</sup> and 2 days incubation)	70	Commercial process - glucose	(Sanchez & Demain, 2014) (Lederberg, 2000)	Batch
<i>Brevibacterium divaricatum</i> NRRL 8-231	100	(3.51) 28.5 h	91	Glucose	(Miesher & Haute, 1975)	Fed-batch
<i>Brevibacterium</i> sp.	96.3	8.3	55	Glucose	Yoshioka et al.	Continuous
<i>C. glutamicum</i> ATCC 13022	93	3.8	54.9	Sugarcane Molasses	(Amin & Al-Talhi, 2007)	Batch
<i>Corynebacterium glutamicum</i> 2262	80	-	-	Glucose	(Delaunay, et al., 2002)	Fed-Batch
<i>C. glutamicum</i> ATCC 13022	73	29.1	75.7	Sugarcane Molasses	(Amin & Al-Talhi, 2007)	Continuous
<i>C. glutamicum</i> ATCC 13022	53.7	-	53.7	Glucose	Aoki et al.	Batch
<i>Brevibacterium</i> (mutant)	52	-	-	Glucose	(Yoshimura, et al., 1982)	-
<i>C. glutamicum</i>	41.42	1.2	53	Glucose	Sun-Uk et al.	Batch
<i>Brevibacterium</i> sp. Tc452	41.4	-	-	Glucose	(S.U. Choi, 2004)	-
<i>Micrococcus glutamicus</i> + <i>pseudomonas reptilivora</i>	37.1	-	-	Glucose	(Sunitha, et al., 1998)	-
<i>Micrococcus glutamicus</i> / <i>Corynebacterium glutamicum</i>	30	-	-	glucose-ammonium (7-10 % glucose is optimum, Ting, 1959)	(S. Kinoshita, 1957)	-
<i>Bacillus</i> strain 14B22	12.5	-	-	Glucose	(Foster, 1959)	-
<i>Brevibacterium</i> species	80 g/kg bagasse	-	-	Sugarcane bagasse (10% glucose), dry bagasse	(Nampoothiri & Pandey, 1996)	SSF, Batch

Table 11 shows strains of glutamic acid producing microorganisms that utilise glucose and glucose-containing substrates. Strains are arranged in descending order in terms of GA concentrations. *C. glutamicum* NCIM-2168 strain produces the highest concentration and yield of 175 g glutamic acid/l and 95% (g/g) respectively, utilizing a sugarcane juice medium (88% sucrose, 7% glucose and 5% fructose) in a membrane-integrated hybrid reactor system as studied by (Pal, et al., 2015). *Bacillus* strain 14B22 species produced the lowest glutamic acid concentration of 12.5 g/l, where the feed utilised in the reactor is glucose (Foster, 1959). However, in terms of productivity performance, *C. glutamicum* ATCC 13022 has the highest productivity of 29.1 g/l/h, with a GA concentration of 73 g/l (continuous operation), meaning that selection of this microorganism is favourable in reducing capital expenditure.

### 3. Study Objectives

In order to ensure future economic competitiveness of sugar mills in SA, it is proposed that multi-product biorefineries with electricity co-generation, utilising sugarcane bagasse and trash as feedstock, be annexed to existing mills. The economic feasibility of xylitol, citric acid (CA) and glutamic acid (GA) biorefinery scenarios annexed to existing sugar mills is analysed individually and compared to a stand-alone CHP baseline scenario. Xylitol is to be produced through batch catalytic hydrogenation of xylose over a Raney nickel catalyst (Mikkola, et al., 1999; Melaja & Hamalainen, 1977). Whereas, CA and GA production will take place under continuous submerged fermentation of glucose utilising their respective microorganisms and process steps (see Table 12). Reasons for the selection of the respective catalyst and microorganisms are discussed and concluded further in the methodology chapter (refer to section 4.4.2).

Table 12 - Biorefinery Scenarios Investigated

Process Conditions	Xylitol Scenario	CA Scenario	GA Scenario	Baseline
Production route	Chemical	Fermentative	Fermentative	n/a
Reactor Configuration	Batch	Continuous	Continuous	n/a
Substrate Utilised	Xylose	Glucose	Glucose	Raw bagasse and trash
Pretreatment	2-stage DA hydrolysis	2-stage DA hydrolysis	2-stage DA hydrolysis	n/a
Revenues				
Xylitol	✓	-	-	-
Citric acid	-	✓	-	-
Glutamic acid	-	-	✓	-
Electricity	✓	✓	✓	✓

The key objectives of this study are summarised as follows:

- Develop 3 conceptual product biorefinery scenarios and model process units in Aspen Plus® using data and process conditions listed in literature.
- Determine whether utilising sugarcane lignocellulose in product biorefinery scenarios with cogeneration is more economically feasible than a stand-alone CHP baseline scenario where bagasse and trash is burnt for electricity and steam (is it more profitable to simply burn bagasse or to process it to products)
- Identify biorefinery scenarios capable of generating profit under the process specifications and model assumptions collected from literature (specifically which biorefinery scenarios are profitable)

- Identify key parameters for which internal rate of return (IRR) values are most sensitive to change and gain insight into what conditions can be altered to improve scenario profitability (which parameters is the profitability of the respective scenarios most sensitive)
- Determine the environmental impact of each biorefinery scenario. (see section 5.7 for environmental impact characterisation)

## 4. Methodology

The following section describes the methodology and various steps required to develop techno-economic analyses to complete the various study objectives given. The software, thermodynamic models, biorefinery process specifications, model descriptions and various economic cost and cash flow sheet assumptions are described further.

### 4.1 Software

Aspen Plus<sup>®</sup> V8.8 process simulation software is used to model the process units for the xylitol, CA and GA biorefinery scenarios, in order to acquire mass and energy balance streams and utility summaries for process equipment (electricity, steam and water consumption). Microsoft Excel<sup>®</sup> 2013 spreadsheets are used to analyse the stream and utility data acquired from Aspen Plus<sup>®</sup> simulations and calculate the key design specifications and costs of equipment and determine biorefinery operating expenditure and ultimately calculate economic indicators of each biorefinery scenario.

### 4.2 Thermodynamic Model

The Electrolyte Non-Random Two-Liquid (ELEC-NRTL<sup>™</sup>) activity coefficient model is used to determine state parameters for all process equipment for each biorefinery scenario with the exception of the auxiliary CHP plant equipment where the STEAM NBS<sup>™</sup> model is used. ELEC-NRTL<sup>™</sup> is able to model mixed solvent and aqueous electrolyte systems and models activity coefficients for ionic species. The ELEC-NRTL<sup>™</sup> model reduces to the regular NRTL<sup>™</sup> model when electrolyte concentrations approach zero (Renon & Prausnitz, 1968). The STEAM NBS<sup>™</sup> property method uses steam table correlations to determine the thermodynamic state properties of pure water and steam systems for temperatures ranging from 273.15 - 2000 K (Haar, et al., 1984).

### 4.3 Techno-economic Analysis Steps

Figure 13 shows the steps required to satisfy the objectives of this investigation. The first step requires a thorough review of past literature related to the current study and thereafter to use this information to construct a process flow diagram (PFD) for each product scenario (xylitol, CA and GA) and then to construct an Aspen Plus<sup>®</sup> process model that calculates the mass and energy flows associated with each process unit. Biorefinery scenarios are required to be energy self-sufficient and will each have different electricity and steam demands based on production steps. The greater the quantity of lignocellulose diverted to the combined heat and power (CHP) plant's boiler for energy, the less there



is available for conversion to products and therefore the lower the energy consumption of a biorefinery. Iterations are made to the biorefinery models' bypass of biomass feed to the CHP plant's boiler, until the biorefinery meets its energy and heating requirements, as well as the steam requirements of the adjacent sugar mill (120 tons/h steam). Process equipment is sized after determining the mass and energy flows for the optimised bypass, as well as the utility consumption of individual process units (steam, electricity and water consumption). Thereafter, capital and operating expenses (CAPEX and OPEX) are calculated and used to determine key financial parameters in a discounted cash flow (DCF) sheet and ultimately analyse whether the individual studied biorefineries annexed to the existing sugar mill are profitable or not. The main financial indicators that will be calculated and used to determine relative profitability, include the net present value (NPV), the internal rate of return (IRR), the total capital investment (TCI) as well as the pay-back period (PBP). These are elaborated and defined in section 4.6.4 (Table 37)

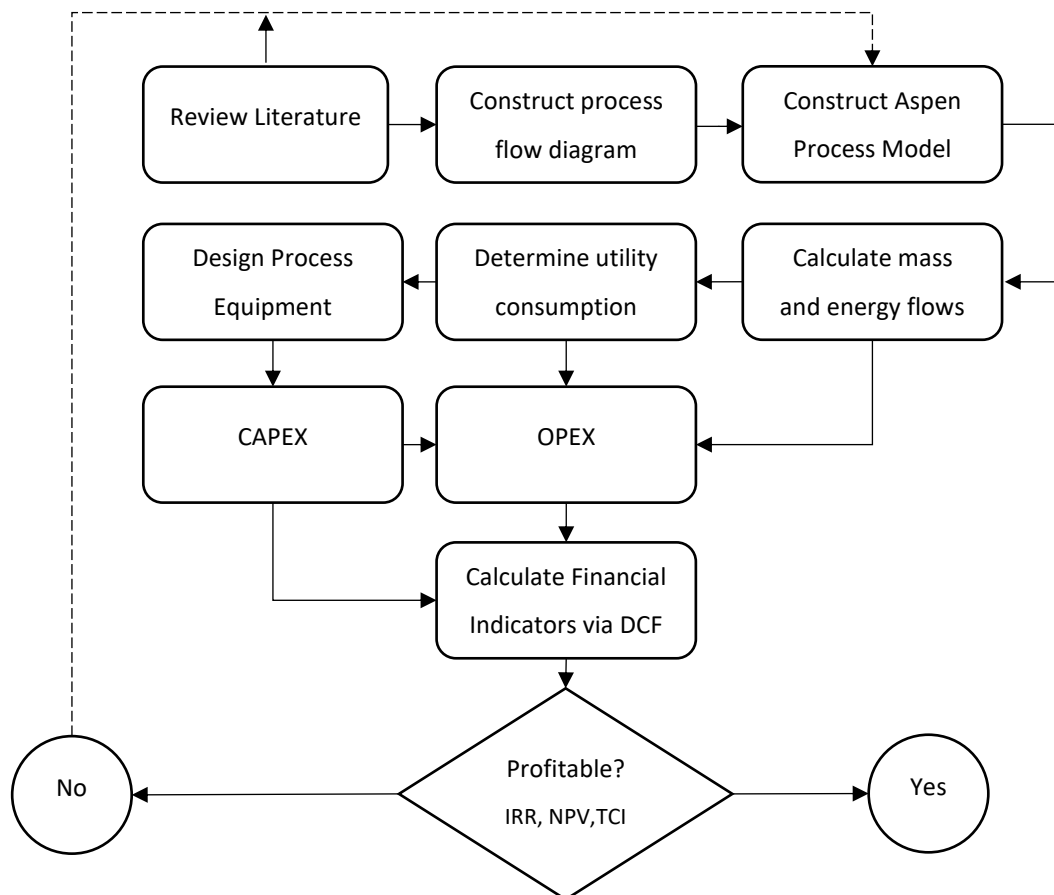


Figure 13 - Steps Required to Perform Economic Analyses

The next section provides an overview of the design assumptions used to construct the process and financial models for each biorefinery scenario. Thereafter, biorefinery scenario process descriptions

are provided that detail the production steps and process conditions required in order to produce the investigated products.

## 4.4 Biorefinery Process Specifications

### 4.4.1 Biorefinery Lignocellulosic Feedstock

Table 13 - Composition of Lignocellulosic Bagasse and Trash Feedstock for Biorefinery Scenarios (Petersen, et al., 2014)

Component	Fraction % (w/w)	Flow (ton/h)
Cellulose	40.7	26.5
Hemicellulose (Glucan)	3.9	2.5
Hemicellulose (Xylan)	17.7	11.5
Hemicellulose (Arabinan)	3.3	2.1
Hemicellulose (Acetate/Acetyl Group)	2.2	1.4
Lignin	21.9	14.3
Ash	3.5	2.3
Extractive	6.7	4.4
<b>Sum Dry Mass</b>	<b>100</b>	<b>65</b>
<b>Water</b>		<b>48.5</b>
<b>Total (Liquid + Solid)</b>		<b>113.53</b>

113.5 tons/h of lignocellulosic biomass is assumed as input for each biorefinery scenario. Table 13 summarises the average polysaccharide compositions of sugarcane bagasse and trash mixed feedstock for various South African sugarcane cultivars (Petersen, et al., 2014). The various hemicellulose polymer compositions were determined by averaging the sugar hydrolysate concentrations after DA hydrolysis for various lignocellulosic feedstocks and reverse calculating the parent hemicellulose chain mass fractions (Moutta, et al., 2011; Cao, et al., 2009; Canilha, et al., 2008; Chandel, et al., 2007; Baek & Kwon, 2007; Sepúlveda-Huerta E, 2006).

### 4.4.2 Catalyst and Microorganism Selection for Biorefinery Scenarios

This section explains the rationale for the catalyst and microorganism selection for Xylitol, CA and GA biorefinery scenarios. The selection of microorganisms for CA and GA biorefinery scenarios are based on a desirability score (Equation 3) ranging from 0 to 1, weighted as a percentage (%w) according to the microorganisms yield, concentration and productivity (Etzkorn, 2012). Each data point ( $X_i$ ) for each of the parameters (yield, concentration or productivity) are normalised against unity ( $X_{i, 0 \text{ to } 1}$ ) according to Equation 4, where  $X_{\max}$  and  $X_{\min}$  represent the maxima and minima among all data points.

$$Desirability = \frac{[(X_{yield} \times \%W_{yield}) + (X_{Conc} \times \%W_{Conc}) + (X_{productivity} \times \%W_{productivity})]}{100\%} \quad [Eq. 3]$$

$$X_{i,0 \text{ to } 1} = \frac{X_i - X_{min}}{X_{max} - X_{min}} \quad [Eq. 4]$$

#### 4.4.2.1 Citric Acid (CA) Scenario Microorganism Selection

Figure 14, ranks CA-producing microorganism strains for batch and continuous submerged fermentation, according to a desirability between a value of 0 and 1, based on a 10% weighting towards productivity and 45 % weighting towards yield and concentration respectively, as described in previous studies (Naleli, 2016; Etzkorn, 2012).

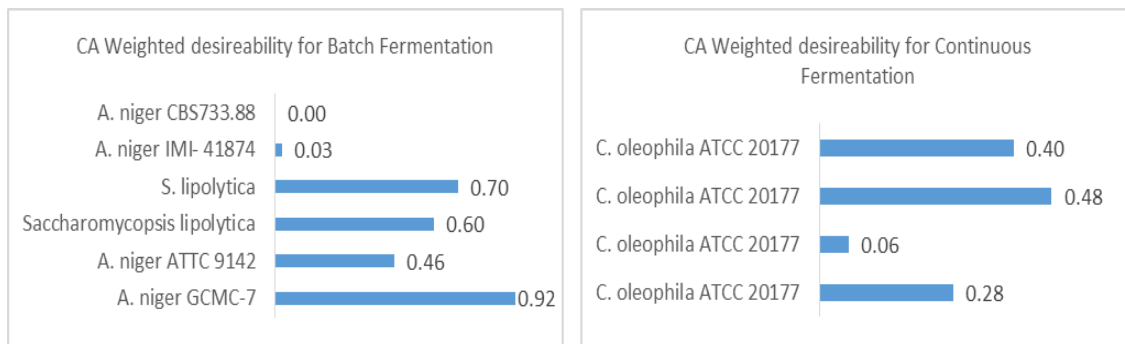


Figure 14 - CA microorganism weighted desirability for batch and continuous production

It can be seen from Figure 14 above, that the strains *A. niger GCMC-7* and *C. oleophila ATCC 20177* achieve the highest desirability in terms of final product concentration, yield and productivity for batch and continuous (including fed-batch) fermentation respectively.

#### 4.4.2.2 Glutamic Acid (GA) Scenario Microorganism Selection

Figure 15 below, ranks GA-producing microorganism strains for batch and continuous submerged fermentation, according to a desirability score between a value of 0 and 1, based on a 10% weighting towards productivity and 45 % weighting towards yield and concentration respectively (Naleli, 2016).

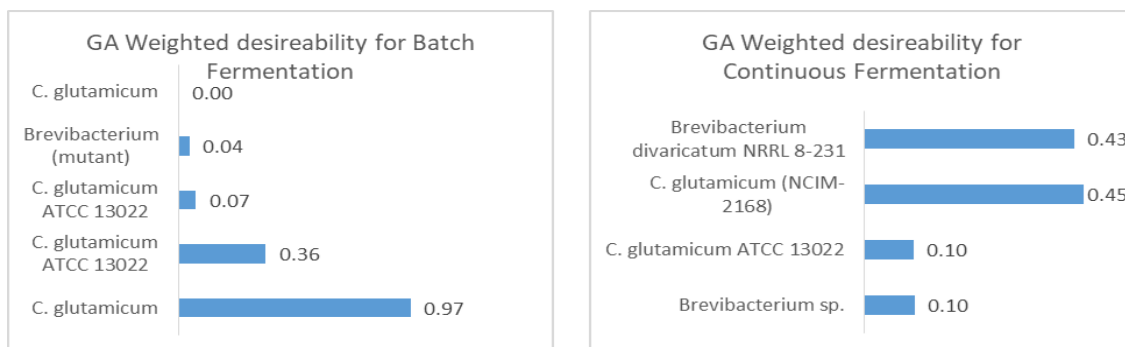


Figure 15 - GA microorganism weighted desirability for batch and continuous production

*C. glutamicum* for batch fermentation and *C. glutamicum* NCIM 2168 and *Brevibacterium divaricatum* NRRL 8-231 for continuous fermentation, are seen as the most desirable microorganisms to select, in terms of final GA concentration, yield and productivity respectively (refer to Figure 15). Production strategies, range from novel processes, to conventional batch, fed-batch and continuous systems as shown in Table 11, section 2.6.2. Continuous processes, report higher productivities and GA yields ranging from 8.3 to 29.1 g/l/h and 55 to 76 % (g/g) in comparison to batch production ranging lower from 1.2 to 3.8 g/l/h and 53-55 % (w/w) respectively (Pal, et al., 2015; Amin & Al-Talhi, 2007). Thus, continuous production is favourable in reducing production time and fermenter capital expenditure.

#### 4.4.2.3 Catalyst Selection for Xylitol Scenario

Yadav, et al., (2011) reported catalytic activities for 3 different Ruthenium (Ru) supported catalysts and a Raney-Ni catalyst and characterised performance in terms of xylose conversion, xylitol and arabinitol selectivity and yield (Table 14).

Table 14 - Comparison of Xylitol Selectivity and Yields Using Different Catalysts adapted from (Yadav, et al., 2011)

Catalysts	% Conversion	Xylitol		Arabinitol		Ave. Cost (\$/kg catalyst)
		% Selc.	% Yield	% Selc.	%Yield	
Ru(1.0%)/NiO(5.0%)–TiO <sub>2</sub>	99.9	99.8	99.7	0.1	0.1	6500-25000
Ru (1.0%)/TiO <sub>2</sub>	97.1	99	96.1	0.1	0.1	6500
Ru (1.0%)/C	96.5	97.5	94	0.2	0.2	73.2
Raney-Ni	96.9	96.7	93.7	1	0.9	18.5

As seen in Table 14, the first catalyst (1% Ru, 5% NiO-TiO<sub>2</sub>) provides the highest product (xylitol) selectivity and yield and the lowest impurity (arabinitol) selectivity in comparison to the subsequent catalysts. The Raney-nickel catalyst provides the third highest xylose conversion of 96.9% and overall has the lowest xylitol selectivity and yield and the highest arabinitol selectivity and yield compared to the other catalysts, although the difference is marginal in comparison.

In terms of cost per unit mass of catalyst, the Raney-Ni catalyst retails lowest at \$18.25/kg catalyst from suppliers in Thailand (Zauba, 2016). Whereas suppliers in China quote prices between \$14-24/kg catalyst, providing an average price of \$19/kg (Alibaba, 2016). Ruthenium and particularly catalysts with traces of titanium (Ti) are priced far higher (Yadav, et al., 2011) with values ranging from as high as \$25/g (\$25000/kg) and other suppliers quoting \$3-8/g (average of \$6500/kg) (Alibaba, 2016). Therefore, the selection of the Raney-Ni catalyst, currently employed in chemical xylitol production,

is the most economical in comparison to the remaining catalysts listed above while still achieving relatively high conversions and selectivities (Heikkila, et al., 1992; Melaja & Hamalainen, 1977).

Hydrogenation of xylose to xylitol typically requires high temperature and pressure conditions. Temperatures and pressures of 120°C at 55 bar for 2h respectively have been reported by (Dinesh Kumar Mishra, 2013; Yadav, et al., 2011) for Ru Zeolite-Y supported catalytic system, and 120°C at 55 bar using ruthenium catalyst on NiO modified TiO<sub>2</sub> support (Yadav, et al., 2011). Whereas, other sources made use of various types of supported Raney-Ni catalysts under process conditions of 80-140°C for temperature and 40-70 bar for pressures (Mikkola, et al., 2003; 2000; 1999).

#### 4.4.3 Process Decisions

- For the xylitol biorefinery scenario the Raney-nickel catalyst was selected based on a lower cost per mass of catalyst (Zaubas, 2016), while simultaneously being able to achieve high conversions and selectivity in comparison to other catalysts. In addition, the catalyst has been extensively studied in the past and is therefore reinforced as a reliable and commercially viable choice (Mountraki, et al., 2017; Yadav, et al., 2011; Mikkola, et al., (2003; 2000; 1999); Melaja & Hamalainen, 1977).
- A Raney-nickel catalyst loading of 5 wt.% is used (Mikkola, et al., 1999).
- Microorganism selection was based on the previously discussed desirability score. *Candida oleophila* ATCC 20177 is chosen as the microorganism for submerged continuous fermentation of glucose to CA due to the yeasts high desirability score, partially achieved due to its high final CA concentration (167 g CA/l), respectable productivity (1.5 g/l/h; 4.85 days fermentation time) and ability to ferment with high glucose feed concentrations (336 g/l glucose in feed medium - high osmo-tolerance) (Anastassiadis & Rehm, 2006).
- *Brevibacterium divaricatum* NRRL 8-231 is a patented microorganism chosen for GA production from glucose utilising continuous submerged fermentation of glucose derived from sugarcane lignocellulose hydrolysate. The choice is based on its desirability score, achieved due to its high final GA concentration (100 g/l) and complete conversion of glucose after 28.5 h (Miesher & Haute, 1975).
- The RSTOIC<sup>®</sup> model from Aspen Plus<sup>®</sup> together with the respective molar conversions for parallel and series reactions is used to model all chemical reactors for each biorefinery scenario.
- Separator blocks are used to model all adsorption, chromatography and filtration separation processes, where the outlet stream purity or the impurity removal is well known.

- All evaporators, crystallisers and drying units are modelled as flash drums in the Aspen Plus® model, where the primary function is to either increase substrate sugar concentrations by removing water or to remove any inhibitors remaining after adsorption.

#### *4.4.3 Process Assumptions*

- Enzymes are assumed to be continuously replaced for CA and GA biorefineries and are not produced on-site. For hydrolysis it is assumed an enzyme dosage of 20 mg/g cellulose fed to the enzymatic hydrolysis reactor (Humbird, et al., 2011; Roche, et al., 2009).
- During heat integration between processes modelled in Aspen Plus®, 10% of energy in the form of heat is assumed to be lost to the surroundings due to process inefficiencies.
- 50% of the Raney-nickel catalyst is assumed to be recovered after the catalytic reaction.
- The steam demand of the annexed sugar mill is assumed to be 0.4 tons of steam per ton of sugarcane processed. The whole sugarcane production line (mill and biorefinery with cogeneration) is assumed to handle a total biomass feed of 300 tons sugarcane/h (Petersen, et al., 2014) and therefore in addition to the respective biorefinery scenarios meeting their own energy and heating needs, the scenarios must also be able to contribute 120 tons/h of steam to the annexed sugar mill. 113.5 tons/h of sugarcane lignocellulosic feed enters each of the biorefineries with CHP plant integration

## 4.5 Biorefinery Scenario Descriptions

The following section describes the various production steps and unit operations for the investigated biorefinery scenarios (CHP baseline, xylitol with electricity co-production, CA with electricity co-production and GA with electricity co-production). This section provides context for results presented in the results section. Mass and energy balances along with Aspen Plus process flow diagrams for xylitol, citric acid and glutamic acid product scenarios have been provided in Appendix A3, B3 and C3 respectively.

### 4.5.1 Standalone CHP Baseline Scenario

The CHP baseline scenario is described by referring to Figure 16 as well as Appendix D (Figure 81), where all available lignocellulose is used to provide process energy to the sugar mill as well as surplus electricity for sale. Within this scenario, 100% of the sugarcane lignocellulose (stream 502) is transported via a conveyor belt (CONV-501) to a storage hopper (HOP-501) before being sent through to the combustion chamber of the boiler (R-501). Air (stream 503) is blown into the combustion chamber to fuel the reactions that release CO<sub>2</sub>, H<sub>2</sub>O and heat. The combustion reactions modelled for the CHP baseline's boiler are provided in Table 15.

Table 15 - CHP Plant Baseline Combustion Reactions

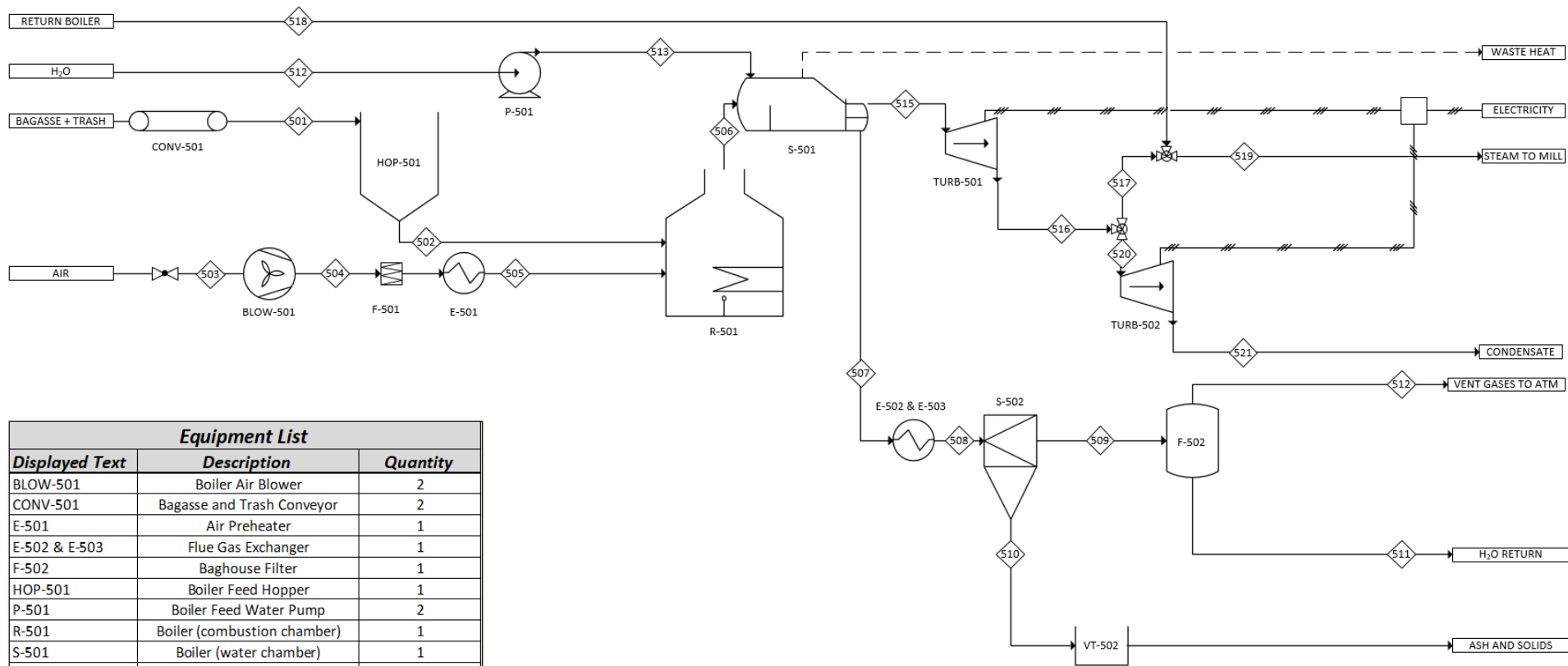
Reactant	Reaction	Product	Conversion
<b>Cellulose</b>	$Cellulose + 6O_2 \rightarrow 5H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100
<b>Glucan</b>	$Glucan + 6O_2 \rightarrow 5H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100
<b>Xylan</b>	$Xylan + 5O_2 \rightarrow 4H_2O + 5CO_2$	Steam, heat, CO <sub>2</sub>	100
<b>Arabinan</b>	$Arabinan + 5O_2 \rightarrow 4H_2O + 5CO_2$	Steam, heat, CO <sub>2</sub>	100
<b>Acetate</b>	$Acetate + 2O_2 \rightarrow 2H_2O + 2CO_2$	Steam, heat, CO <sub>2</sub>	100
<b>Lignin</b>	$Lignin + 8.5O_2 \rightarrow 4H_2O + 8CO_2$	Steam, heat, CO <sub>2</sub>	100
<b>Extract</b>	$Extract + 6O_2 \rightarrow 6H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100

The steam and heat formed is circulated through a heat exchange chamber (E-502 & E-503) adjacent to the wall of a water boiler (S-501), where 185 tons/h of water (stream 13) is vaporised to steam (stream 15) before entering the first stage of a condensing steam turbine (CEST) (TURB-501). The flue gases released during combustion (stream 507) are sent to a dust cyclone (S-502) and baghouse filter (F-502) to remove any ash and solid particles, which are stored in a skip bin (VT-502) before being removed for solids waste collection. 185 tons/h of steam at 857 °C and 64 atm (stream 515) pass through the first stage of the CEST, to produce 16.1 MW of electricity and steam exiting (stream 516) at 694 °C and 28 atm. A portion of the exiting steam is split to the sugar mill and mixed with water in

a let-down station (V-501) to form 120 tons/h of steam (stream 519). The remaining portion of steam (stream 520) is passed through the second stage of the CEST where 40.8 MW of electricity is produced. A part of the resulting exiting steam (80 tons/h at 125°C, stream 521) is mixed in the let-down station to form a partial closed loop of water and meet the steam requirement of the sugar mill.



## COMBINED HEAT AND POWER PLANT BASELINE



<b>Equipment List</b>		
<b>Displayed Text</b>	<b>Description</b>	<b>Quantity</b>
BLOW-501	Boiler Air Blower	2
CONV-501	Bagasse and Trash Conveyor	2
E-501	Air Preheater	1
E-502 & E-503	Flue Gas Exchanger	1
F-502	Baghouse Filter	1
HOP-501	Boiler Feed Hopper	1
P-501	Boiler Feed Water Pump	2
R-501	Boiler (combustion chamber)	1
S-501	Boiler (water chamber)	1
S-502	Flue Gas Cyclone	1
TURB-501	Condensing Steam Turbine (Stage 1)	1
TURB-502	Condensing Steam Turbine (Stage 2)	1
VT-502	Solids Skip Bin	1

Figure 16 - CHP Baseline Flow Sheet

#### 4.5.2 Xylitol and Electricity Co-production Scenario

Xylitol production can be split into six sections, namely pretreatment (area 100), detoxification (area 200), catalytic reaction (area 300), purification (area 400), the CHP plant (area 500) and wastewater treatment (WWT, area 600). The production steps are described for each respective area. Appendix A2 (Figure 55 to Figure 61) can be referred to further understand the configuration of process units in Aspen Plus®.

##### 4.5.2.1 Pretreatment (Area 100)

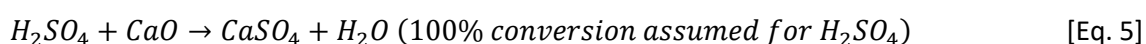
Figure 17 shows the PFD for the xylitol biorefinery pretreatment section (Area 100). 113.5 tons/h of sugarcane lignocellulosic biomass (bagasse and trash) is initially stored in a vessel (VT-101) and then loaded onto the first conveyor belt (CONV-101). A portion of the raw feed is bypassed to the xylitol biorefinery via a shorter conveyor (CONV-102) and a portion to the CHP plant (Area 500). An optimised bypass to the CHP plant (Area 500) was determined iteratively until the steam requirements of the adjacent sugar mill and the steam and electricity needs of the xylitol biorefinery itself were satisfied (see section 5.1).

Lignocellulosic biomass entering the biorefinery is sent through pretreatment to firstly undergo hydrolysis in the dilute acid (DA) tank reactor (R-101). Pre-mixed DA at a concentration of 2.9 g H<sub>2</sub>SO<sub>4</sub>/100 ml H<sub>2</sub>O (stream 6) is sent to the DA tank reactor such that a solids to liquid ratio of 0.25 ton lignocellulose/ton DA is maintained ( $S/L = \frac{1}{4}$ ) (Moutta, et al., 2011). During DA hydrolysis, the contents of the reactor are held at a temperature of 130°C for 30 min by heating with steam (stream 2) to liberate xylose, arabinose, glucose and inhibitor (furfural, 5-HMF and acetic acid) monomers from the main and side chains of the hemicellulosic polymers (Moutta, et al., 2011). The crystalline cellulose and lignin portions (cellulignin) (apart from a small percentage of amorphous cellulose) are insoluble, whereas the monomer sugars and inhibitor groups are assumed to be fully dissolved in the exiting hemicellulosic hydrolysate stream (stream 9). A list of the key reactions occurring during DA hydrolysis are presented in Table 16 (Humbird, et al., 2011).

Table 16 - DA Hydrolysis Chemical Reactions and Conversions (Humbird, et al., 2011)

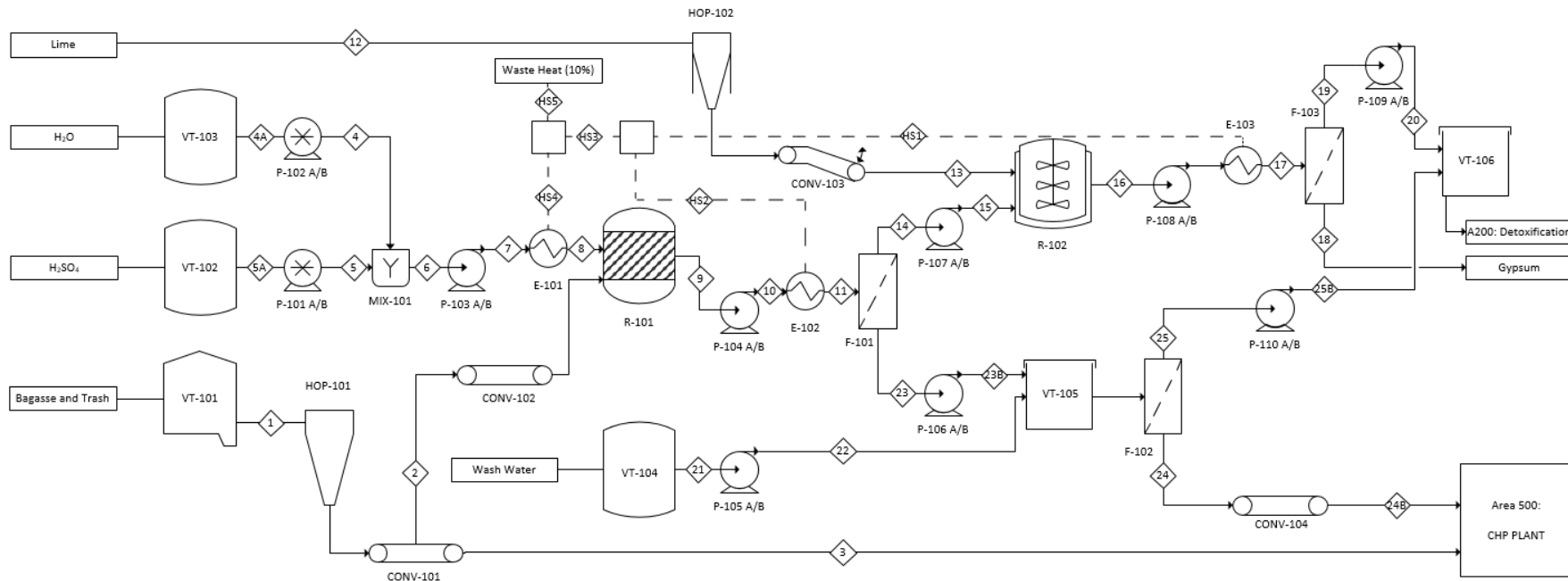
Reactant	Reaction	Product	Conversion (%)
Cellulose	$(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_6H_{12}O_6$	Glucose	9.9
Cellulose	$2(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_{12}H_{22}O_{11}$	Cellobiose	0.3
Hemicellulose (Glucan)	$(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_6H_{12}O_6$	Glucose	9.9
Hemicellulose (Xylan)	$(C_5H_8O_4)_n + nH_2O \rightarrow nC_5H_{10}O_5$	Xylose	90
Hemicellulose (Arabinan)	$(C_5H_8O_4)_n + nH_2O \rightarrow nC_5H_{10}O_5$	Arabinose	100
Hemicellulose (Acetyl)	$(CH_3COO^- - H^+)_n \rightarrow nCH_3COOH$	Acetic Acid	100
Xylose	$C_5H_{10}O_5 \rightarrow C_5H_4O_2 + 3H_2O$	Furfural	4.17
Glucose	$C_6H_{12}O_6 \rightarrow C_6H_6O_3 + 3H_2O$	HMF	3.97
Lignin	$C_8H_8O_3 \rightarrow C_8H_8O_3$	Soluble lignin comp	1.5

The resulting hydrolysate (stream 9) is then sent through a cooler (E-102) to bring the temperature down from 130°C to 60°C. The cooled hydrolysate (stream 11) is then filtered through presses (F-101) to separate the insoluble cellulignin portion (including the solid ash and extract) (stream 23), from the soluble sugars and inhibitors (stream 14), such that the minimum moisture content of the exiting insoluble cellulignin (stream 23) is 50% (w/w). The separated hydrolysate rich in sugars (stream 14) is thereafter transported to a lime tank (R-102), where quick-lime (CaO, which will be referred to plainly as lime throughout this study) (stream 12) is released from a hopper (HOP-102) and transported via a solids conveyor (CONV-103) to the tank to neutralise the H<sub>2</sub>SO<sub>4</sub> and prevent possible downstream catalyst deactivation. During neutralisation, the H<sub>2</sub>SO<sub>4</sub> in the hemicellulosic hydrolysate reacts exothermically with lime to form solid gypsum (CaSO<sub>4</sub>) precipitate as seen in Equation 5.



CaO is supplied at 15% in excess of stoichiometric requirements and a 15 min reaction time is assumed according to industrial practices (Lewis & Boyton, 2000). Excess lime and gypsum (stream 18) is separated in a solids filter (F-103) from the sugar-rich neutralised hydrolysate (stream 19). The cellulignin (stream 23) from the filter presses (F-101) is then sent through to a series of wash water filtration tanks (represented as VT-105 and F-102) in order to recover any remaining sugars (stream 25) contained within the moisture. The washed and filtered cellulignin slurry exiting the wash tanks (stream 24) is sent via conveyor (CONV-104) to the CHP plant (Area 500) for combustion to produce electricity and steam. The recovered sugars are sent to a holding tank (VT-106) to be mixed with the sugar-rich hydrolysate (stream 19) to form a combined hydrolysate (stream 26) rich in pentoses (mostly xylose) (Figure 17).

# Area 100: Pretreatment



Equipment List					
Equipment ID	Description	Quantity	Equipment ID	Description	Quantity
CONV-101	Bagasse and Trash Conveyor	2	P-104 A/B	Dilute Acid Hydrolysate Pump	2
CONV-102	Bypassed Biomass Conveyor	2	P-105 A/B	Hemicellulosic Hydrolysate Pump	2
CONV-103	Lime Conveyor	2	P-106 A/B	Lime Tank Exit Pump	2
CONV-104	Cellulignin Conveyor	2	P-107 A/B	Cellulignin Slurry Pump	2
E-101	DA Preheater	1	P-108 A/B	Wash Water Pump	2
E-102	DA Hydrolysate Cooler	1	P-109 A/B	Neutralised Hydrolysate Pump	2
E-103	Lime Tank Cooling Jacket	1	P-110 A/B	Recovered Sugars Pump	2
F-101	Cellulignin-Sugars Filter Press	18	R-101	Dilute Acid Tank Reactor (2 stage)	1
F-102	Gypsum Rotary Filter Drum	35	R-102	Lime Neutralising Tank	1
F-103	Cellulignin Filter Press	3	VT-101	Bagasse and Trash Holding Tank	1
HOP-101	Bagasse and Trash Hopper	1	VT-102	Sulphuric Acid Storage Tank	1
HOP-102	Lime Hopper	1	VT-103	Dilution Water Storage Tank	1
P-101 A/B	Sulphuric Acid Pump	2	VT-104	Wash Water Storage Tank	1
P-102 A/B	Dilution Water Pump	2	VT-105	Wash Water Filtration Tank	3
P-103 A/B	Dilute Acid Feed Pump	2	VT-106	Hydrolysate Sugar Holding Tank	1

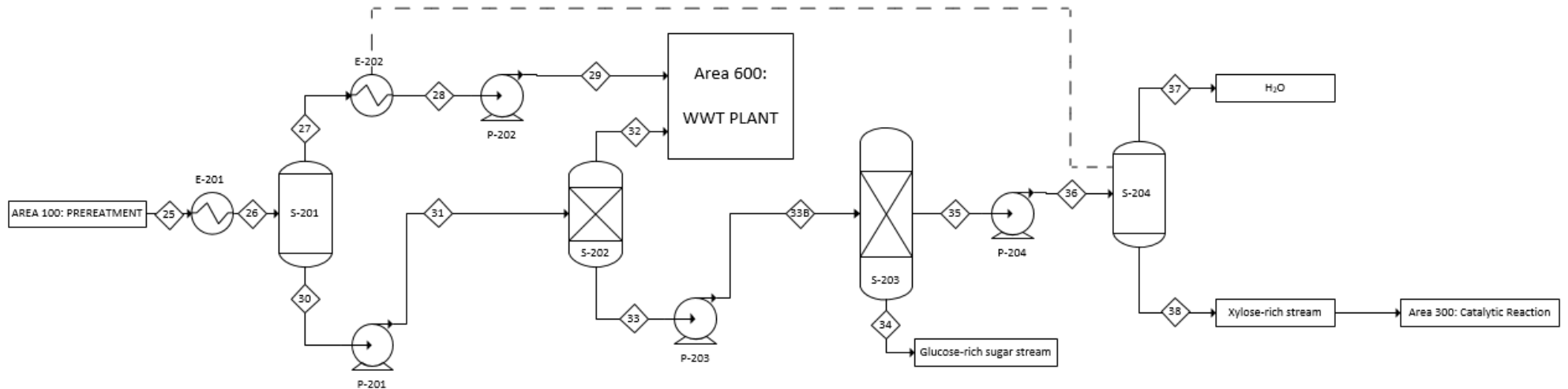
Figure 17 - Xylitol Biorefinery Pretreatment Section (Area 100)

#### 4.5.2.2 Detoxification (Area 200)

Figure 18 provides an overview of the detoxification section (Area 200) of the xylitol biorefinery. The objective of this section of the biorefinery is to produce a xylose-rich stream by removing any inhibitors and competing sugars (arabinose and glucose) that may interfere with down-stream catalytic reactions (catalyst deactivation) and produce by-products. This reduces the number of down-stream purification steps that will be required.

The combined sugar-rich hydrolysate exiting the mixing tank (from Area 100) is preheated in a heat exchanger (E-201) (stream 26) before entering an evaporation column (S-201). Here, the hydrolysate is heated to remove a large majority of the inhibitors as well as a portion of the water. The inhibitor rich vapour (stream 29) is then cooled in a condenser (E-202), before being pumped to the wastewater treatment (WWT) plant (Area 600) to undergo anaerobic and aerobic digestion. The xylose-rich hydrolysate (stream 31), is thereafter pumped to an activated carbon (AC) adsorption column (S-202), where the remaining inhibitors (stream 32) are removed and sent through to WWT (Area 600). The exiting detoxified xylose-rich fraction (stream 33) is then transported to a simulated moving bed (SMB) ion-exchange chromatography column (S-203), the purpose of which is to separate xylose from arabinose and glucose. The unit is packed with a weak base anionic (WBA) resin consisting of porous microspheres coated with  $\text{SO}_4^{2-}$  functional groups made from a methacrylate structure cross-linked with 6 % DVB (Pia, Heikkila, & Hurme, 2010). The WBA resin displays higher relative separation factors for both xylose and arabinose compared to glucose. Glucose is therefore completely removed, together with a portion of the arabinose (stream 34). The purified xylose-rich fraction (stream 35) is then sent to a second evaporation column (S-204) to remove enough water (stream 37) to produce a xylose-rich feed (stream 38) with a minimum xylose content of 78 wt.% on (dry basis) (Melaja & Hamalainen, 1977), before being transported to the catalytic biorefinery section (Area 300).

# Area 200: Detoxification



Equipment List		
Equipment ID	Description	Quantity
E-201	Evaporator Preheater	1
E-202	Condenser	1
P-201	AC Column Feed Pump	2
P-202	Wastewater Feed Pump	2
P-203	SMB Column Feed Pump	2
P-204	Crystalliser/Dryer Feed Pump	2
S-201	Inhibitor Evaporator	1
S-202	AC Column	1
S-203	SMB Chromatography Column	1
S-204	Crystalliser/Dryer	1

Figure 18 - Xylitol Detoxification Section (Area 200)

#### 4.5.2.3 Catalytic Reaction and Purification (Area 300 and 400)

Figure 19 provides the flowsheets for the reaction (Area 300) and purification (Area 400) plant sections for the xylitol biorefinery. After exiting the evaporation column (S-204) and achieving the desired sugar concentration (78 wt%), the xylose-rich feed is sent to a set of 3 batch catalytic reactors (R-301). Raney-nickel catalyst is mixed in the feed with a solids loading, equivalent to 5 % (w/w) of the xylose content in the entering stream (Mikkola, et al., 2000). The reactors are pressurised using hydrogen gas that enters at a pressure of 40 atm (C-301 A/B) (stream 48). The reactor contents are heated to a temperature of 135°C for 2.5 hours, where xylose is catalytically hydrogenated over the Raney-nickel catalyst active sites to form xylitol (Mikkola, et al., 1999; Melaja & Hamalainen, 1977). The remaining arabinose in the xylose-rich feed (stream 45), as well as a small part of the xylose are converted to arabinitol. The catalytic hydrogenation reactions are listed in Table 17 (Yadav, et al., 2011; Mikkola, et al., 1999).

Table 17 - Catalytic Reactions of Xylose and Arabinose to Xylitol and Arabinitol

Reactant	Reaction	Product	Conversion
<b>Xylose</b>	$C_5H_{10}O_5 + H_2 \rightarrow C_5H_{12}O_5$	Xylitol	97
<b>Arabinose</b>	$C_5H_{10}O_5 + H_2 \rightarrow C_5H_{12}O_5$	D-Arabinitol	97
<b>Xylose</b>	$Xylose (C_5H_{10}O_5) \rightarrow xylulose (C_5H_{10}O_5)$	D-Xylulose	2
<b>Xylulose</b>	$C_5H_{10}O_5 + H_2 \rightarrow C_5H_{12}O_5$	D-Arabinitol	100
<b>Xylose</b>	$C_5H_{10}O_5 \rightarrow C_5H_4O_2 + 3H_2O$	Furfural	0

After exiting the reactor, the xylitol-rich product is cooled in a heat exchanger (E-303) from 135°C to 65°C to prepare for further downstream purification (Area 400). Once the xylitol product has formed in the catalytic reaction section (Area 300), it needs to be purified and dried in order to create a marketable product. Once cooled, the xylitol-rich product (stream 46) is transported to a second SMB ion-exchange chromatography column (S-401). The unit is loaded with a strong acid cation (SAC) resin, consisting of porous microspheres containing  $Ca^{2+}$  functional groups, supported on a polystyrene sulfonate structure cross-linked with 3.5 % DVB (Melaja & Hamalainen, 1977). The SBA resin displays higher relative separation factors for both xylitol and arabinitol and allows high separation into a purified xylitol syrup (stream 48) and an arabinitol-rich syrup (stream 47). The xylitol-rich syrup is pumped to an evaporation column (S-402) to remove moisture (stream 49) and produce a high concentration syrup (stream 50) before it is sent for crystallisation and drying (S-403) to remove any remaining moisture (stream 51) to produce high purity crystalline xylitol product (stream 52).

## Area 300: Catalytic Reactor

## Area 400: Purification

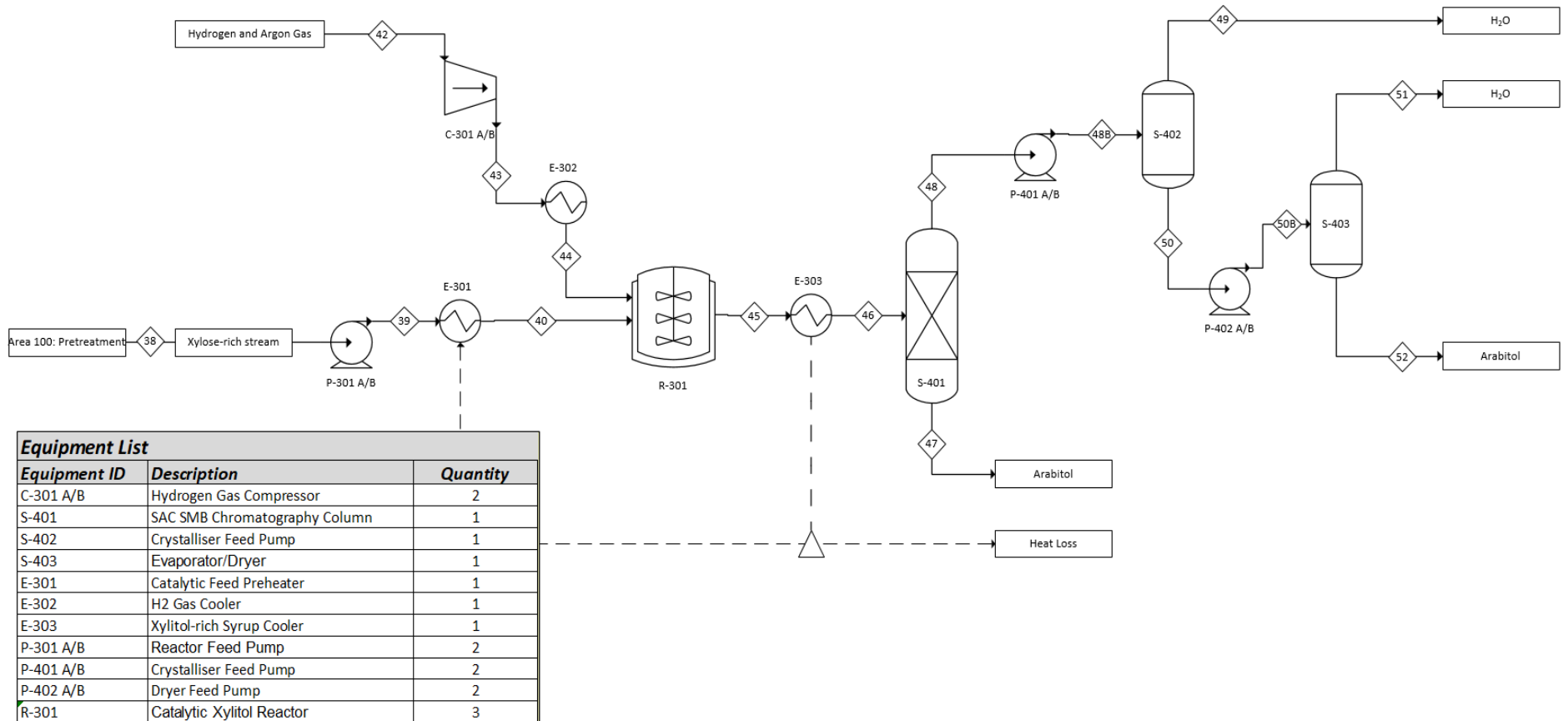


Figure 19 - Xylitol Catalytic Reaction and Purification Section (Area 300 & Area 400)



#### 4.5.2.4 Wastewater Treatment Plant (Area 600)

Figure 20 shows the WWT section (Area 600) of the xylitol scenario. The objective of this section is to reduce the chemical oxygen demand (COD) of the incoming wastewater and produce methane for additional energy production. Anaerobic digestion and activated sludge treatment (aerobic digestion) are the two main technologies utilized to treat wastewater for the product biorefineries investigated. Anaerobic digestion for the production of bio-methane and hydrogen has been reported to successfully treat sugar hydrolysates from lignocellulose pretreatment without any significant inhibition from by-products (acetic acid, furans and phenolic components) and is therefore an applicable technology for the present work (Akobi, et al., 2017; Monlau, et al., 2014; Barakat, et al., 2012).

Inhibitors (stream 601) from the evaporation column (S-201) and waste sugars (stream 602) from the carbon adsorption column (S-202) in the detoxification section (Area 200) are initially stored in the wastewater feed holding tank (VT-601). From there the inhibitors and sugars (stream 604) are pumped to an anaerobic digester (R-601) along with compressed ammonia (stream 605) which is used as a nitrogen source. During anaerobic digestion, 4 main reactions take place, namely acidogenesis where organic components react to form volatile fatty acids (VFA), acetogenesis where VFAs are converted to acetic acid and methanogenesis where acetic acid and any unreacted VFAs are converted to CO<sub>2</sub> and CH<sub>4</sub> (Tchobanoglous, et al., 2004). Propionic and butyric acid are considered as the main forms of VFA's produced with ammonia being assumed as the initial nitrogen source for biomass protein formation (C<sub>5</sub>H<sub>7</sub>NO<sub>2</sub>) as seen in Table 18 (Rajendran, et al., 2014; Tchobanoglous, et al., 2004). The WWT plant needs to process 0.8 MI/day of wastewater with an estimated theoretical oxygen demand of 27 000 mg/l. Anaerobic digestion is suitable for treating large concentrations of organic wastes and is able to achieve an 80 - 90 % reduction in chemical oxygen demand (COD) (Tchobanoglous, et al., 2004). Gases produced during anaerobic digestion (stream 8) are sent to a gas vent and scrubber (S-101) before being compressed and transported to the CHP plant section of the biorefinery (Area 500). The mixed liquor exiting the anaerobic digester (stream 606) is cooled from 52°C to a temperature of 40 °C (E-601) before being pumped to the membrane integrated activated sludge system (stream 610 to R-602).

Table 18 - Anaerobic Digestion Reactions (Rajendran, et al., 2014; Ostrem, 2004; Kalyuzhnyi, 1997; Bilitewski, et al., 1997)

Reactant	Reaction	Main Product	Conversion (%)
<b>Glucose</b>	GLUCOSE + 0.1115 NH <sub>3</sub> --> 0.0115 BIOMASS + 0.744 AACID + 0.5 PRO-ACID + 0.4409 BUT-ACID + 0.951007 CO <sub>2</sub> + 1.0254 H <sub>2</sub> O	VFAs	77
<b>Xylose</b>	3 XYLOSE + 5 H <sub>2</sub> O --> 5 AACID + 10 H <sub>2</sub> + 5 CO <sub>2</sub>	Acetic Acid	77
<b>Arabinose</b>	3 ARABINOS + 5 H <sub>2</sub> O --> 5 AACID + 10 H <sub>2</sub> + 5 CO <sub>2</sub>	Acetic Acid	77
<b>Propionic Acid</b>	PRO-ACID + 0.06198 NH <sub>3</sub> + 0.314336 H <sub>2</sub> O --> 0.06198 BIOMASS + 0.9345 AACID + 0.660412 CH <sub>4</sub> + 0.160688 CO <sub>2</sub> + 0.00055 H <sub>2</sub>	Methane	95
<b>Butyric Acid</b>	BUT-ACID + 0.0653 NH <sub>3</sub> + 0.8038 H <sub>2</sub> O + 0.0006 H <sub>2</sub> + 0.5543 CO <sub>2</sub> --> 0.0653 BIOMASS + 1.8909 AACID + 0.446 CH <sub>4</sub>	Methane	85
<b>Hydrogen</b>	14.4976 H <sub>2</sub> + 3.8334 CO <sub>2</sub> + 0.0836 NH <sub>3</sub> --> 0.0836 BIOMASS + 3.4154 CH <sub>4</sub> + 7.4996 H <sub>2</sub> O	Methane	100
<b>Acetic Acid</b>	AACID + 0.022 NH <sub>3</sub> --> 0.022 BIOMASS + 0.945 CH <sub>4</sub> + 0.066 H <sub>2</sub> O + 0.945 CO <sub>2</sub>	Methane	90
<b>Furfural</b>	2 FURFURAL + 6 H <sub>2</sub> O --> 5 CO <sub>2</sub> + 5 CH <sub>4</sub>	Methane	100
<b>HMF</b>	HMF + 3 H <sub>2</sub> O --> 3 CH <sub>4</sub> + 3 CO <sub>2</sub>	Methane	100
<b>Carbon dioxide</b>	CO <sub>2</sub> + H <sub>2</sub> O --> HCO <sub>3</sub> <sup>-</sup> + H <sup>+</sup>	Carbonic Acid	9.5

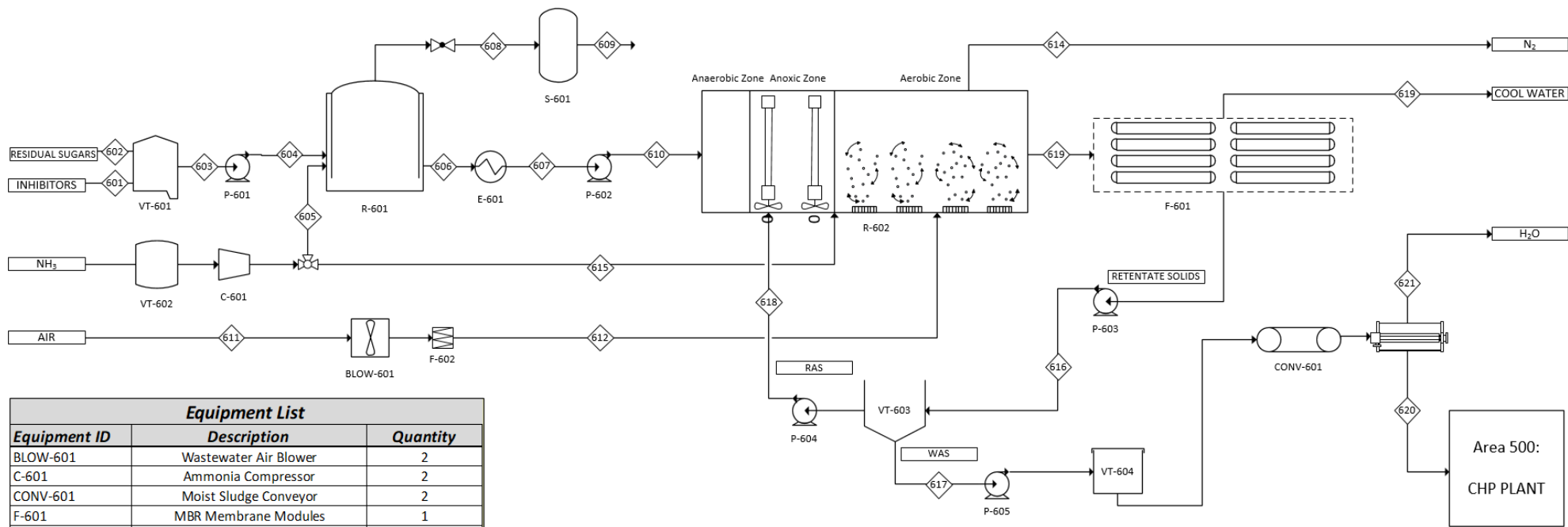
After leaving the anaerobic digester and being cooled, the mixed liquor (stream 610) needs to undergo aerobic digestion (R-602) to remove the remaining oxygen demand exerted by the organic components in the wastewater. Table 19 provides a list of stoichiometric equations used to account for various components, including sugars, inhibitors and acidic components. Different chemical process stages in WWT have also been accounted for, including nitrification (conversion of ammonia to nitrates) in an anoxic stage and denitrification in an aerobic stage (conversion nitrates to nitrogen) (Tchobanoglous, et al., 2004). Air (stream 612) is blown into the digester, together with ammonia (stream 615) as a nitrogen source for denitrifying bacteria.

Table 19 - Aerobic Digestion Reactions (Tchobanoglous, et al., 2004)

Reactant	Reaction	Main Product	Conversion (%)
<b>Glucose</b>	GLUCOSE + O <sub>2</sub> + NH <sub>3</sub> --> BIOMASS + 4 H <sub>2</sub> O + CO <sub>2</sub>	Biomass	100
<b>Xylose</b>	2 XYLOSE + 5 O <sub>2</sub> + NH <sub>3</sub> --> BIOMASS + 5 CO <sub>2</sub> + 8 H <sub>2</sub> O	Biomass	100
<b>Arabinose</b>	2 ARABINOS + 5 O <sub>2</sub> + NH <sub>3</sub> --> BIOMASS + 5 CO <sub>2</sub> + 8 H <sub>2</sub> O	Biomass	100
<b>Biomass</b>	BIOMASS + 5 O <sub>2</sub> --> 5 CO <sub>2</sub> + NH <sub>3</sub> + 2 H <sub>2</sub> O	Ammonia	33
<b>Ammonia</b>	NH <sub>3</sub> + 2 O <sub>2</sub> --> NO <sub>3</sub> <sup>-</sup> + H <sup>+</sup> + H <sub>2</sub> O	Nitrate	10
<b>Acetic Acid</b>	5 AACID + 8 NO <sub>3</sub> <sup>-</sup> --> 4 N <sub>2</sub> + 10 CO <sub>2</sub> + 6 H <sub>2</sub> O + 8 OH <sup>-</sup>	Nitrogen	100
<b>Glucose</b>	H <sup>+</sup> + OH <sup>-</sup> --> H <sub>2</sub> O	Water	100
<b>Acetic Acid</b>	AACID + 2 O <sub>2</sub> --> 2 CO <sub>2</sub> + 2 H <sub>2</sub> O	Mixed vapour	100
<b>Phenolic</b>	2 LIG-SOL + 17 O <sub>2</sub> --> 8 H <sub>2</sub> O + 16 CO <sub>2</sub>	Mixed vapour	100
<b>Propionic Acid</b>	2 PRO-ACID + 7 O <sub>2</sub> --> 6 H <sub>2</sub> O + 6 CO <sub>2</sub>	Mixed vapour	100
<b>Butyric Acid</b>	BUT-ACID + 5 O <sub>2</sub> --> 4 H <sub>2</sub> O + 4 CO <sub>2</sub>	Mixed vapour	100

Gases produced during the aerobic reactions (stream 614) are naturally vented to the atmosphere (N<sub>2</sub>, O<sub>2</sub>, CO<sub>2</sub>, H<sub>2</sub>O and excess NH<sub>3</sub>). The mixed liquor suspended solids (MLSS) exiting the activated sludge digester (stream 615) is then filtered through membrane modules (F-601) where solids sludge including biomass growth (stream 616) is separated from the water (stream 619). The separated sludge (stream 616) is then sent to a sludge holding tank (VT-603). A portion of the activated sludge is returned (RAS) back to the anoxic zone of the activated sludge reactor (R-602) to re-seed the reactor with bacteria, whereas the other portion of the activated sludge from the tank is wasted (WAS) as a purge and is sent to belt presses to produce a solids cake that can either be burnt in the boilers in the CHP plant section (Area 500) or that can be used as fertilizer for any agriculture in the vicinity of the biorefinery.

## AREA 600: WASTEWATER TREATMENT PLANT



Equipment List		
Equipment ID	Description	Quantity
BLOW-601	Wastewater Air Blower	2
C-601	Ammonia Compressor	2
CONV-601	Moist Sludge Conveyor	2
F-601	MBR Membrane Modules	1
F-602	Air Filter	1
E-601	Anaerobic Cooler	1
P-601	Wastewater Feed Pump	2
P-602	Activated Sludge MBR Feed Pump	2
P-603	Solids Retentate Pump	2
P-604	RAS Pump	2
P-605	WAS Pump	2
R-601	Anaerobic Digester	1
VT-601	Wastewater Holding Tank	1
VT-602	Ammonia Gas Cylinder	1
VT-603	Activated Sludge Holding Tank	1
VT-604	Waste Sludge Holding Tank	1
S-601	Natural Gas Scrubber	1

Figure 20 - Xylitol WWT Plant (Area 600)

#### 4.5.2.5 CHP Plant (Area 500)

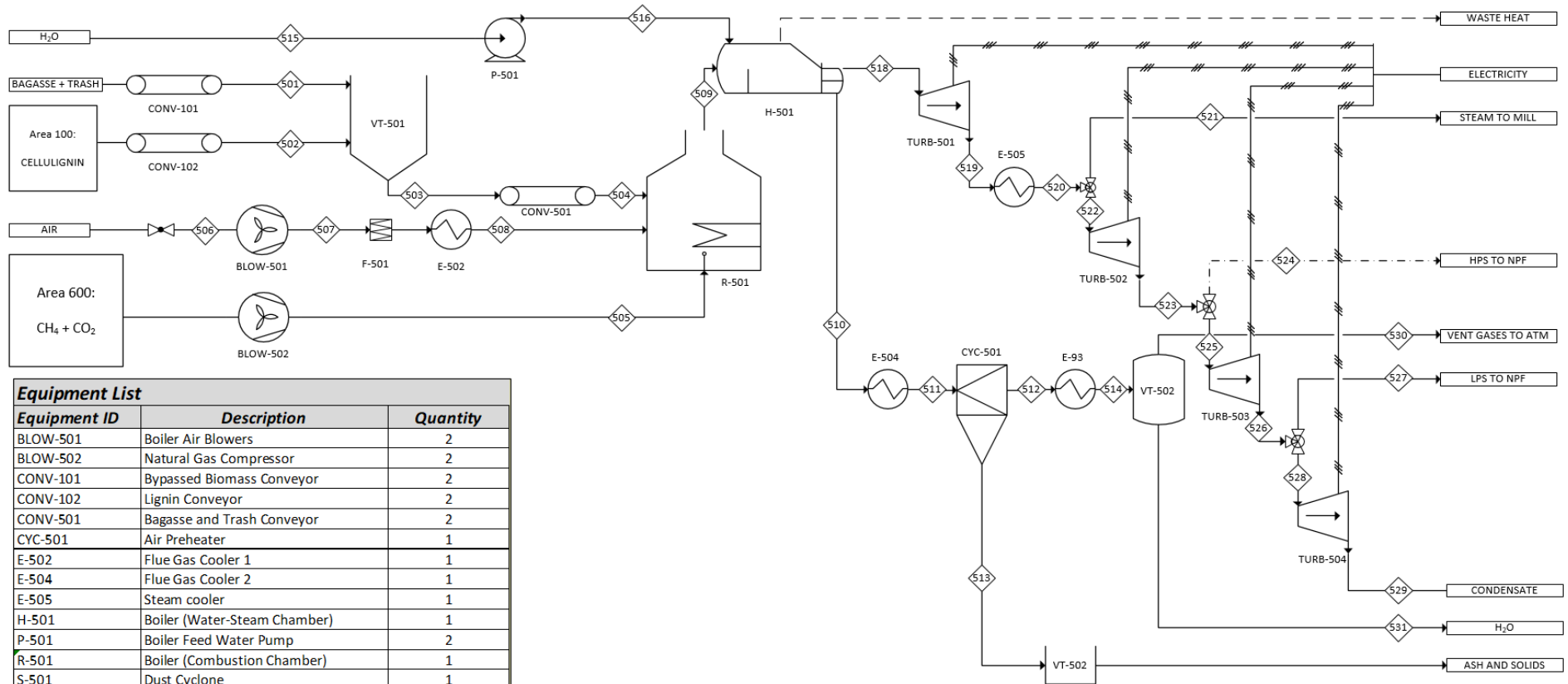
Figure 21 provides a layout of the CHP plant section of the xylitol biorefinery, whereas the flowsheet of the Aspen model for this section is provided in Appendix B (A2.6. Area 500 Combined Heat and Power Plant). The purpose of this section is to allow the biorefinery to produce its own electrical and heating utility needs, in order to make the plant self-sustaining and produce additional steam for the adjacent sugar mill (not shown).

Within the CHP plant (Area 500), bypassed bagasse and trash (stream 501), cellulignin slurry with residual sugars (stream 502) and methane gas (stream 505) (produced during anaerobic digestion) are combusted in a fired heater (R-501). The heater is modelled as an RSTOIC<sup>®</sup> reactor in Aspen Plus<sup>®</sup>, where air is blown into the combustion chamber to control the temperature of the reaction at a maximum boiler temperature of 870 °C (refer to Table 20 for combustion reactions). The resulting flue gas from combustion (stream 509 and 518) transfers its heat to the feed water (stream 516) sent to the boiler (H-501), where the water is heated to high-pressure steam (stream 518). 223 tons/h of high-pressure steam at 422 °C and 64 atm enters a 2-stage condensing steam turbine (CEST). The turbine is modelled as 4 turbine units in Aspen Plus<sup>®</sup>. After the first turbine unit (TURB-501), steam exits at 352 °C and 28 bar and enters a cooler to bring a portion of the steam to 340°C before 120 tons/h of it (stream 521) is sent to the sugar mill as a design requirement. The remaining portion of steam (103 tons/h) is sent through the remaining stages of the CEST (TURB-502, TURB-503 & TURB-504) to produce electricity and low pressure steam (LPS) for the xylitol biorefinery's own electrical and heating needs.

Table 20 - Boiler Combustion Reactions (Görgens, et al., 2015)

Reactant	Reaction	Product	Conversion
Cellulose	$Cellulose + 6O_2 \rightarrow 5H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100
Glucan	$Glucan + 6O_2 \rightarrow 5H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100
Xylan	$Xylan + 5O_2 \rightarrow 4H_2O + 5CO_2$	Steam, heat, CO <sub>2</sub>	100
Arabinan	$Arabinan + 5O_2 \rightarrow 4H_2O + 5CO_2$	Steam, heat, CO <sub>2</sub>	100
Acetate	$Acetate + 2O_2 \rightarrow 2H_2O + 2CO_2$	Steam, heat, CO <sub>2</sub>	100
Lignin	$Lignin + 8.5O_2 \rightarrow 4H_2O + 8CO_2$	Steam, heat, CO <sub>2</sub>	100
Cellobiose	$Cellulose + 12O_2 \rightarrow 11H_2O + 12CO_2$	Steam, heat, CO <sub>2</sub>	100
Glucose	$Glucose + 6O_2 \rightarrow 6H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100
Xylose	$Xylose + 5O_2 \rightarrow 5H_2O + 5CO_2$	Steam, heat, CO <sub>2</sub>	100
Arabinose	$Arabinose + 5O_2 \rightarrow 5H_2O + 5CO_2$	Steam, heat, CO <sub>2</sub>	100
Furfural	$Furfural + 5O_2 \rightarrow 2H_2O + 5CO_2$	Steam, heat, CO <sub>2</sub>	100
Acetic Acid	$Acetic Acid + 2O_2 \rightarrow 2H_2O + 2CO_2$	Steam, heat, CO <sub>2</sub>	100
HMF	$HMF + 6O_2 \rightarrow 3H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100
Extract	$Extract + 6O_2 \rightarrow 6H_2O + 6CO_2$	Steam, heat, CO <sub>2</sub>	100

## AREA 500: COMBINED HEAT AND POWER PLANT



Equipment List		
Equipment ID	Description	Quantity
BLOW-501	Boiler Air Blowers	2
BLOW-502	Natural Gas Compressor	2
CONV-101	Bypassed Biomass Conveyor	2
CONV-102	Lignin Conveyor	2
CONV-501	Bagasse and Trash Conveyor	2
CYC-501	Air Preheater	1
E-502	Flue Gas Cooler 1	1
E-504	Flue Gas Cooler 2	1
E-505	Steam cooler	1
H-501	Boiler (Water-Steam Chamber)	1
P-501	Boiler Feed Water Pump	2
R-501	Boiler (Combustion Chamber)	1
S-501	Dust Cyclone	1
S-502	Gas Vent with Bag Filter	1
TURB-501	Condensing Steam Turbine (2-stage)	1
TURB-502	Condensing Steam Turbine (2-stage)	0
TURB-503	Condensing Steam Turbine (2-stage)	0
TURB-504	Condensing Steam Turbine (2-stage)	0
VT-501	Bypassed Biomass Hopper	1
VT-502	Ash and Dust Skip Bin	1

Figure 21 - Xylitol CHP Plant (Area 500)

#### 4.5.3 Citric Acid (CA) and Electricity Co-production Scenario

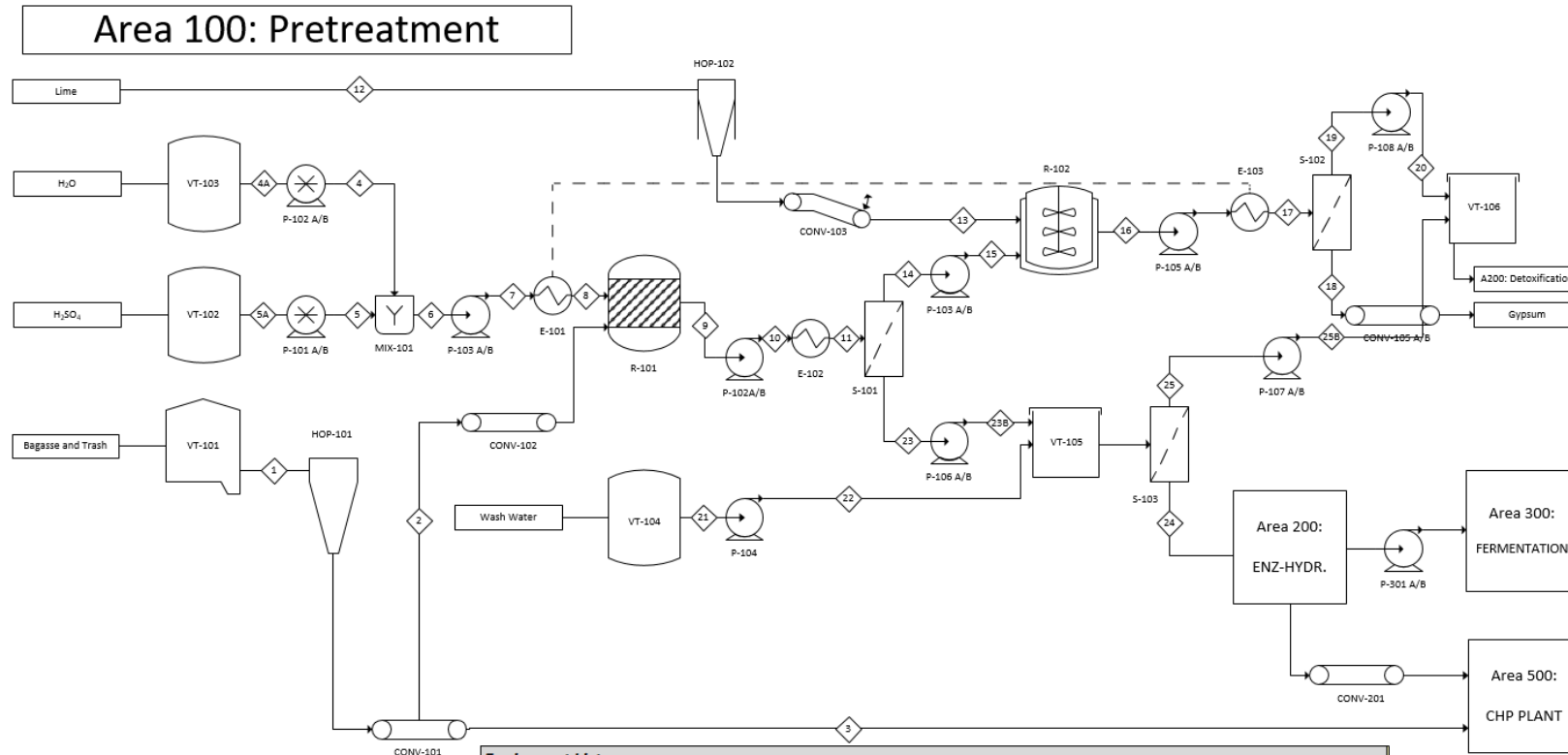
Citric acid (CA) follows a cellulosic valorisation pathway and production is divided into 7 sections. These include, pretreatment (area 100), enzymatic hydrolysis (area 200), fermentation (area 300), purification (area 400), the CHP plant (area 500), WWT (area 600) and the fermentation seed train (area 700). Production steps are discussed for areas 100-400. CA process steps for area 500 and 600 (Figure 26 and Figure 27) are similar to xylitol production and have previously been discussed.

##### 4.5.3.1 Pretreatment (Area 100)

Figure 22 shows the pretreatment section of the CA plant. 113.5 tons/h of sugarcane lignocellulose is sent via conveyor (CONV-101) in much the same way as for the xylitol biorefinery with the exception that a 45% optimised bypass of biomass to the CHP plant (Area 500, Figure 26) was required to ensure sufficient steam production for the adjacent sugar mill but also meet the requirements of the biorefinery itself.

Therefore, 62.4 tons/h of sugarcane bagasse and trash is sent through to pretreatment (Area 100). There the lignocellulose is subjected to hydrolysis in a DA reactor (R-101) to liberate sugars for the hemicellulosic portion. An acid concentration of 2.9 g H<sub>2</sub>SO<sub>4</sub>/100 ml solution, temperature of 130 °C and a solids to liquid ratio of 0.25 (1:4) is selected and the contents are held in the reactor for 30 min (Moutta, et al., 2011). The reactions can be viewed in the previously listed Table 16. Two portions exit the reactor, a soluble sugar hydrolysate (rich in xylose) and an insoluble cellulose and lignin portion. The soluble hydrolysate containing remnant DA is separated using solid-liquid filtration (S-101) and sent to the lime tank (R-102) where the acid is neutralised with the aid of quick lime (CaO) in 15 % excess of the stoichiometric requirement. The neutralisation process produces heat from the exothermic reaction and is jacket cooled (E-103) with utility water (Görgens, et al., 2015). A 15 min reaction time is assumed according to industrial practices utilizing CaO as a base (Lewis & Boyton, 2000), as previously described for the xylitol biorefinery scenario. The difference from the xylitol scenario is that glucose and not xylose is utilised as the substrate for fermentation and therefore the cellulose portion of lignocellulose needs to be broken down into individual glucose monomer units in order to feed the microorganisms used during fermentation (Area 300). Therefore, the solid portion (stream 23) that was separated from the sugar hydrolysate (stream 14), must undergo further treatment before being sent to the next section of the plant, namely enzymatic hydrolysis (Area 200). The solid portion is washed in a wash water tank (represented collectively as VT-105 and S-103) to recover any residual sugars that may be contained within the moist solid cellulignin. The washed solids (stream 24) is then sent to enzymatic hydrolysis (Area 200).





Equipment List					
Equipment ID	Description	Quantity	Equipment ID	Description	Quantity
CONV-101	Bagasse and Trash Conveyor	2	P-105 A/B	Wash Water Feed Pump	2
CONV-102	Bypassed Biomass Conveyor	2	P-106 A/B	Neutralised Hydrolysate Pump	2
CONV-103	Lime Conveyor	2	P-107 A/B	Recovered Sugar Pump	2
CONV-105	Gypsum Conveyor	2	P-108 A/B	WWT Transfer Pump	2
CONV-201	Cellulignin Conveyor	2	R-101	Dilute Acid Tank	1
E-101	Dilute Acid Tank Heating Jacket	1	R-102	Lime Neutralisation Tank	1
E-102	Hydrolysate Cooler	1	S-101	Cellulignin Filter Press	15
E-103	Lime Tank Cooling Jacket	1	S-102	Gypsum Filter Press	29
HOP-101	Bagasse and Trash Hopper	1	S-103	Wash Water Tank Filter	3
HOP-102	Lime Hopper	1	VT-101	Bagasse and Trash Holding Tank	1
MIX-101	Dilute Acid Mixer	1	VT-102	Sulphuric Acid Storage Tank	1
P-101 A/B	Hydrolysate Cooler	2	VT-103	Dilution Water Storage Tank	1
P-102 A/B	Dilute Acid Hydrolysate Pump	2	VT-104	Wash Water Storage Tank	1
P-102A/B	Dilute Acid Feed Pump	2	VT-105	Wash Water Filtration Tank	3
P-103 A/B	Dilute Acid Hydrolysate Pump	2	VT-106	Recovered Sugar Holding Tank	2
P-104 A/B	Gypsum Filter Pump	2	P-301 A/B	Detoxified Glucose Rich Feed Pump	2

Figure 22 - CA Pretreatment Section (Area 100)

#### 4.5.3.2 Enzymatic Hydrolysis (Area 200)

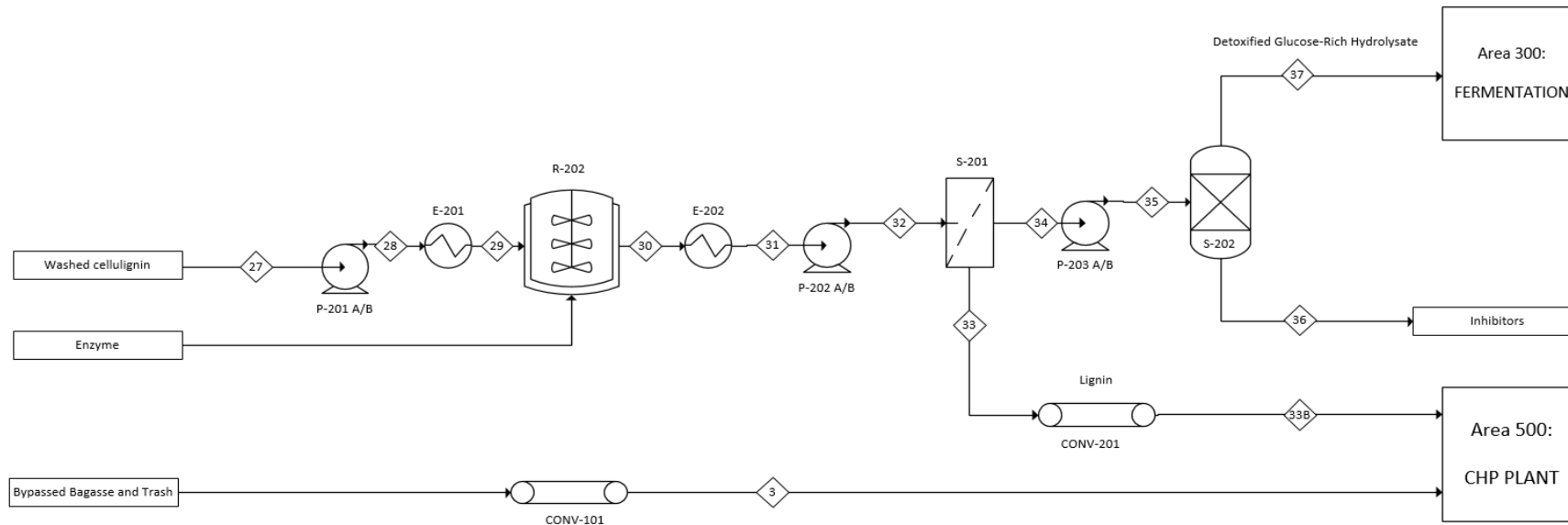
Figure 23 shows the enzymatic hydrolysis section of the CA biorefinery. During enzymatic hydrolysis, cellulose polymer chains are hydrolysed in the presence of cellulase enzymes inside a reactor to liberate monomeric glucose (R-201). The reactor operates at a temperature of 48 °C for 84h in a two-stage process (24 h continuous and 60 h batch) under atmospheric pressures, where 20 mg cellulase/g cellulose is dosed to achieve the conversions listed in Table 21 (Humbird, et al., 2011).

Table 21 - Enzymatic Hydrolysis Reactions and Conversions (Humbird, et al., 2011)

Reactant	Reaction	Product	Conversion (%)
Cellulose	$(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_6H_{12}O_6$	Glucose	90
Cellulose	$2(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_{12}H_{22}O_{12}$	Cellobiose	1.2
Cellobiose	$C_{12}H_{22}O_{12} + H_2O \rightarrow 2C_6H_{12}O_6$	Glucose	100

After enzymatic hydrolysis the resulting glucose rich hydrolysate (stream 32) is transported to a set of filter presses (S-201) that separate the solid lignin portion (stream 33) from the solubilised sugars (stream 34). The lignin is transported via conveyor (CONV-201) to the CHP plant (Area 500) for electricity and steam production, whereas the glucose-rich sugars are sent through to carbon adsorption (S-202) to remove soluble inhibitors, as well as any soluble ions released during the pretreatment steps, before being sent to the fermentation section (Area 300).

## Area 200: Enzymatic Hydrolysis



Equipment List		
Equipment ID	Description	Quantity
CONV-101	Bagasse and Trash Conveyor	2
CONV-201	Bypassed Biomass Conveyor	2
E-201	Cellulignin Slurry Cooler	1
E-202	Enzymatic Cooling Jacket	1
P-201 A/B	Enzymatic Tank Feed Pump	2
P-202 A/B	Enzymatic Hydrolysate Pump	2
P-203 A/B	Activated Carbon Column Feed Pump	2
R-202	Enzymatic Hydrolysis Tank (2-stage)	1
S-201	Lignin Solids Filter	2
S-202	Activated carbon column	2

Figure 23 - CA Enzymatic Hydrolysis Section (Area 200)

#### 4.5.3.3 Citric Acid Fermentation (Area 300)

After glucose has been liberated from cellulose during enzymatic hydrolysis and purified in activated carbon, the glucose rich stream is combined with nutrients and makeup water in a mixing tank (VT-301) to bring the glucose concentration to 336 g/L (Anastassiadis & Rehm, 2006). The composition of the resulting feed medium mixture is given in Table 22.

Table 22 - CA fermentation medium sugar and nutrient concentration, table adapted from (Anastassiadis & Rehm, 2006)

Component	Concentration in medium (g/l)
Glucose	336
NH <sub>4</sub> Cl	4.5
KH <sub>2</sub> PO <sub>4</sub>	1.05
MgSO <sub>4</sub> ·7H <sub>2</sub> O	0.525
MnSO <sub>4</sub> ·4H <sub>2</sub> O	0.2507
FeSO <sub>4</sub> ·7H <sub>2</sub> O	0
CuSO <sub>4</sub> ·5H <sub>2</sub> O	0.00015
ZnSO <sub>4</sub> ·7H <sub>2</sub> O	0.0315
CoSO <sub>4</sub> ·7H <sub>2</sub> O	0.006
H <sub>3</sub> BO <sub>3</sub>	0.06
CaCl <sub>2</sub>	0.15
NaCl	0.15
KJ	0.15
CA	2.5
NA <sub>2</sub> MoO <sub>4</sub> ·2H <sub>2</sub> O	0.0003
Thiamine	0.003
Biotin	0.000375
Pyridoxine	0.0009375
Ca-D-Pantothenate	0.0009375
Nicotinic acid (niacin)	0.00075

The combined glucose and nutrient feed is then sterilized for 30 min at 121°C in a sterilization unit (represented by a combination of E-301A and E-301B) before being mixed with the yeast inoculum from the seed train section (Area 700) of the biorefinery (Figure 28) in a second mixing tank (VT-302). The sterilized feed with the inoculum is thereafter sent to 3 fermenters (R-301) operating at 30 °C at a pH < 2, where yeast converts glucose to citric acid. The main stoichiometric reactions are given in Table 23, where oxalic acid is assumed to be the main by-product formed alongside CA (Heinzle, et al., 2006). It is assumed that 98% of the organic acids present in the product stream are comprised of CA and the remainder being mainly oxalic and negligible quantities of gluconic acid (Datta & Bergemann, 1996).

Table 23 - CA Fermenter Stoichiometric Reactions (Heinzle, et al., 2006)

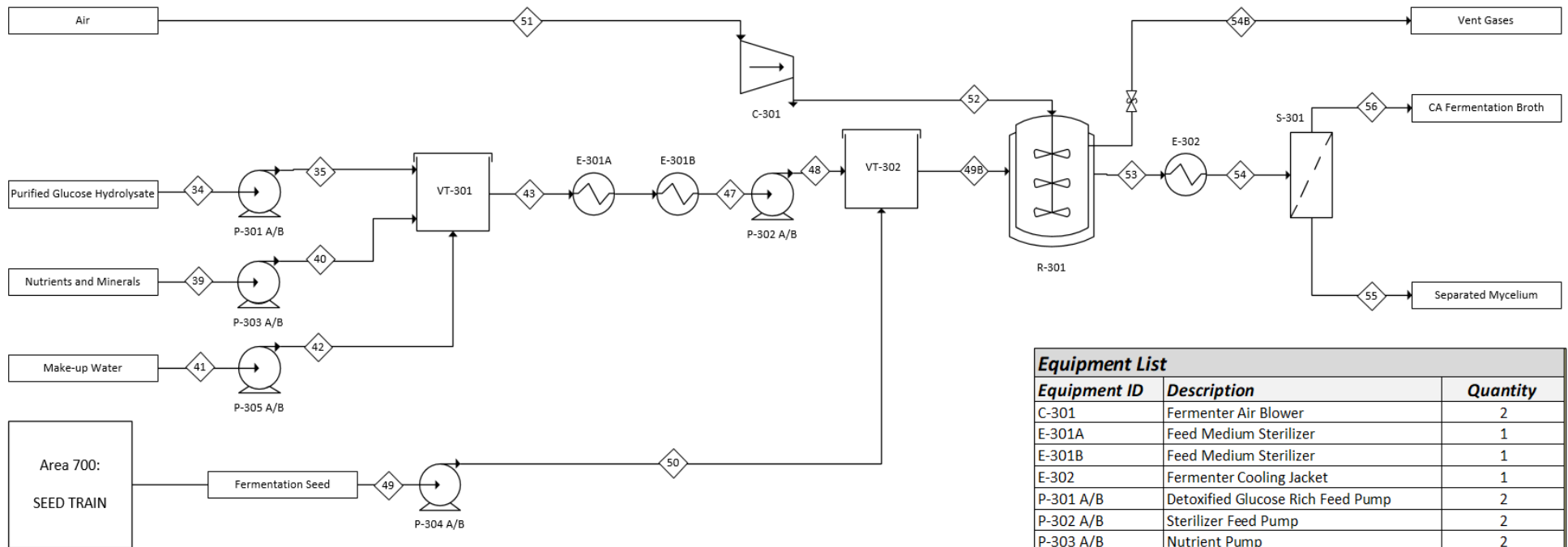
Reactant	Reaction	Product	Conversion (%)
Glucose	$C_6H_{12}O_6 + 1.5 O_2 \rightarrow C_6H_8O_7 + 2H_2O$	Citric Acid	90
Glucose	$2C_6H_{12}O_6 + 9 O_2 \rightarrow 6C_2H_2O_4 + 6H_2O$	Oxalic Acid	2
Glucose	$3C_6H_{12}O_6 + 8 O_2 + 2NH_3 \rightarrow 2C_5H_7NO_2 + 14H_2O + 8CO_2$	Biomass	0.8

In addition it is assumed that 15% of the CA produced remain intracellular within the mycelia that form (Lengon, et al., 1999). This can be press filtered (not shown) and dried to recover and form a cake containing not more than 0.2 % of CA (Lengon, et al., 1999). The exiting reactor stream (stream 53) is thereafter filtered pressed (S-301) to recover part of the CA before being sent through to the purification section (Area 400) of the CA plant.

#### 4.5.3.4 Purification (Area 400)

Figure 25 shows the separation steps used to purify the CA, namely solvent extraction (SE), crystallisation and drying (Area 400). Once the cell mass has been separated from the outlet reactor broth (stream 56), impurities as a result of side-reactions in the fermenter need to be removed. As previously mentioned, the main impurity assumed to form part of the outlet fermentation broth is oxalic acid. SE takes place using 2 absorption columns (S-402). One column utilises an amine solvent consisting of a mixture of 5% (wt/v) n-octanol, 47% (wt/v) Alamine<sup>®</sup> 336 and 50% (wt/v) Isopar K and the other stripping water at 40 °C (Baniel, et al., 2003; Grewal & Kalra, 1995). SE eliminates the need for additional lime, sulphuric acid and wash water. Furthermore the solvent is regenerated by stripping with water, allowing the costly amine solvent to be recovered and recycled back into the process (refer to Figure 25).

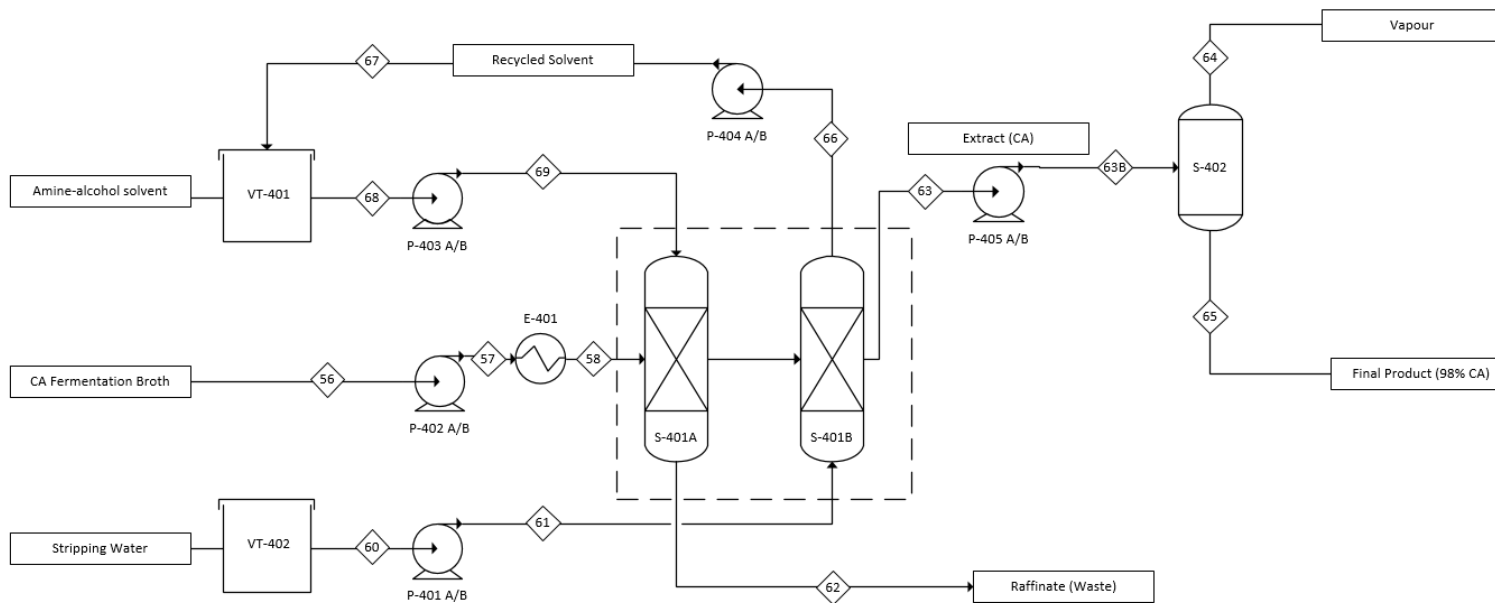
## Area 300: CA Fermentation



Equipment List		
Equipment ID	Description	Quantity
C-301	Fermenter Air Blower	2
E-301A	Feed Medium Sterilizer	1
E-301B	Feed Medium Sterilizer	1
E-302	Fermenter Cooling Jacket	1
P-301 A/B	Detoxified Glucose Rich Feed Pump	2
P-302 A/B	Sterilizer Feed Pump	2
P-303 A/B	Nutrient Pump	2
P-304 A/B	Fermenter Seed Feed Pump	2
P-305 A/B	Makeup Water Pump	2
P-306 A/B	Fermentation Product Pump	2
R-301	Citric Acid Fermenter	3
S-301	Mycelium Filter Press	1
VT-301	Fermenter Feed Medium Holding Tank 1	1
VT-302	Fermenter Feed Medium Holding Tank 2	1

Figure 24 - CA Fermentation Section (Area 300)

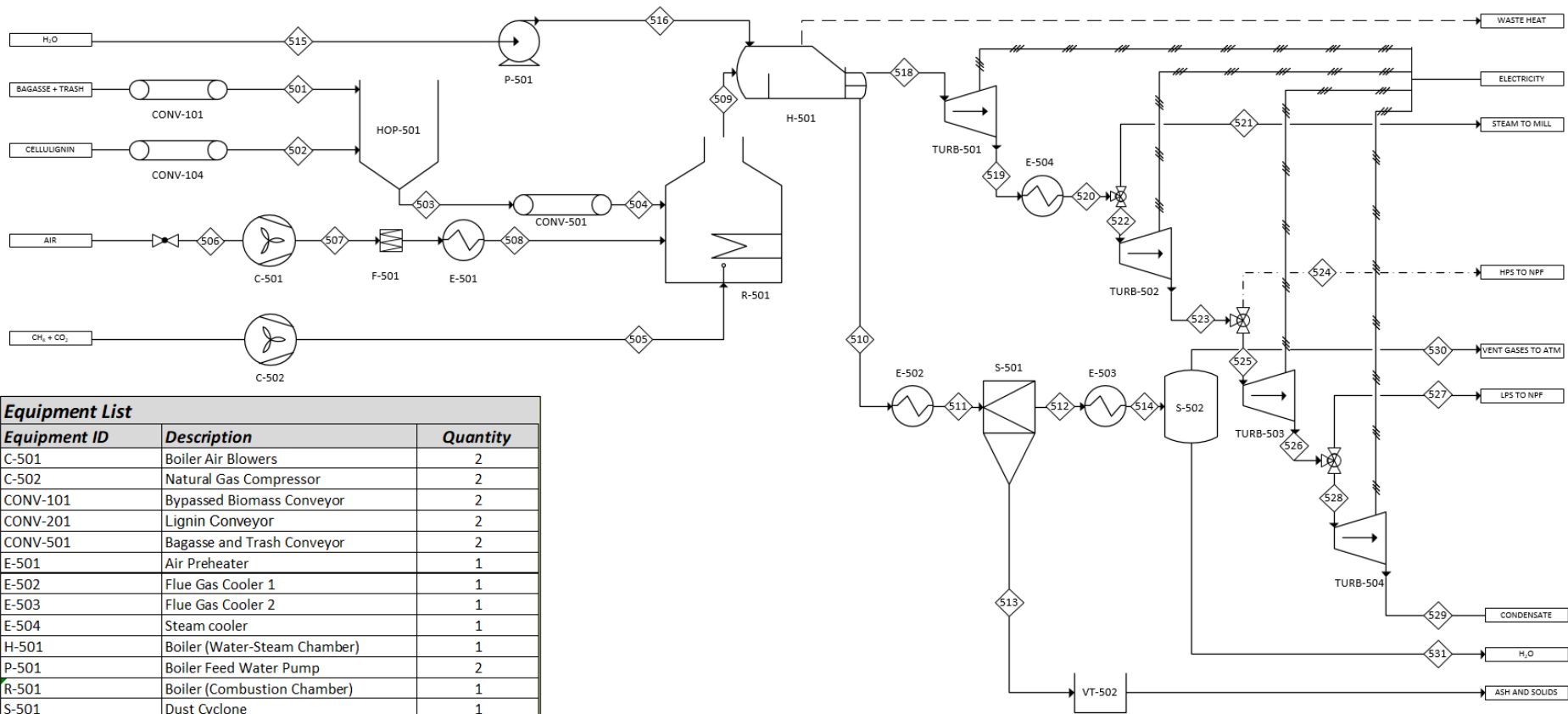
## Area 400: Purification



Equipment List		
Equipment ID	Description	Quantity
E-401	Fermentation Broth Cooler	1
P-401 A/B	Stripping Water Feed Pump	2
P-402 A/B	Fermentation Broth Pump	2
P-403 A/B	Amine-solvent Pump	2
P-404 A/B	Recycled Solvent Pump	2
P-405 A/B	Crystalliser Feed Pump	2
S-401A	Amine-water extraction column	1
S-401B	Amine-water stripping column	1
S-402	Evaporator-Crystalliser	1
VT-401	Amine-solvent Holding Tank	1
VT-402	Stripping Water	1

Figure 25 - CA Purification Section (Area 40)

## AREA 500: COMBINED HEAT AND POWER PLANT

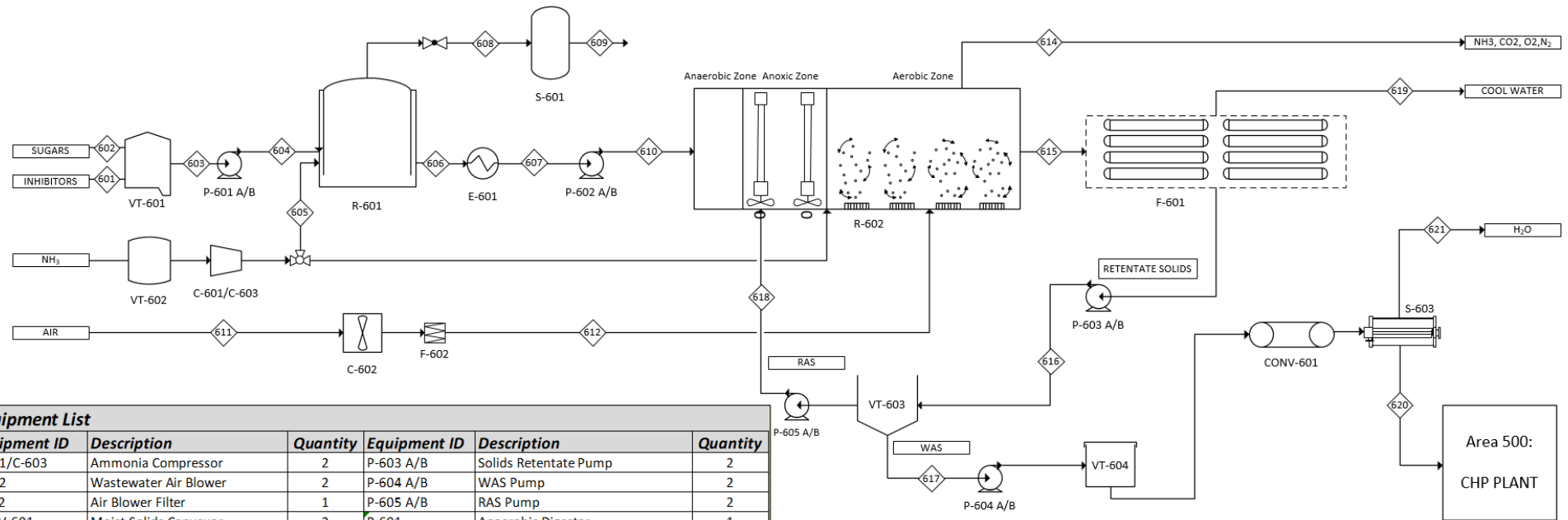


Equipment List		
Equipment ID	Description	Quantity
C-501	Boiler Air Blowers	2
C-502	Natural Gas Compressor	2
CONV-101	Bypassed Biomass Conveyor	2
CONV-201	Lignin Conveyor	2
CONV-501	Bagasse and Trash Conveyor	2
E-501	Air Preheater	1
E-502	Flue Gas Cooler 1	1
E-503	Flue Gas Cooler 2	1
E-504	Steam cooler	1
H-501	Boiler (Water-Steam Chamber)	1
P-501	Boiler Feed Water Pump	2
R-501	Boiler (Combustion Chamber)	1
S-501	Dust Cyclone	1
S-502	Gas Vent with Bag Filter	1
TURB-501	Condensing Steam Turbine (2-stage)	1
TURB-502	Condensing Steam Turbine (2-stage)	0
TURB-503	Condensing Steam Turbine (2-stage)	0
TURB-504	Condensing Steam Turbine (2-stage)	0
HOP-501	Bypassed Biomass Hopper	1
VT-502	Ash and Dust Skip Bin	1

Figure 26 - CA CHP Plant Section (Area 500)



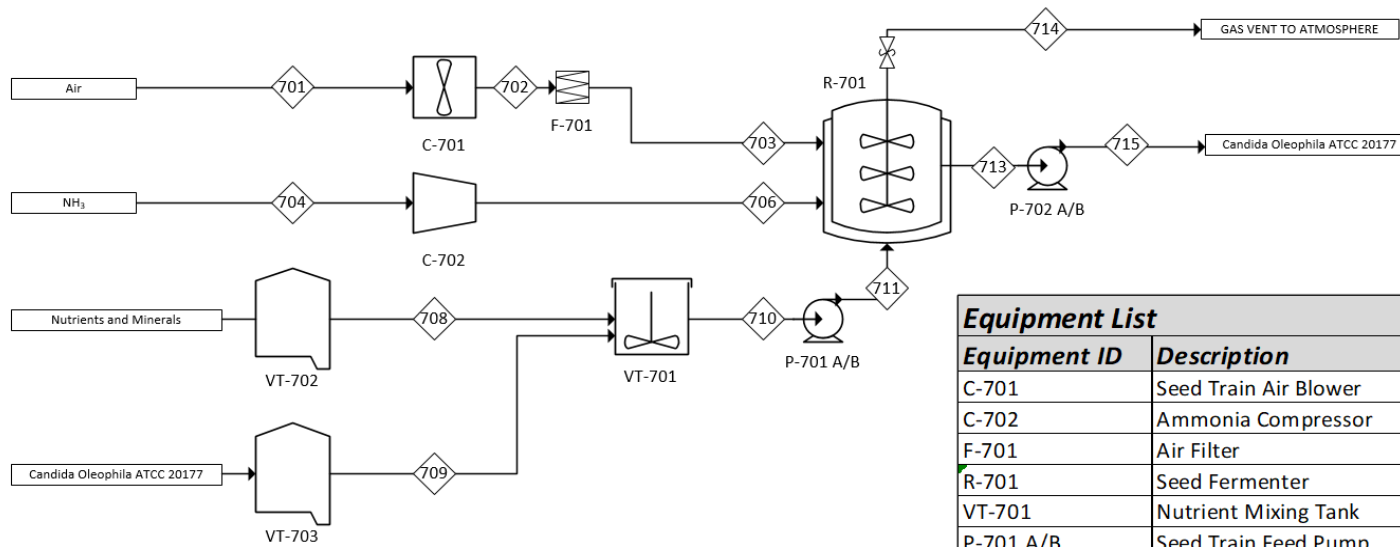
**AREA 600: WASTEWATER TREATMENT PLANT**



Equipment List					
Equipment ID	Description	Quantity	Equipment ID	Description	Quantity
C-601/C-603	Ammonia Compressor	2	P-603 A/B	Solids Retentate Pump	2
C-602	Wastewater Air Blower	2	P-604 A/B	WAS Pump	2
F-602	Air Blower Filter	1	P-605 A/B	RAS Pump	2
CONV-601	Moist Solids Conveyor	2	R-601	Anaerobic Digester	1
R-602	Activated Sludge Reactor	1	S-601	Anaerobic Gas Scrubber	1
F-601	MBR Membrane Module	1	VT-601	Wastewater Feed Holding Tank	1
S-603	Solids Beltpress	1	VT-602	Ammonia Gas Cylinder	1
E-601	Anaerobic Cooler	1	VT-603	Activated Sludge Holding Tank	1
P-601 A/B	Waste water Feed Pump	2	VT-604	Activated Sludge Holding Tank	1

Figure 27 - CA WWT Plant Section (Area 600)

## AREA 700: FERMENTER SEED TRAIN



<b>Equipment List</b>		
<b>Equipment ID</b>	<b>Description</b>	<b>Quantity</b>
C-701	Seed Train Air Blower	2
C-702	Ammonia Compressor	2
F-701	Air Filter	1
R-701	Seed Fermenter	1
VT-701	Nutrient Mixing Tank	1
P-701 A/B	Seed Train Feed Pump	2
P-702 A/B	Seed Train Effluent Pump	2
VT-702	Nutrient Holding Tank	1
VT-703	Seed Inoculum Tank	1

Figure 28 - CA Seed Train Section (Area 700)

#### 4.5.4 Glutamic Acid (GA) and Electricity Co-production Scenario

Figure 29, Figure 30 and Figure 31 show the PFD for the pretreatment (Area 100), enzymatic hydrolysis (Area 200) and fermentation (Area 300) sections of the GA biorefinery for a 35% biomass bypass to the CHP plant (Area 500). The pretreatment (Figure 29) and enzymatic hydrolysis (Figure 30) steps are similar to those described for the CA and xylitol biorefinery scenarios and are for all purposes equivalent with regards to chemical reactions. Consequently only the GA fermentation (area 300) section of the plant will be described. Aspen process flow diagrams (Figure 72 - Figure 79) for all areas of the GA scenario are provided in Appendix C2.

##### 4.5.4.1 Glutamic Acid Fermentation (Area 300)

The slurry exiting the enzyme reactor (stream 29, Figure 30), is filtered through a rotary drum filter (S-201), forming a solid fraction (stream 31), containing mostly insoluble lignin and un-hydrolysed cellulose and a sugar, rich in glucose (stream 30). The latter is then sent to an activated carbon column (S-202) to remove soluble inhibitors (stream 33). The purified glucose (stream 34) is pumped to a mixing tank (VT-301), where the glucose medium is mixed with micronutrients and salts (stream 38) before passing through a feed-medium sterilizer (E-301) that heats the mixture to 120 °C for 30 min and a cooler (E-303) that subsequently cools the glucose-rich medium to a temperature between 32-33 °C (stream 41) to prepare it for fermentation (Figure 31) (Miesher & Haute, 1975). The glucose-rich medium (stream 41A) along with the makeup water (stream 36) is pumped to a second mixing tank (VT-302) such that the appropriate concentration of nutrients and glucose is maintained (see Table 24) (Miesher & Haute, 1975) (refer to Figure 31).

Table 24 – GA Fermentation Feed Medium Concentrations (Miesher & Haute, 1975)

Component	Concentration (g/l)
Glucose	123
KH <sub>2</sub> PO <sub>4</sub>	1
K <sub>2</sub> SO <sub>4</sub>	1
MgSO <sub>4</sub>	1
FeSO <sub>4</sub> (mg/l)	6
MNSO <sub>4</sub> (mg/l)	6
H <sub>2</sub> O (ml)	4860
Antifoam (Polypropylene glycol)	0.1
NH <sub>4</sub> NO <sub>3</sub>	5
CSL	6
H <sub>2</sub> O (ml)	5000
Oleic acid (ml)	61

Seed inoculum (stream 43), from the seed fermenter trains (R-701) (Area 700), containing the patented *Brevibacterium divaricatum* NRRL B-2311, along with filtered air from C-301 (stream 45),

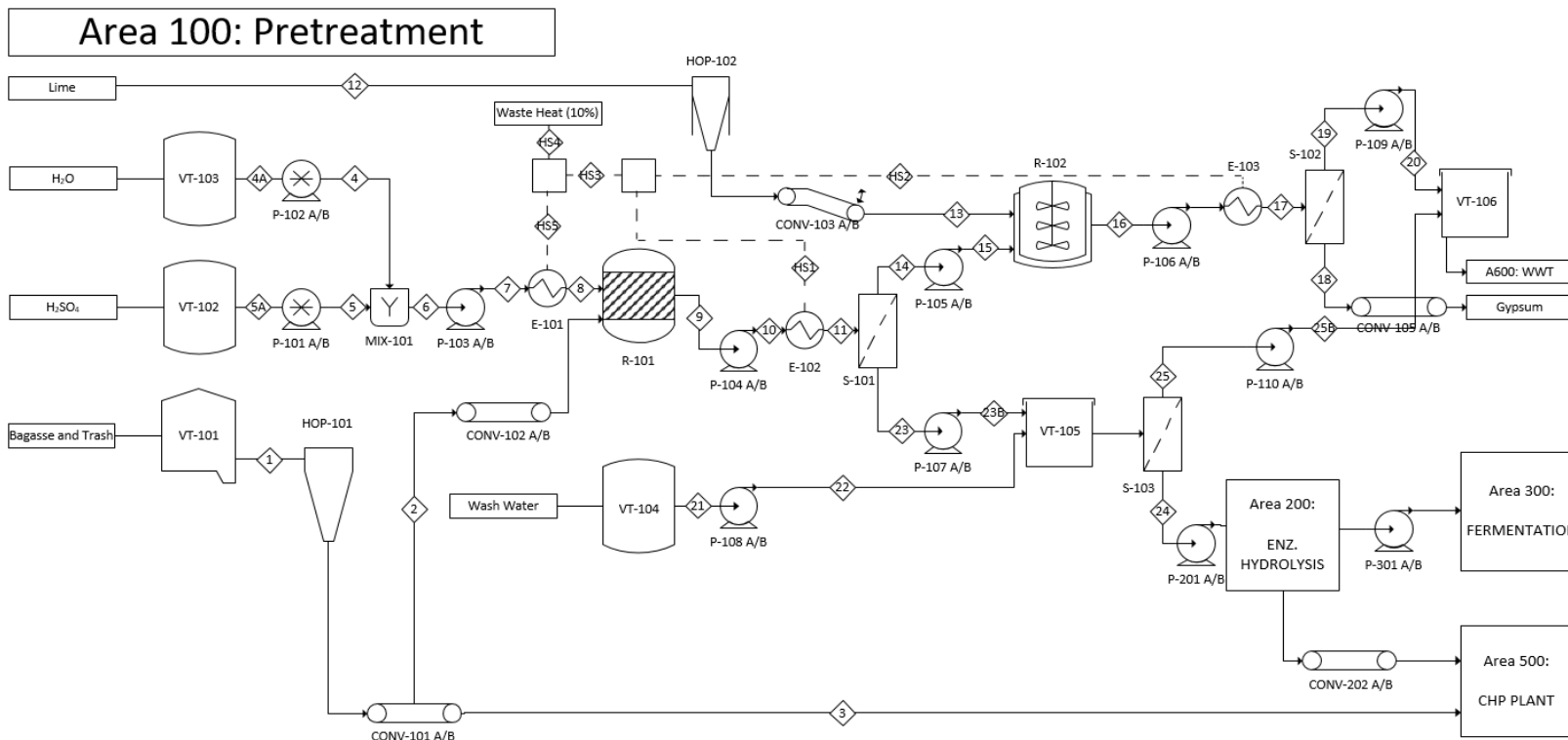
glucose medium from VT-302 and ammonia (stream 47) are continuously fed to a train of 8 stirred fermenters in parallel (R-301). The feed is held at atmospheric conditions in the reactors for a residence time of 28.5 h at a temperature of 33 °C resulting in a GA concentration of 100 g/l with only 2% of glucose from the feed after fermentation (stream 52) (Miesher & Haute, 1975). pH levels are controlled at 6.5 by continuously adding fresh feed medium and separating product within a combined integrated micro and nanofiltration membrane system c. The addition of 2-3 µg/l biotin is also included (cost included, not shown) to provide suboptimal growth and ensure extracellular production of GA and prevent by-product formation (Kumar, et al., 2014; Lederberg, 2000) (Figure 31). Stoichiometric reactions are provided in Table 25.

Table 25 - GA Chemical Reactions (Pandey, et al., 2015; Delaunay, et al., 2002; Lederberg, 2000)

Reactant	Reaction	Product	Conversion (%)
Glucose	$C_6H_{12}O_6 + NH_3 + O_2 \rightarrow C_5H_9NO_4 + CO_2 + 3H_2O$	Glutamic Acid	87
Glucose	$3C_6H_{12}O_6 + 8O_2 + 2NH_3 \rightarrow 2C_5H_7NO_2$	Biomass	0.09

After fermentation, the cell protein (stream 53) is removed from the resulting broth exiting the fermenter (stream 52) by using PVDF microfiltration (MF) spiral wound modules (F-401) operating at 2.5 bar pressure with a permeate flux and cross flow velocity of 350 L/(m<sup>2</sup>h) and 1.2m/s respectively (Pal, et al., 2016; 2015). This produces a diluted permeate of 64.5 g/l GA (Pal, et al., 2016; 2015). Part of the retentate solid cells are recycled back to the fermenter to maintain high yields and productivity. The micro filtered broth (stream 57) is then fed to a train of 2 nanofiltration (NF) modules (F-402 and F-403) to firstly remove mineral ions and recover any residual glucose and rejected GA (stream 60) (F-402) and secondly to concentrate GA (F-403, where GA reports to the retentate, stream 66) (Pal, et al., 2015); (Pal, et al., 2016)). 15 % of GA is rejected in F-402 (reports to stream 60) and 94% in F-403 (6% GA permeate, stream 61), which are sent to a holding tank (VT-401) along with the residual glucose and minerals (Pal, et al., 2016; 2015). The trains are operated at a transmembrane (TMB) pressure of 15 bar (max operating pressure 40 bar) resulting in a GA rich retentate with a GA concentration of 175 g/l leaving F-403 (stream 66) (Pal, et al., 2016; 2015). The resulting GA-rich retentate (stream 66) is thereafter sent to an evaporation tank (S-401) and vacuum crystalliser (S-402) in order to liberate the solids and remove any remaining moisture (stream 68 and stream 72). The dried product (stream 73) is then packaged resulting in a 98% pure GA product ready for shipment to consumers (refer to Figure 31).

Areas 500 (CHP), 600 (WWT) and 700 (Seed Trains) have previously been described for the CA fermentative scenario and are therefore not detailed further for GA as they are similar to this scenario.



Equipment List					
Equipment ID	Description	Quantity	Equipment ID	Description	Quantity
CONV-101 A/B	Bagasse and Trash Conveyor	2	P-107 A/B	Cellulignin slurry pump	2
CONV-102 A/B	Bypassed Biomass Conveyor	2	P-108 A/B	Wash Water Feed Pump	2
CONV-103 A/B	Lime Conveyor	2	P-109 A/B	Filtered Hydrolysate Pump	2
CONV-105 A/B	Gypsum Conveyor	2	P-110 A/B	Recovered Sugar Pump	2
CONV-202 A/B	Lignin Conveyor	2	P-201 A/B	Wash Cellulignin Feed Pump	2
E-101	DA Preheater	1	R-101	Dilute Acid Tank Reactor	1
E-102	DA Hydrolysate Cooler	1	R-102	Lime Neutralising Tank	1
E-103	Lime Tank Cooling Jacket	1	S-101	Cellulignin-Sugars Filter Press	20
HOP-101	Bagasse and Trash Hopper	1	S-102	Gypsum Rotary Filter Drum	17
HOP-102	Lime Hopper	1	S-103	Lignin Filter Press	2
MIX-101	DA Mixer	1	VT-101	Bagasse and Trash Storage Tank	1
P-101 A/B	Dilute Acid Feed Pump	2	VT-102	DA Storage Tank	1
P-102 A/B	Dilute Acid Hydrolysate Pump	2	VT-103	DA Water Storage Tank	1
P-103 A/B	Gypsum Filter Pump	2	VT-104	Wash Water Storage Tank	1
P-104 A/B	Wash Water Feed Pump	2	VT-105	Wash Water Tank	3
P-105 A/B	Lime Tank Feed Pump	2	VT-106	Recovered Sugar Holding Tank	1
P-106 A/B	Neutralised Hydrolysate Pump	2	P-301 A/B	Detoxified Glucose Rich Feed Pump	2

Figure 29 - GA Pretreatment Section (Area 100)

## Area 200: Enzymatic Hydrolysis

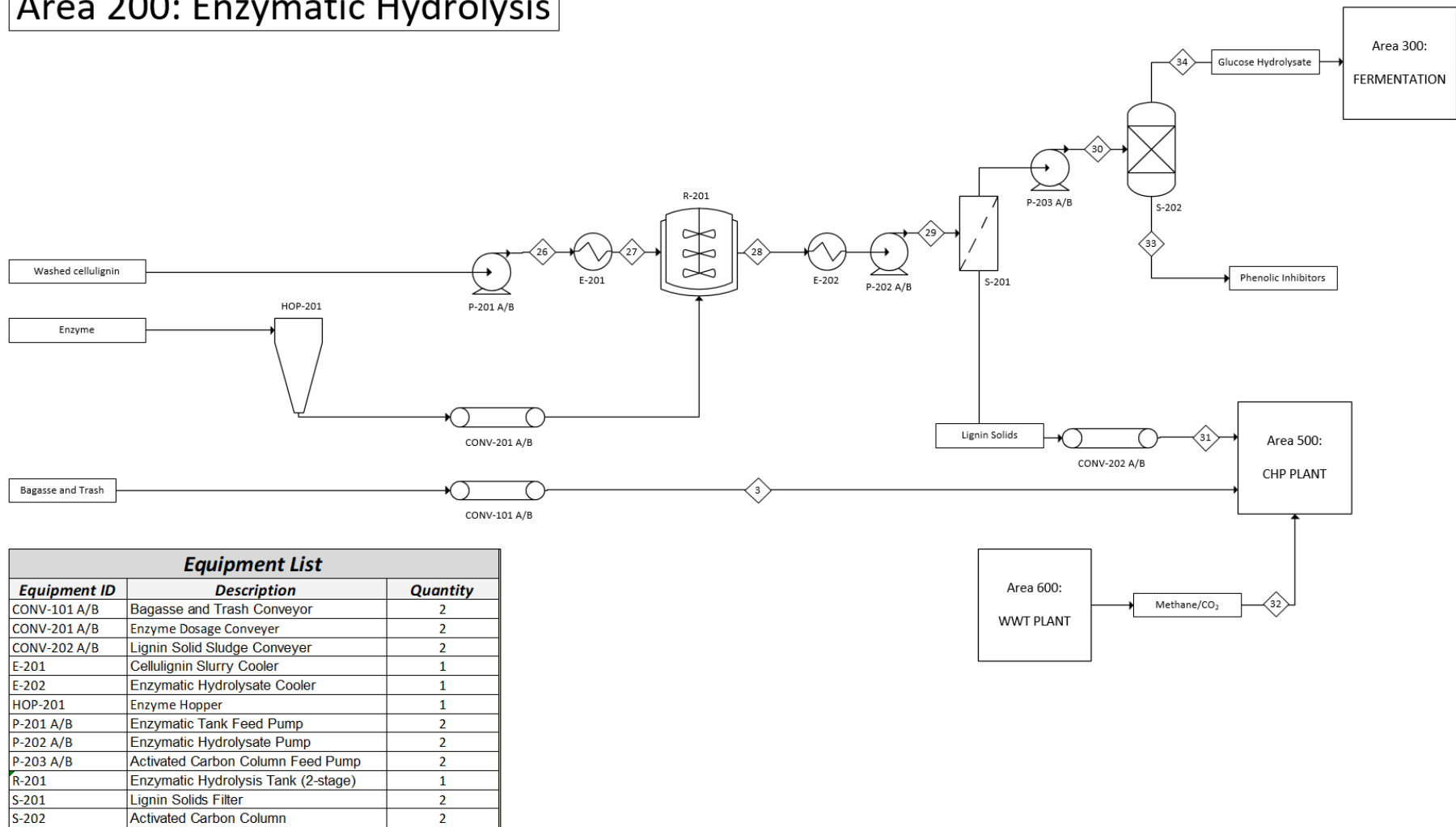
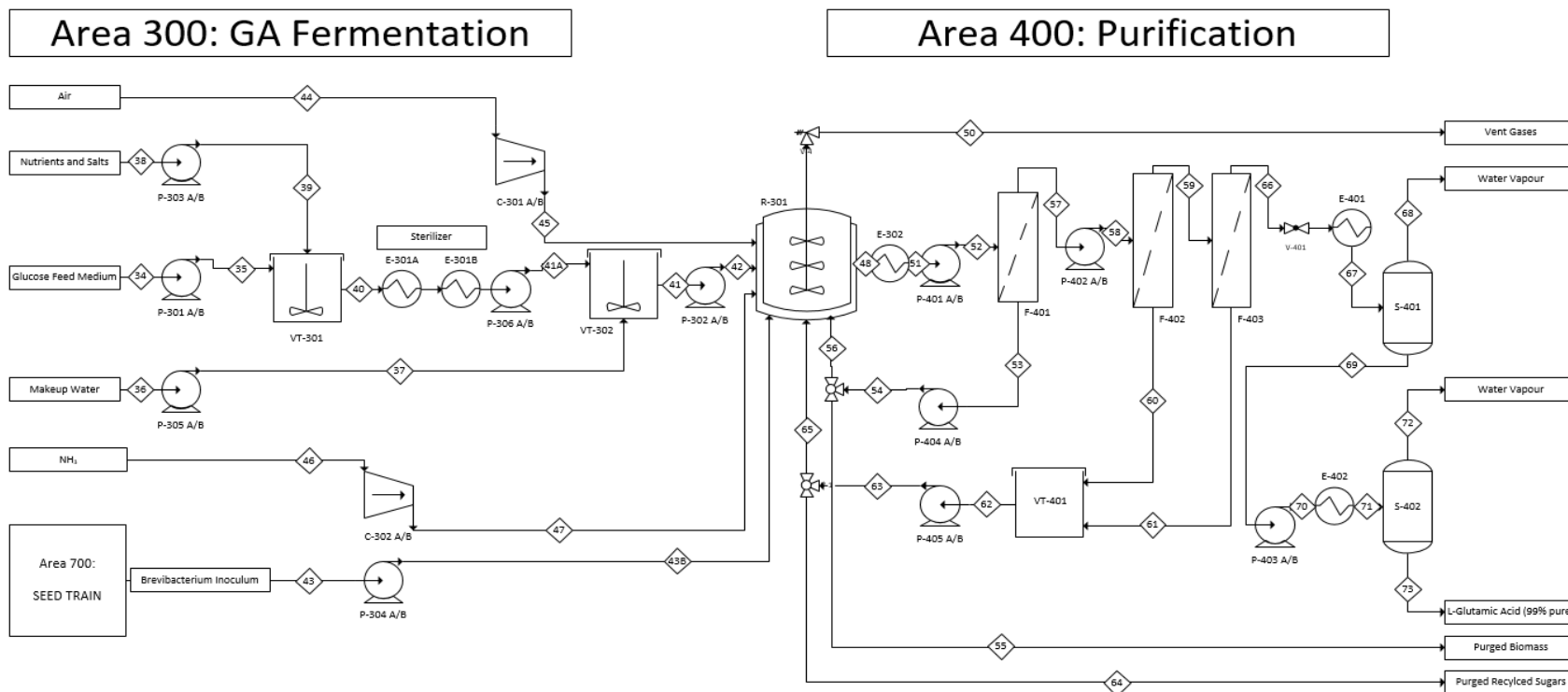


Figure 30 - GA Enzymatic Hydrolysis Section (Area 200)



Equipment List					
Equipment ID	Description	Quantity	Equipment ID	Description	Quantity
C-301 A/B	Air Blower	2	P-305 A/B	Makeup Water Pump	2
C-302 A/B	Ammonia Compressor	2	P-306 A/B	Fermenter Feed Medium Holding Tank 2	2
E-301A	Feed Medium Sterilizer	1	P-401 A/B	Microfilter Feed Pump	2
E-301B	Feed Medium Sterilizer	1	P-402 A/B	Nanofiltration Feed Pump 1	2
E-302	Fermenter Cooling Jacket	1	P-403 A/B	GA Crystalliser Feed Pump	2
E-401	Evaporation Column Preheater	1	P-404 A/B	Biomass Recycle Pump	2
E-402	Precrystallisation Cooler	1	P-405 A/B	Sugar Recycle Pump	2
F-401	Microfilter Train	1	R-301	Glutamic Acid Fermenter	8
F-402	Ultrafiltration Train NF20	1	S-401	Evaporation Column	1
F-403	Ultrafiltration Train NF1	1	S-402	Vacuum Crystalliser	1
P-301 A/B	Detoxified Glucose Rich Feed Pump	2	VT-301	Fermenter Feed Medium Holding Tank 1	1
P-302 A/B	Sterilizer Feed Pump	2	VT-302	Fermenter Feed Medium Holding Tank 2	1
P-303 A/B	Nutrient Pump	2	VT-401	Sugar Recycle Holding Tank	1
P-304 A/B	Fermenter Seed Feed Pump	2			

Figure 31 - GA Fermentation and Purification Sections (Area 300 and 400)

## 4.6 Economic Evaluation

### 4.6.1 Capital Cost Estimation

The equipment purchase costs ( $C_e$ ) are based on a 2007 US Gulf Coast basis and are calculated using Equation 6 (Towler & Sinnott, 2012). The sizing constants, (a) and (b), as well as the sizing exponent (n) are specific to the unique category of equipment and are detailed in Appendix H (see Table 85), according to the equipment category and size factor (S) range, which takes on different units for different equipment (Towler & Sinnott, 2012).

$$C_e = a + bS^n \quad [\text{Eq. 6}]$$

Equation 7 is used to determine the new equipment cost ( $C_2$ ) from the base or reference equipment cost ( $C_1$ ), the size scaling factor (n), the new sizing parameter ( $S_2$ ) and the base sizing parameter ( $S_1$ ) which may be found in another literature source or industry vendor for which Equation 6 may not be suitable. A list of scaling exponents for various equipment types are listed in Table 26 (Humbird, et al., 2011).

Table 26 - Equipment Capacity Scaling Components (Humbird, et al., 2011)

Scaling Exponents, $C_{\text{new}} = C_{\text{base}}(S_{\text{new}}/S_{\text{base}})^n$	
Equipment	Exponent
Agitators	0.5
Compressors, motor driven	0.6
Distillation columns	0.6
Heat exchangers	0.7
Inline mixers	0.5
Skidded equipment	0.6
Pressure Vessels	0.7
Pumps	0.8
Tanks, atmospheric	0.7
Solids handling equipment	0.8

$$C_2 = C_1 \times \left(\frac{S_2}{S_1}\right)^n \quad [\text{Eq. 7}]$$

Once the equipment purchase cost is determined, it needs to be multiplied by the appropriate installation factor to account for the various added costs associated with installation. Installation factors to be used during this study can be viewed in Table 27 ( (Humbird, et al., 2011); (Görgens, et al., 2015)).

$$C_{\text{installed}} = C_e \times \text{installation factor} \quad [\text{Eq. 8}]$$



Table 27 - Equipment Installation Factors (Humbird, et al., 2011)

Installation Factors	
Equipment	Multiplier
Agitators, carbon steel	1.6
Agitators, stainless steel	1.5
Boiler	1.8
Compressors, motor driven	1.6
Cooling Tower	1.5
Distillation columns, stainless steel	2.4
Heat exchangers, shell & tube, SS	2.2
Heat exchangers, plate & frame, SS	1.8
Heat exchangers, plate & frame, air cooled	2.8
Inline mixers	1
Skidded equipment	1.8
Solids handling equipment (incl. filters)	1.7
Pressure vessels, carbon steel	3.1
Pressure vessels, SS	2
Pretreatment reactor system	1.5
Pumps, SS	2.3
Pumps, carbon steel	3.1
Tanks, field-erected, carbon steel	1.7
Tanks, field-erected, SS	1.5
Tanks, storage, plastic	3
Tanks, storage, carbon steel	2.6
Tanks, storage, SS	1.8
Turbo-generator	1.8

Purchase costs calculated from Equation 6 are on a 2007 year basis, whereas costs determined from Equation 7 could be calculated for different years and are based purely on sizing parameters. In order to account for inflation of equipment costs over time, Equation 9 can be used to determine the most up-to-date cost ( $C_{new}$ ) from the older cost ( $C_{old}$ ) by multiplying it with the most recent Chemical Engineering Price Cost Index (CEPCI) value and dividing by the old CEPCI value (Towler & Sinnott, 2012).

$$C_{new} = C_{base} \times \frac{CEPCI_{new}}{CEPCI_{base}} \quad [\text{Eq. 9}]$$

Table 28 contains CEPCI values for the relevant years used to adjust price.

Table 28 - CEPCI Values for Different Years

Cost Escalation CEPCI Values	
Year	CE Index
1998	390
2007	525
2009	522
2010	551
2016	536

Once installed equipment costs have been calculated, the inside battery limit (ISBL) costs can be determined by summing all plant section costs associated with manufacturing the product (does not including the CHP plant, storage (5% of ISBL), utility costs (6.5%)) Together with warehousing (4% of ISBL), Site development (9% ISBL) and additional piping costs (4.5%), one can determine the total direct costs (TDC). Total indirect costs (TIDC) include prorated expenses (10% TDC), field expenses (10%), office construction (20%), contingency (10%) and other costs (10%). Fixed capital investment (FCI) can then be calculated by adding TDC and TIDCs. 5% working capital added to this provides us with total capital investment.

#### 4.6.2 Operating Cost Estimation

The following section summarises the costs for chemical consumables, waste removal, operational staff, maintenance, as well as insurance and tax operating costs used to conduct the economic analysis.

##### 4.6.1 Chemical Consumable Costs

Table 29, Table 30 and Table 31 provide a list of prices per ton of chemical consumed for the xylitol, CA and GA scenarios respectively. The price per ton is multiplied by the amount of chemical consumed, which is determined from the mass balances generated by the respective Aspen Plus® simulations.

Table 29 - Chemical Consumables Cost Summary for Xylitol Biorefinery Scenario (Görgens, et al., 2015)

Chemical Consumables and Products Summary	
Consumables	Price (US \$/ton)
Raney-nickel catalyst	\$18 500 <sup>a</sup>
Activated carbon	\$600 <sup>b</sup>
H <sub>2</sub> O (\$/kl)	\$0.0019 <sup>c</sup>
Feedstock (Bagasse & Trash)	\$11 <sup>d</sup>
H <sub>2</sub> SO <sub>4</sub>	\$112 <sup>d</sup>
NH <sub>3</sub>	\$335 <sup>e</sup>
CaO (lime)	\$75 <sup>b</sup>
H <sub>2</sub> gas	\$6 500 <sup>f</sup>
IE Chromatography resin (SAC)	\$0.10 <sup>b</sup>
IE Chromatography resin (WBA)	\$0.10 <sup>b</sup>

- 
- a - International price (Zauba, 2016)  
 b - International price (Alibaba, 2016)  
 c - eThekweni Municipal Commercial Tariff for >45 kl  
 d - Estimated cost of SA bagasse and trash (Görgens, et al., 2015)  
 e - International price (Jones, 2016)  
 f - Average cost \$3-10/kg (Eichman, et al., 2016) (James, et al., 2016)
- 

Table 30 - Chemical Consumables Cost Summary for CA Biorefinery Scenario (Anastassiadis & Rehm, 2006)

Chemical Consumables and Products	
Consumable	Price (US \$/ton)
NH <sub>3</sub>	\$335 <sup>a</sup>
H <sub>2</sub> O (\$/L)	\$0.0032 <sup>b</sup>
Feedstock (Bagasse & Trash)	\$11 <sup>c</sup>
H <sub>2</sub> SO <sub>4</sub>	\$112 <sup>d</sup>
CaO (lime)	\$75 <sup>e</sup>
Solvent	2747 <sup>f</sup>
Enzyme (Offsite Price)	\$6 000 <sup>e</sup>
Antifoam (Tween 60)	\$1 800 <sup>e</sup>
Activated carbon	\$600 <sup>e</sup>
<b>Fermentation medium</b>	
NH <sub>4</sub> Cl	\$220 <sup>e</sup>
KH <sub>2</sub> PO <sub>4</sub>	\$950 <sup>e</sup>
MgSO <sub>4</sub> .7H <sub>2</sub> O	\$95 <sup>e</sup>
MnSO <sub>4</sub> .4H <sub>2</sub> O	\$450 <sup>e</sup>
FeSO <sub>4</sub> .7H <sub>2</sub> O	\$95 <sup>e</sup>
CuSO <sub>4</sub> .5H <sub>2</sub> O	2250 <sup>e</sup>
ZnSO <sub>4</sub> .7H <sub>2</sub> O	730 <sup>e</sup>
CoSO <sub>4</sub> .7H <sub>2</sub> O	\$7750 <sup>e</sup>
H <sub>3</sub> BO <sub>3</sub>	\$815 <sup>e</sup>
CaCl <sub>2</sub>	\$160 <sup>e</sup>
NaCl	\$87 <sup>e</sup>
CA	1102 <sup>g</sup>
Na <sub>2</sub> MoO <sub>4</sub> .2H <sub>2</sub> O	8000 <sup>e</sup>
Thiamine	30000 <sup>e</sup>
Biotin	400000 <sup>e</sup>
Pyridoxine	25000 <sup>e</sup>
Ca-D-Pantothenate	20000 <sup>e</sup>
Nicotinic acid (niacin)	6000 <sup>e</sup>
Inoculum	\$0.024 <sup>h</sup>

- a - International price (Jones, 2016)  
 b - eThekweni Municipal Commercial Tariff for >45 kl  
 c - Estimated cost of SA bagasse and trash (Görgens, et al., 2015)  
 d - Basic Chemical, Omaha via HGI  
 e - International price (Alibaba, 2016)  
 f - Weighted average cost of all solvent chemicals from international prices (Zauba, 2016)  
 g - Average price calculated from literature and industry suppliers (Gurpreet Singh Dhillon et al., 2011)  
 h - Weighted average price for all inoculum chemicals (Alibaba, 2016)
-

Table 31 - Chemical Consumables Cost Summary for GA Biorefinery Scenario (Pal, et al., 2016; Miesher &amp; Haute, 1975)

Chemical Consumables and Products	
Consumable	Price (US \$/ton)
NH <sub>3</sub>	\$335 <sup>a</sup>
H <sub>2</sub> O (\$/L)	\$0.0019 <sup>b</sup>
Feedstock (Bagasse & Trash)	\$11 <sup>c</sup>
H <sub>2</sub> SO <sub>4</sub>	\$112 <sup>d</sup>
CaO (lime)	\$75 <sup>e</sup>
Biotin	\$400 000 <sup>e</sup>
Oleic acid	\$2 000 <sup>e</sup>
KH <sub>2</sub> PO <sub>4</sub>	\$950 <sup>e</sup>
MgSO <sub>4</sub>	\$95 <sup>e</sup>
FeSO <sub>4</sub> .7H <sub>2</sub> O	\$95 <sup>e</sup>
MnSO <sub>4</sub> .7H <sub>2</sub> O	\$450 <sup>e</sup>
Urea	\$242 <sup>a</sup>
Glucose	\$400 <sup>e</sup>
Enzyme (Offsite Price)	\$6 000 <sup>e</sup>
Antifoam (Tween 60)	\$1 800 <sup>e</sup>
Activated carbon	\$600 <sup>e</sup>

a - International price (Jones, 2016)

b - eThekweni Municipal Commercial Tariff for >45 kl

c - Estimated cost of SA bagasse and trash (Görgens, et al., 2015)

d - Basic Chemical, Omaha via HGI

e - International price (Alibaba, 2016)

#### 4.6.2 Waste Removal Costs

The main areas of waste for all biorefinery scenarios include ash and gypsum removal. The price attributed to removal is determined from past studies (Görgens, et al., 2015) and current industry sources (regyp, 2017). The costs associated with each waste are provided in Table 32 along with the annual waste removal costs for each biorefinery.

Table 32 - Waste Removal Costs

Solid Waste	Waste Removal Cost (\$/ton)
Ash	33.57
Gypsum	76.5

#### 4.6.3 Operational Staff Costs

Similarly operational staff costs (Table 33) are determined from past studies utilising similar compositions and quantities of biomass feed (Görgens, et al., 2015). The main distinction between biorefinery scenario staff costs, is that the xylitol scenario utilises a chemical synthesis route using a metal catalyst and not a fermentative biotechnological route. Therefore enzyme handling and

operation shift staff is naturally only considered for the CA and GA fermentative scenarios and therefore operational staff costs will be higher for these scenarios.

Table 33 - Staff Salary Summary (Görgens, et al., 2015)

Plant Staff Salary Costs						
Position	Salary (\$)	Year of salary quote	2016 Salary (\$)	# Xylitol	#CA	#GA
Plant Manager	147000	2009	168458	1	1	1
Plant Engineer	70000	2009	80218	2	2	2
Maintenance Supervisor	57000	2009	65320	1	1	1
Maintenance Technician	40000	2009	45839	8	8	8
Lab Manager	56000	2009	64174	1	1	1
Lab Technician	40000	2009	45839	2	2	2
Lab Tech-Enzyme	40000	2009	45839	0	2	2
Shift Supervisor	48000	2009	55007	4	4	4
Shift Operators	40000	2009	45839	15	15	15
Shift Oper-Enzyme	40000	2009	45839	0	6	6
Yard Employees	28000	2009	32087	3	3	3
Clerks & Secretaries	36000	2009	41255	3	3	3

#### 4.6.4 Other Costs (Maintenance, Insurance and Tax)

Maintenance, property insurance and taxes are determined as a percentage of the inside battery limit cost (ISBL) and fixed capital investment (FCI) and are displayed in Table 34. The respective ISBL for each biorefinery scenario is calculated by summing the total installed equipment cost for all process sections excluding the CHP plant, WWT plant, storage and utilities.

Table 34 - Maintenance, Insurance and Tax Operational Costs (Görgens, et al., 2015)

Other Costs	% Contribution
Maintenance	(3% of ISBL)
Property Insurance and Tax	(0.7% of FCI)

#### 4.6.3 Cash Flow Sheet Assumptions

The main assumptions for discounted cash flow analysis calculations are summarised in Table 35. Assumptions are all based on methods applied in previous work (Görgens, et al., 2015). The biorefineries operate 9 months per year as a result of annual harvest cycles for sugarcane. The biorefineries, also do not initially run at full capacity. The production capacity increases gradually from 50% in year 1 to 75% in year 2 and only reaches maximum production capacity after the 3<sup>rd</sup> year from the initial investment. The FCI is divided over a 3 year period prior to operation as seen in Table 35. The biorefinery project spans over a 25 year period and assumes a real term hurdle rate of 9.7% for the project to be considered profitable (Görgens, et al., 2015). Depreciation is assumed to follow a linear trend with an annual depreciation rate of 20% over a 5 year period to allow a 0 salvage value.

Operating costs and revenues are assumed to remain constant over the projects lifespan. Income over the projects lifespan is taxed at a rate of 28%. The average selling prices for products were determined from prices quoted in international markets and literature. A list of market prices for xylitol, CA and GA are provided in Appendices A7, B7 and C7 respectively.

Table 35 – Main Cash Flow Sheet Assumptions ( (Görgens, et al., 2015); (Farzad, et al., 2017))

Parameter	Value
Annual operating hours (9 months)	6480 h
Depreciation Type	Linear
Annual Depreciation (%)	20%
Equipment salvage value (\$)	\$0
Income Tax (%)	28%
Hurdle Rate (%)	15%
Ramp-up until operational at full capacity	
% year 1	50%
% year 2	75%
% year 3	100%
Plant Life (years)	25
Equity (%)	100%
% FCI spent in year -2	10%
% FCI spent in year -1	60%
% FCI spent in year 0	30%
Working Capital (% of FCI)	5%
Electricity Selling Price (\$/kWh)	\$0.08975
Average xylitol Selling Price (\$/ton)	\$4 538
Average CA Selling Price (\$/ton)	\$1 189
Average GA Selling Price (\$/ton)	\$3 625

#### 4.6.4 Profitability Indicators

Profitability indicators can be determined after capital and operating expenses have been calculated and processed in cash flow sheet analysis. Table 37 provides definitions for main profitability indicators that will be compared to the CHP baseline scenario in this study. Indicators include the net present value (NPV), total capital investment (TCI) and the internal rate of return (IRR).

Table 36 - Economic criteria to be compared between biorefinery and CHP simulation scenarios for economic viability

Performance Criteria	Description	Positive Outcome (Higher/Lower)
<b>TCI</b>	Total capital investment	Lower
<b>NPV</b>	Net Present Value - Cumulative discounted cash position at the end of the bio refinery project (Turton, et al., 2013). The sum of the present values of the future cash flows.	Higher
<b>DPBP</b>	Discounted Pay Back Period – Time after start-up required to recover the fixed capital investment for the project with all cash flows discounted from time zero	Lower
<b>IRR/DCFROR</b>	Interest rate for which the NPV of the bio refinery is equal to zero (Turton, et al., 2013)	Higher dependent on min acceptable rate of return on investment

#### 4.7 Environmental Impact

The environmental impact is quantified in terms of a few basic variables, namely greenhouse gas (GHG) emissions, solid waste output and water consumption. GHG emissions are evaluated in terms of annual biogenic and non-biogenic CO<sub>2</sub> emissions, with the former derived from biomass combustion (biomass carbon neutrality assumed) and the latter evaluating emissions derived from fossil fuel combustion and other process unit operations (excluding the boiler section) (Liu, et al., 2017). Total annual solid outputs include the disposal of ash, gypsum and biomass wastes. South Africa (SA) is categorised as a water scarce country, with recent droughts adding additional stress to water resources. A biorefinery is therefore considered more environmentally sustainable, the lower its annual water consumption. This is done in a basic manner in order to determine the scenario with the highest overall environmental impact.

It can however be more useful identifying scenarios that show a good balance between high profitability and low environmental impact. In this study this can be quantified by determining a weighted normalized score between 0 and 1 for both IRR and an overall environmental impact score. Equation 4 (section 2.6) can be used to normalize GHG emissions, water consumption and solid waste production. Equation 3 is then used to weight each impact into an overall score. A 60% weighting is applied to water consumption, 25% towards GHGs and 15% towards solid waste. Equation 4 and 3 can then be used again but weighted towards IRR and the environmental impact instead (80% and 20% respectively) to form an overall environmental-profitability composite score which will give indication as to which scenarios are able to generate high profit with low environmental impact (highest score between 0 and 1).

## 5. Results and Discussion

### 5.1 Utility and Bioproduct Production

The following section presents the mass and energy flow results for electricity, low pressure steam (LPS) and cooling medium consumption (water and air) for equipment in the xylitol, CA and GA biorefinery scenarios described in chapter 4. The highest utility consuming equipment as well as the utility distribution by plant area and the relationship of the results with regards to the bypass of biomass to the CHP plant are discussed. This section shows whether the utility requirements of the biorefineries are being under met or satisfied.

#### 5.1.1 Xylitol Equipment Utility Consumption

##### 5.1.1.1 Electricity Consumption

From Figure 32 and Table 37 we can see electricity consumption summarised according to plant area, with purification (Area 400), the CHP plant (Area 500) and catalytic reactor (Area 300) consuming the majority of the electricity (79%, 10% and 7% respectively). The xylitol scenario consumes under 8.22 MW of electricity and is able to produce 10.76 MW of electricity therefore providing an excess of 2.54 MW. The xylitol biorefinery scenario is therefore able to satisfy its entire electrical consumption.

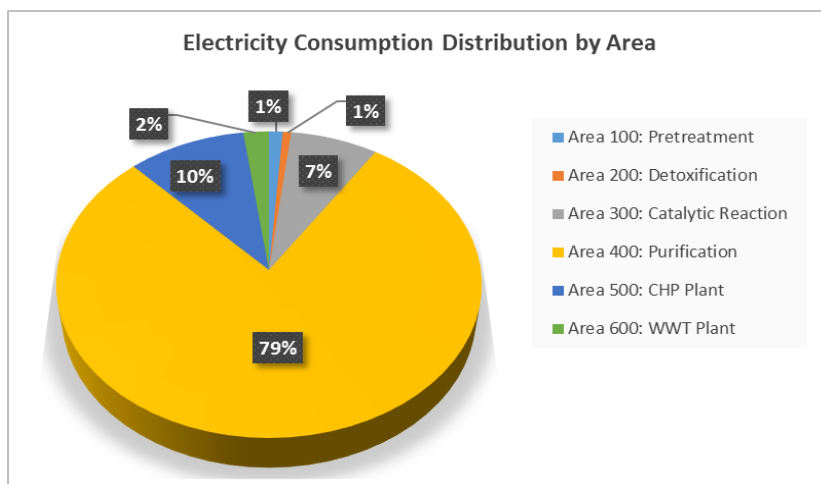


Figure 32 - Xylitol Electricity Distribution by Plant Area

Table 37 –Summary of Electricity Consumption According to Plant Sections for Xylitol Biorefinery

Plant Section	Electricity Consumption (kWh)
Area 100: Pretreatment	94
Area 200: Detoxification	57
Area 300: Catalytic Reaction	601
Area 400: Purification	6480
Area 500: CHP Plant	811
Area 600: WWT Plant	175
<b>Total Consumed</b>	<b>8218</b>



<b>Total Produced</b>	<b>10762</b>
<b>Excess</b>	<b>2543</b>

Insights into the energy distribution per area are highlighted in Figure 33 and Table 38 which summarise the distribution of the total electricity consumption for the xylitol scenario in terms of the top 7 electricity consuming pieces of equipment. The crystallisation and drying unit (combined as one unit) (S-403) is the highest energy consumer, accounting for 79% of the total electricity consumption (6475 kW), followed by the boiler feed pump (P-501), hydrogen gas compressor (C-301) and the boiler air blower (C-501) each consuming 6% (481 kW), 6% (473 kW) and 4% (329 kW) of the total electrical consumption respectively.

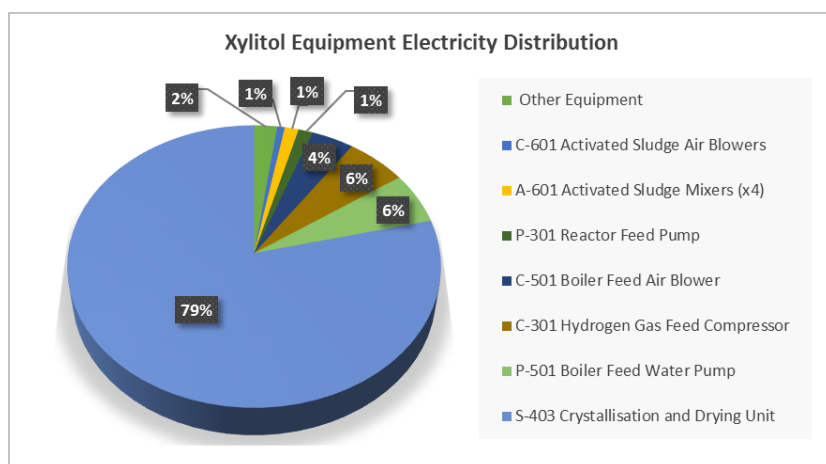


Figure 33 - Xylitol Equipment Electricity Distribution (Top 7 Consumers)

Table 38 – Summary of Highest Electricity Consumers for Xylitol Biorefinery

Equipment ID	Equipment Description	Equipment classification	Electricity Consumption (kWh)
	Other Equipment	Mixed	180
C-601	Activated Sludge Air Blowers	Compressor	58
A-601	Activated Sludge Mixers (x4)	Agitator	109
P-301	Reactor Feed Pump	Pump	113
C-501	Boiler Feed Air Blower	Compressor	329
C-301	Hydrogen Gas Feed Compressor	Compressor	473
P-501	Boiler Feed Water Pump	Pump	481
S-403	Crystallisation and Drying Unit	Separator	6475
<b>Total Consumed</b>			<b>8218</b>
<b>Total Produced</b>			<b>10762</b>
<b>Excess Electricity</b>			<b>2543</b>

#### 5.1.1.2 Low Pressure Steam (LPS) Consumption

LPS consumption by equipment type for the xylitol scenario is summarised in Table 39. It can be observed that the dilute acid tank reactor (R-101) consumes the most LPS (57.1 tons/h) followed by the evaporation column (S-402) that consumes the remaining amount of steam (40.95 tons/h) for a total of 98 tons/h. The xylitol process is able to generate 102.1 tons/h of steam, 4.05 tons/h in excess

of its heating requirements. The xylitol biorefinery scenario is therefore able to meet its steam heating requirements.

Table 39 - Summary of LPS Consumed by Equipment for Xylitol Biorefinery

Equipment ID	Equipment Description	Equipment classification	LPS Consumed (tons/h)
R-101	Dilute Acid Tank Reactor	Reactor	57.1
S-402	Evaporation Column	Separator	40.95
<b>LPS Consumed</b>			<b>98</b>
<b>LPS Produced</b>			<b>102.1</b>
<b>Excess LPS</b>			<b>4.05</b>

### 5.1.1.3 Cooling Water and Air

The xylitol scenario is the only biorefinery that utilizes air as a cooling medium. Cooling air and water consumption is listed in Table 40. The pre-feed cooler (E-302) uses the majority of air (311.6 tons/h) to cool hydrogen gas that enters the catalytic reactor. The remainder of cooling air is used by the cooler (E-505) after the first stage of the steam turbine (TURB-501, 10.53 tons/h).

Table 40 - Summary of Cooling Water and Air Consumed by Equipment for Xylitol Biorefinery

Utility	Equipment ID	Equipment Description	Equipment Classification	Cooling Air/Water (tons/h)
Cooling Air	E-302	Pre-feed Hydrogen Gas Cooler	Exchanger	311.64
Cooling Air	E-505	Turbine Outlet Steam Cooler	Exchanger	10.53
Cooling Water	E-601	Anaerobic Digester Cooling Jacket	Exchanger	48.57

## 5.1.2 CA Equipment Utility Consumption

### 5.1.1.1 Electricity

Figure 34 and Table 41 summarise CA electricity consumption by plant section. In descending order, fermentation (Area 300), wastewater treatment (WWT, Area 600) and pre-treatment (Area 100) consume the highest percentage of electricity (67%, 17.1% and 11.3% respectively), the majority of which is attributed to the sterilizer (E-201) (previously referred to in Figure 35). The biorefinery is able to meet its own electrical consumption needs and produces an excess of 6.35 MWh of excess electricity which can be sold for additional revenue.

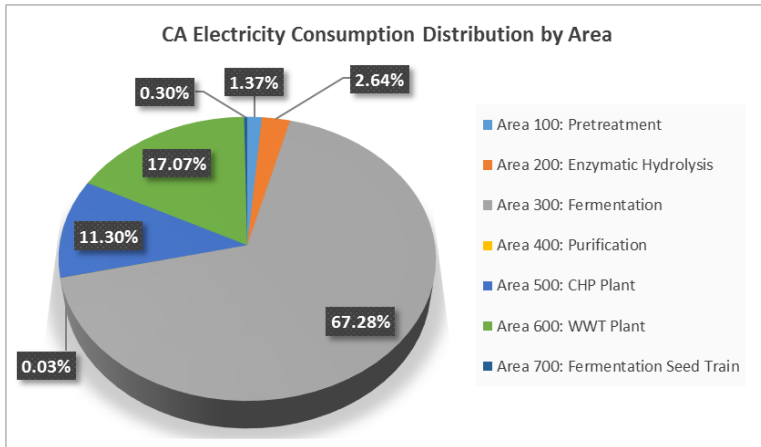


Figure 34 - CA Electricity Distribution by Plant Area

Table 41 - Summary of Electricity Distribution among Plant Sections for Xylitol Biorefinery

Plant Section	Electricity Consumption (kWh)
Area 100: Pretreatment	97
Area 200: Enzymatic Hydrolysis	187
Area 300: Fermentation	4764
Area 400: Purification	2
Area 500: CHP Plant	800
Area 600: WWT Plant	1209
Area 700: Fermentation Seed Train	21
<b>Total Consumed</b>	<b>5999</b>
<b>Total Produced</b>	<b>12348</b>
<b>Excess</b>	<b>6349</b>

Figure 35 and Table 42 detail the highest electricity consuming equipment for the CA biorefinery scenario. In descending order, the feed medium sterilizer (E-301), aerobic agitators (A-602), aerobic digestion air blowers (C-602) and the boiler feed pump (P-501) represent 89% of the electricity consumption (4616, 685, 502 and 463 kW respectively).

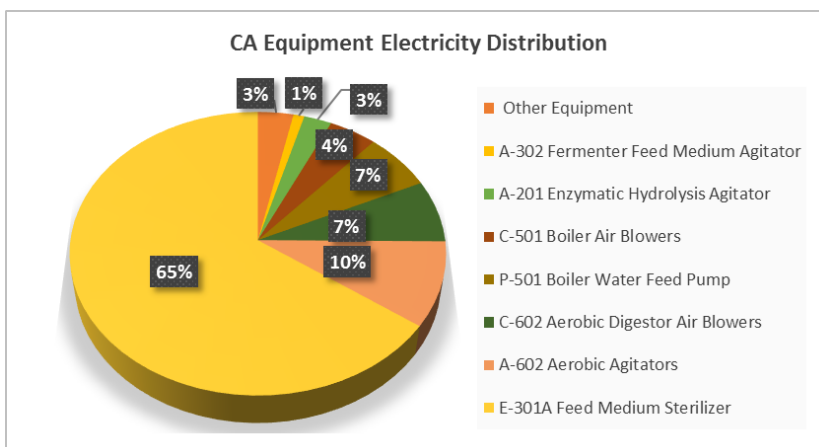


Figure 35 - CA Equipment Electricity Distribution

Table 42 - Summary of Highest Electricity Consumption among Equipment for CA Biorefinery

Equipment ID	Equipment Description	Equipment classification	Electricity Consumption (kWh)
	Other Equipment	Mixed	237
A-302	Fermenter Feed Medium Agitator	Agitator	73
A-201	Enzymatic Hydrolysis Agitator	Agitator	185
C-501	Boiler Air Blowers	Compressor	320
P-501	Boiler Water Feed Pump	Pump	463
C-602	Aerobic Digester Air Blowers	Compressor	502
A-602	Aerobic Agitators	Agitator	685
E-301A	Feed Medium Sterilizer	Exchanger	4616
<b>Total Consumed</b>			<b>5999</b>
<b>Total Produced</b>			<b>12348</b>
<b>Excess</b>			<b>6349</b>

### 5.1.1.2 Low Pressure Steam

Table 43 summarises the LPS consumption for the CA biorefinery scenario. The evaporation column (S-402) consumes the most LPS (64.5 ton/h), followed by the DA tank (13.1 tons/h) and CA fermenter jacket (9.7 tons/h). The CA biorefinery scenario is able to meet its own steam heating requirements and produces 15.16 tons/h of excess steam.

Table 43 - Summary of LPS Consumed by Equipment for CA Biorefinery

Equipment ID	Equipment Description	Equipment classification	LPS Consumed (tons/h)
R-101	Dilute Acid Tank	Reactor	13.14
E-302	Fermenter Jacket	Exchanger	9.71
S-402	Evaporation Column	Separator	64.52
<b>LPS Consumed</b>			<b>79.07</b>
<b>LPS Produced</b>			<b>94.2</b>
<b>Excess</b>			<b>15.16</b>

### 5.1.1.3 Cooling Water

In terms of cooling water consumption for the CA biorefinery, the pre-crystallisation cooler (E-401) consumes the highest quantity of water (130.5 tons/h, 83.2%), whereas the cellulignin slurry cooler (E-201) consumes the lowest quantity of water (9.1 tons/h, 5.8%) (Refer to Table 44).

Table 44 - Summary of Cooling Water Consumed by Equipment for CA Biorefinery

Equipment ID	Equipment Description	Equipment classification	Cooling Water Consumed (tons/h)
E-201	Cellulignin Slurry Cooler	Exchanger	9.1
E-202	Enzymatic Cooling Jacket	Exchanger	17.3
E-401	Pre-crystallisation Cooler	Exchanger	130.5
<b>Total Consumed</b>			<b>157</b>

### 5.1.3 GA Equipment Utility Consumption

#### 5.1.1.1 Electricity

GA equipment electricity consumption by plant area is provided in Figure 36 and Table 45. The WWT plant consumes the majority of electricity (47.96%). The CHP plant (Area 500) consumes the second highest quantity (25.41%), followed by enzymatic hydrolysis (Area 200) and pretreatment sections (Area 100) each consuming 7.84 and 5.65% respectively. The GA scenario has a total electricity consumption over 3 MW and produces under 14.5 MW, providing an excess of 11.4 MW of electricity. Therefore this scenario demonstrates the ability to meet its own energy requirements.

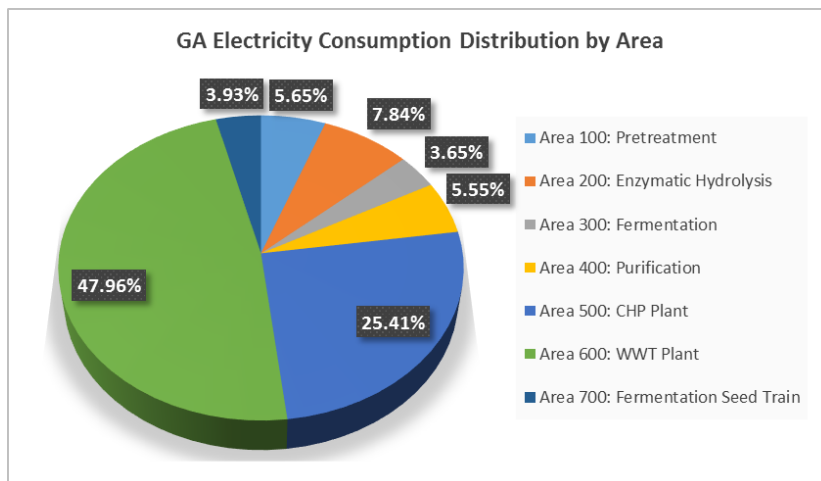


Figure 36 - GA Electricity Distribution by Plant Area

Table 45 - Summary of Electricity Usage among Plant Sections for GA Biorefinery

Plant Section	Electricity Consumption (kWh)
Area 100: Pretreatment	174
Area 200: Enzymatic Hydrolysis	241
Area 300: Fermentation	112
Area 400: Purification	171
Area 500: CHP Plant	782
Area 600: WWT Plant	1476
Area 700: Fermentation Seed Train	121
<b>Total Consumed</b>	<b>3079</b>
<b>Total Produced</b>	<b>14494</b>
<b>Excess</b>	<b>11415</b>

The highest electricity consuming equipment for the GA biorefinery scenario are detailed in Figure 37 and Table 46. The top 4 electricity consumers in descending order are the feed medium sterilizer (E-301), aerobic agitators (A-602), aerobic digester air blowers (C-602) and the boiler feed pump (P-501) (30%, 16%, 15% and 11% respectively).

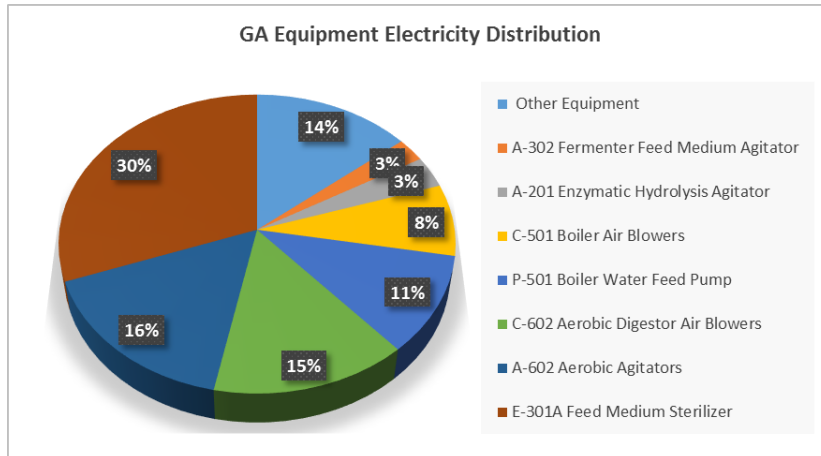


Figure 37 -GA Equipment Electricity Distribution

Table 46 - Summary of Highest Electricity Consumption among Equipment for GA Biorefinery

Equipment ID	Equipment Description	Equipment classification	Electricity Consumption (kWh)
	Other Equipment	Mixed	434
A-302	Fermenter Feed Medium Agitator	Agitator	79
A-201	Enzymatic Hydrolysis Agitator	Agitator	104
C-501	Boiler Air Blowers	Compressor	239
P-501	Boiler Water Feed Pump	Pump	330
C-602	Aerobic Digester Air Blowers	Compressor	452
A-602	Aerobic Agitators	Agitator	502
E-301A	Feed Medium Sterilizer	Exchanger	939
<b>Total Consumed</b>			<b>3079</b>
<b>Total Produced</b>			<b>14494</b>
<b>Excess</b>			<b>11415</b>

#### 5.1.1.2 Low Pressure Steam

Table 48 shows the LPS consumption of the GA biorefinery scenario. It can be seen that the scenario consumes 74 tons/h of steam with a total steam production of 89.1 tons/h, therefore providing 15.1 tons of excess steam. The scenario is able to meet its steam heating requirements.

Table 47 - Summary of LPS Consumed by Equipment for GA Biorefinery

Equipment ID	Equipment Description	Equipment classification	LPS Consumed (tons/h)
R-101	Dilute Acid Tank	Reactor	12.27
S-401	Evaporation Column	Separator	61.36
<b>Total Consumed</b>			<b>74</b>
<b>Total Produced</b>			<b>89.1</b>
<b>Excess</b>			<b>15.1</b>

### 5.1.1.3 Cooling Water

Table 48 summarises the cooling water consumption for the GA biorefinery scenario. It can be seen that the seed fermenter train as well as the recrystallization cooler consume the highest quantity of cooling water. The scenario’s total water consumption amounts to 175 tons/h.

Table 48 - Summary of Cooling Water Consumed by Equipment for GA Biorefinery

Equipment ID	Equipment Description	Equipment classification	Water Consumed (tons/h)
R-701	Seed Fermenter Train	Reactor	98.92
E-201	Cellulignin Slurry Cooler	Exchanger	10.31
E-202	Enzymatic Cooling Jacket	Exchanger	20.43
E-402	Pre-crystallisation Cooler	Exchanger	45.51
<b>Total Consumed</b>			<b>175</b>

## 5.2 Bioproduct and Electricity Production

### 5.2.1 Product Quantity and Purity Comparison

Figure 38 compares the different biorefinery scenarios in terms of product quantity per annum, product purity (% w/w) and surplus electricity production (kWh). It can be observed that the xylitol biorefinery produces the lowest quantity of product out of all the biorefineries and has the highest product purity in comparison to the CA and GA scenarios (38 789 tons of xylitol/year with a 99% purity). The CA scenario produces the highest quantity of product per annum with the second highest purity (97 893 tons CA/year, 98% purity), followed by the GA scenario with the second highest quantity of product produced per annum and the lowest product purity (83 005 tons GA/annum, 97.8% purity).

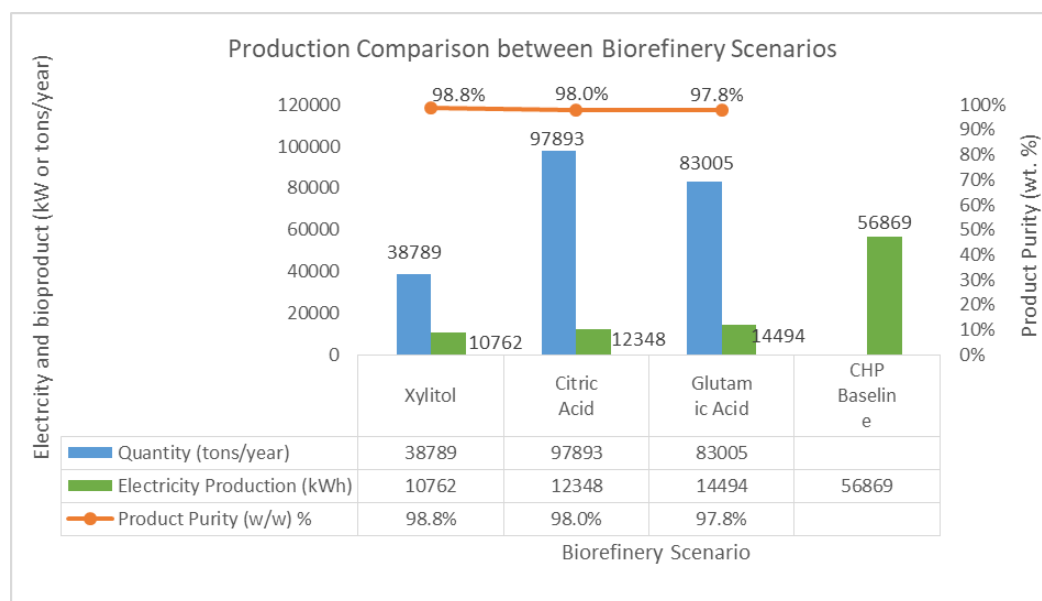


Figure 38 – Comparison of Bioproduct Production (Xylitol, CA, GA, and Baseline)

The composition of cellulose is higher than hemicellulose in the sugarcane bagasse and trash feedstock investigated, therefore it is evident, that with the conversions reviewed for DA hydrolysis and enzymatic hydrolysis, that higher quantities of CA and GA would naturally be produced.

### 5.2.2 Electricity Production Comparison

Figure 39 compares the electricity production and consumption of all biorefinery scenarios. The CHP plant naturally produces the highest quantity of electricity (56.1 MW) as all biomass is bypassed to the boiler (100% biomass bypass). When comparing biorefineries relative to each other (excluding the baseline), the GA scenario simultaneously produces the highest quantity of electricity (35% bypass) and consumes the least electricity giving a total excess of 11.4 MW. Reasons for simultaneous high electricity production and low electricity consumption for the GA scenario could be attributed to the use of membrane integrated reactor systems that eliminate a number of the separation units that would typically consume large amounts of electricity in conventional processes. The xylitol scenario consumes the highest quantity of electricity of all the scenarios and produces the lowest quantity of excess energy (8.22 MW and 2.54 MW respectively). This is mainly due to the utilization of electricity as an energy source for the crystallisation and dryer, in place of low pressure steam. This decision was undertaken to ensure that the correct balance of electricity and steam was produced, so that the xylitol biorefinery scenario could be energy self-sufficient.

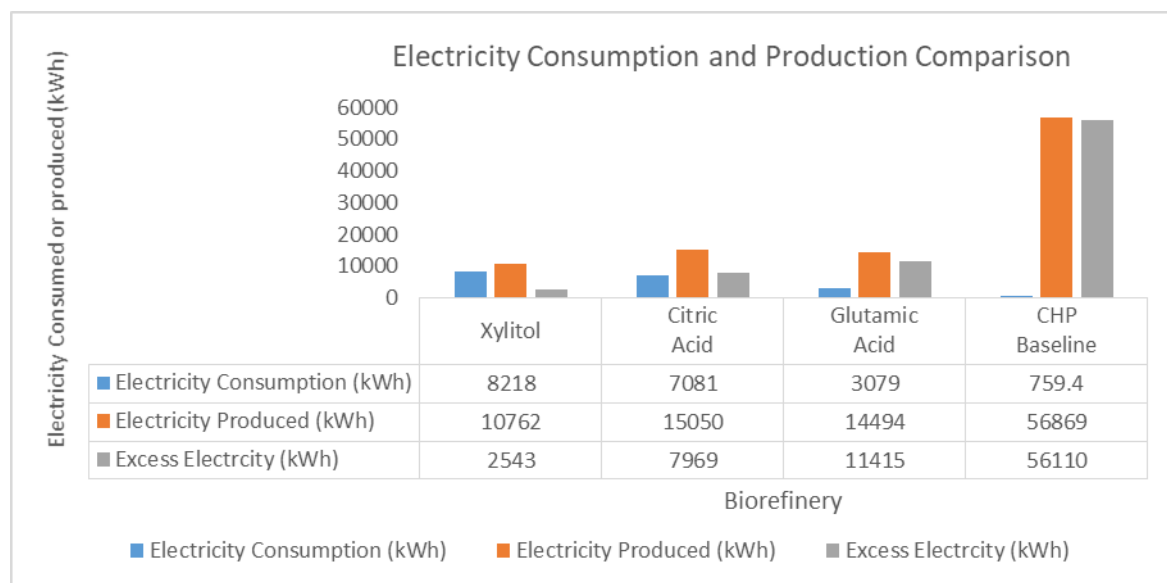


Figure 39 - Comparison of Electricity Production (Xylitol, CA, GA, and Baseline)



### 5.3 Operating Expenditure

In this section, operating expenses are summarised according to their classification (both variable and fixed, direct and indirect) for each biorefinery scenario and later compared to the CHP plant baseline to rank the scenario with the lowest total operating costs (TOC).

#### 5.3.1 Xylitol OPEX

The variable operating costs for the xylitol scenario consist of chemical expenses and waste disposal costs, which together constitute \$75.8 million (Table 49). From Figure 40 it can be observed that the majority of the operating expenditure is comprised of activated carbon, the Raney-nickel catalyst, hydrogen feed gas and feedstock costs. Together these constitute 77% of total variable operating costs, \$ 57.9 million.

A significant amount of the variable operating costs could be reduced if the amount of catalyst recovered after the reactor section (Area 300) could be increased. In addition, all of the hydrogen currently utilised for catalytic hydrogenation of xylose is bought off-site. If hydrogen was produced on site via hydrolysis or charcoal gasification, the variable operating costs could potentially be significantly reduced, albeit at a compromised higher capital cost. Resin costs for both chromatography columns amount to over \$3.1 million per annum. An increase in the amount of resin recovered or sourcing resin at a cheaper price from competitive suppliers can also lower costs.

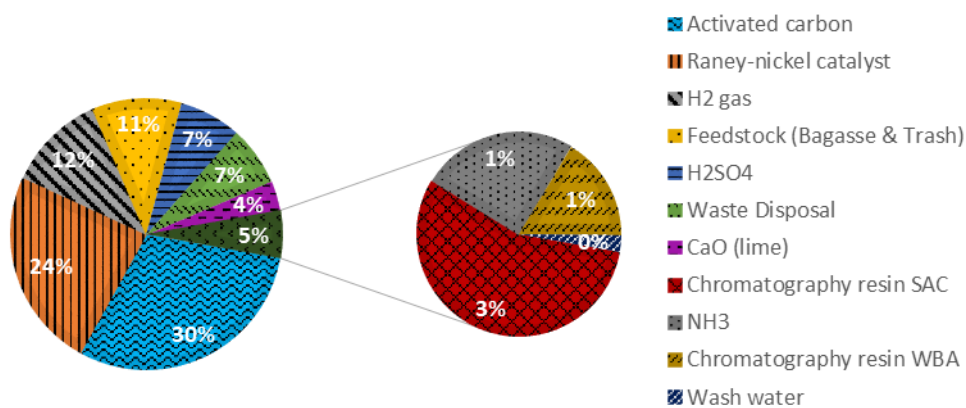


Figure 40 - Xylitol Variable Operating Cost Distribution

Table 49 - Xylitol Variable Operating Costs

Operating Cost	Cost (\$ million/annum)
Activated carbon	\$22.9
Raney-nickel catalyst	\$18.0
H <sub>2</sub> gas	\$8.9
Feedstock (Bagasse & Trash)	\$8.0
H <sub>2</sub> SO <sub>4</sub>	\$5.6
CaO (lime)	\$5.4
Chromatography resin SAC	\$2.7
NH <sub>3</sub>	\$2.4
Chromatography resin WBA	\$1.0
Waste Disposal	\$0.7
Wash water	\$0.1
<b>Total Variable Operating Costs</b>	<b>\$75.8</b>

The total operating cost (TOCs) consisting of variable and fixed costs are summarised in Table 50. Maintenance, labour and insurance make up a total fixed operating cost amounting to \$3.48 million. This consists of maintenance costs, property insurance, tax and total staff salaries. The TOC for the xylitol scenario amounts to \$79.25 million.

Table 50 - Xylitol TOC Summary

Other Costs	Cost (\$ million/annum)
Maintenance (3% ISBL)	1.17
Property Insurance and Tax (0.7% of FCI)	0.27
Total Staff Salaries	2.04
Chemical Expenses (\$/year)	70.4
<b>Total Variable Operating Costs(\$/year)</b>	<b>75.8</b>
<b>Total Fixed Operating Costs (\$/year)</b>	<b>3.48</b>
<b>Total Operating Cost (TOC) \$/year</b>	<b>79.25</b>

### 5.3.2 Citric Acid (CA) Production Scenario OPEX

CA variable costs are detailed in Figure 41 and Table 51. A clear observation is that the majority of the total variable operating costs are attributed to the amine-alcohol solvent utilized during solvent extraction and enzymes utilized during the cellulose hydrolysis steps (28% and 18% respectively). Feedstock and H<sub>2</sub>SO<sub>4</sub> provide the third and fourth highest contribution to variable operating costs respectively (14% and 8%). Other costs contribute 7%, with this mainly consisting of fermentation medium, activated carbon and inoculum medium costs (refer to Table 51). Total variable operating costs for the CA biorefinery scenario amount to \$55.5 million per annum. 28% of all variable operating costs could be reduced if all of the amine-alcohol solvent after solvent extraction could be recovered.

Onsite enzyme production for this scenario is another suggestion to greatly reduce operating costs for this process but at the expense of additional capital investment. Humbird, et al. (2011) highlighted that 57% of the onsite cost of enzyme production from *Trichoderma Reesi* for bioethanol production was attributed to the cost of the glucose substrate. Total on-site enzyme costs amounted to \$4240/ton protein (Humbird, et al., 2011). When adjusted to present prices this would total \$4599/ton, which is a 23.4% reduction in cost from the currently used off-site enzyme cost of \$6000/ton.

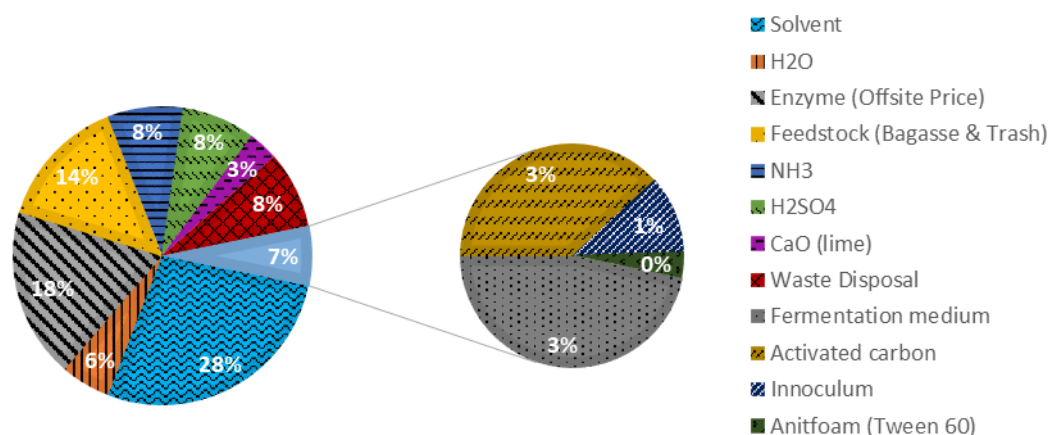


Figure 41 - CA Variable Operating Cost Distribution

Table 51 - CA Variable Operating Costs

Operating Cost	Cost (\$ million/annum)
Solvent	15.5
H2O	3.0
Enzyme (Offsite Price)	10.3
Feedstock (Bagasse & Trash)	8.0
NH3	4.6
H2SO4	4.5
CaO (lime)	1.8
Waste Disposal	4.6
<b>Other</b>	<b>3.6</b>
Fermentation medium	1.7
Activated carbon	1.4
Inoculum	0.4
Antifoam (Tween 60)	0.1
<b>Total</b>	<b>56.0</b>

The total operating cost (TOCs) consisting of variable and fixed costs are summarised in Table 52 for the CA scenario. Total fixed operating costs amount to \$4.01 million/year, whereas TOC are \$60 million/year.

Table 52 - CA TOC Summary

Other Costs	Cost (\$ million/annum)
Maintenance (3% ISBL)	1.3
Property Insurance and Tax (0.7% of FCI)	0.3
Total Staff Salaries	2.41
Chemical Expenses (\$/year)	51.36
<b>Total Variable Operating Costs(\$/year)</b>	<b>55.99</b>
<b>Total Fixed Operating Costs (\$/year)</b>	<b>4.01</b>
<b>Total Operating Cost (TOC) \$/year</b>	<b>60</b>

### 5.3.3 GA OPEX

GA variable operating costs are summarised in Figure 42 and Table 53. It can be seen that the majority of the cost is distributed among enzyme, ammonia and feedstock use (75% of total variable operating cost). The remainder of the costs are allocated to H<sub>2</sub>SO<sub>4</sub>, water, lime and activated carbon consumption. Together with other costs, the total variable operating cost amounts to \$100.9 million. The GA scenario utilises micro and nanofiltration modules to separate impurities from the outlet fermentation broth. This reduces chemical costs normally associated with other purification techniques such as the solvent extraction or IE chromatography route utilised for the CA and xylitol respectively. In addition a lower amount of energy would need to be utilised during evaporation and crystallisation steps in comparison to the other biorefineries (xylitol and CA). The GA scenario bypasses 35 % of the total biomass to the CHP plant and therefore processes a larger quantity of bagasse and trash in the biorefinery, this results in higher total operating costs in comparison to the fermentative CA scenario (45% bypass). Enzyme and ammonia costs for the GA scenario are more than 4 times greater than the CA scenario. Therefore a higher bypass of biomass to the CHP plant equates to lower operating costs as a result of processing less feedstock in the biorefinery.

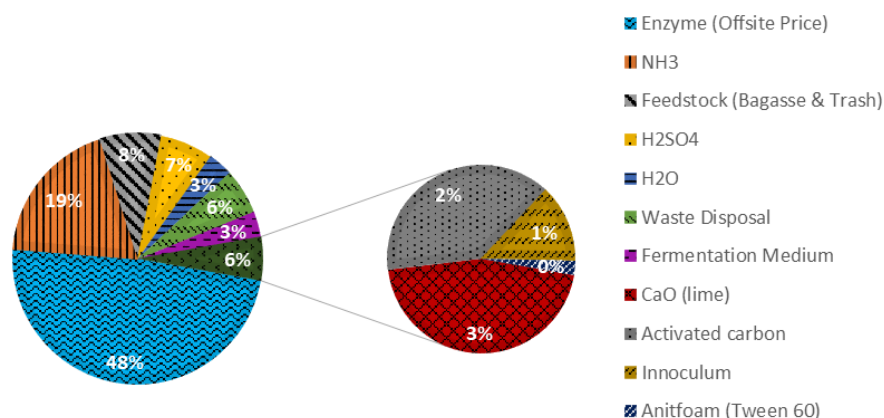


Figure 42 - GA Variable Operating Cost Distribution

Table 53 - GA Variable Operating Costs

Operating Cost	Cost (\$ million/annum)
Enzyme (Offsite Price)	48.9
NH3	19.0
Feedstock (Bagasse & Trash)	8.0
H2SO4	6.9
H2O	3.2
Waste Disposal	6.0
Fermentation Medium	3.3
CaO (lime)	2.5
Activated carbon	2.2
Inoculum	0.7
Antifoam (Tween 60)	0.1
<b>Total</b>	<b>100.9</b>

A summary of total variable, fixed and TOC is provided in Table 54. Maintenance, property insurance, tax and staff salaries provide total fixed operating costs amounting to \$8.2 million. Together the TOC amounts to \$109.1million.

Table 54 - GA TOC Summary

Other Costs	Cost (\$ million/annum)
Maintenance (3% ISBL)	3.0
Property Insurance and Tax (0.7% of FCI)	2.8
Chemical Expenses (\$/year)	94.9
<b>Total Variable Operating Costs(\$/year)</b>	<b>100.9</b>
<b>Total Fixed Operating Costs (\$/year)</b>	<b>8.2</b>
<b>Total Operating Cost (TOC) \$/year</b>	<b>109.1</b>

### 5.3.4 OPEX Comparison (Including CHP Baseline Scenario)

Table 55 compares the TOCs and revenue per annum of biorefinery scenarios to the CHP plant baseline. The GA scenario had the highest TOC (\$109.1 million) and revenue (\$303 million) out of all the biorefineries, followed by the xylitol (\$79 million), CA (\$60 million) and CHP baseline scenarios (\$13 million). Profit margins for each of the scenarios are calculated as a percentage by dividing the difference between total sales revenue and TOCs by the total sales. The GA scenario had the highest profit margin (64%) out of all the scenarios as a result of its high sales revenue and low TOCs, as physical membrane separation avoids the use of additional chemicals used during other purification methods and the high selling price of GA combined with high bypass of biomass to the biorefinery substantially increases revenue. However in terms of overall TOCs, a lower bypass of lignocellulose to the CHP plant for the GA (35% bypass) biorefinery scenarios in comparison to the CA scenario (45% bypass) increases the amount of chemical consumables required to process the feed. Therefore a lower bypass of biomass to the CHP plant i.e. higher bypass to biorefinery, increases TOC.

Table 55 - OPEX and Revenue Comparison Summary (\$million/year)

Biorefinery	% Bypass	TOC (\$/year)	Total Sales Revenue (\$/year)	Profit Margin (%)
Xylitol	46%	\$79	\$175	55%
Citric Acid	45%	\$60	\$120	50%
Glutamic Acid	35%	\$109	\$303	64%
CHP Baseline	100%	\$13	\$33	61%

## 5.4 Capital Expenditure

The following section details the capital costs associated with each biorefinery scenario organised according to plant sections. The capital expenditure (TCI) of each biorefinery scenario is compared to the CHP baseline in order to determine the plant with the highest investment requirements.

### 5.4.1 Xylitol Biorefinery CAPEX

Figure 43 displays the distribution of equipment costs for the xylitol biorefinery scenario whereas Table 57 summarises the total direct costs (TDC), the total indirect costs (TIDC) as well as the FCI and TCI. It can be seen that the CHP plant (Area 500) contributes the most to the capital equipment cost (56%), followed by the catalytic reaction section (Area 300), WWT (Area 600) and pretreatment sections (Area 100), contributing 17%, 9% and 7% to capital equipment costs respectively.

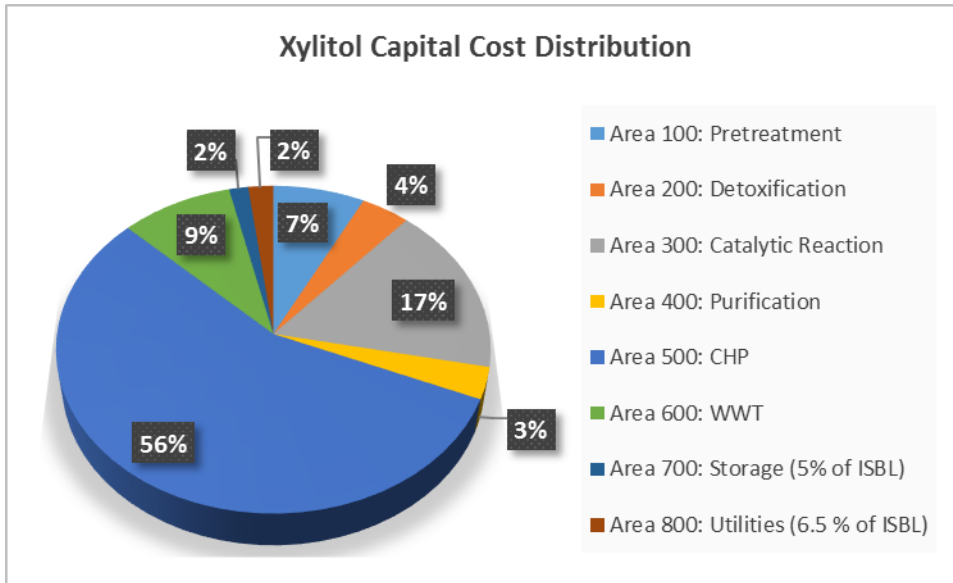


Figure 43 – Xylitol Equipment Cost Distribution per Area

Table 56 - Installed Cost Distribution (Xylitol-Electricity Co-production)

Equipment ID	Equipment Description	Installed Equipment Cost (\$ million)	%
R-501	CHP Boiler	\$ 48.01	39.8%
R-301	Catalytic Xylitol Reactor (x3)	\$ 18.27	15.1%
C-501	Boiler Feed Air Blower	\$ 6.97	5.8%
TURB-502	Condensing Steam Turbine Stage 2	\$ 6.56	5.4%
R-601	Anaerobic Digester	\$ 6.33	5.2%
E-503	Flue gas heat exchanger 1	\$ 3.87	3.2%
S-401	SAC SMB Chromatography Unit	\$ 3.35	2.8%
TURB-501	Condensing Steam Turbine Stage 1	\$ 3.09	2.6%
C-301	Hydrogen Gas Feed Compressor	\$ 1.91	1.6%
Other	Other	\$ 22.32	18.5%

The TDC and TIDC for the xylitol scenario amounts to \$131 million and \$78.6 million resulting in a FCI of \$209.6 million and TCI of \$220.1 million (see Table 57).

Table 57 - Xylitol Capital Expenditure Summary

<b>Equipment Summary</b>		
<b>Area</b>	<b>Area</b>	<b>Installed Cost (\$ million)</b>
Area 100: Pretreatment	100	\$9.3
Area 200: Detoxification	200	\$5.0
Area 300: Catalytic Reaction	300	\$20.8
Area 400: Purification	400	\$3.9
Area 500: CHP	500	\$69.4
Area 600: WWT	600	\$11.4
Area 700: Storage (5% of ISBL)	700	\$1.9
Area 800: Utilities (6.5 % of ISBL)	800	\$2.5
<b>ISBL Total</b>		<b>\$38.9</b>
<b>Total Equipment Cost</b>		<b>\$124.2</b>
Warehouse (4 % of ISBL)		\$1.56
Site Development (9 % of ISBL)		\$3.50
Additional Piping (4.5 % of ISBL)		\$1.75
<b>Total Direct Costs (TDC)</b>	<b>% of TDC</b>	<b>\$130.99</b>
Prorate able Expenses	10	\$13.1
Field Expenses	10	\$13.1
Home Office and Construction	20	\$26.2
Project Contingency	10	\$13.1
Other Costs incl. start-up, permits etc.	10	\$13.1
<b>Total Indirect Costs (TIDC)</b>		<b>\$78.59</b>
<b>Fixed Capital Investment (FCI = TDC + TIDC)</b>		<b>\$209.58</b>
Working Capital (WC 5% of FCI)		\$10.48
<b>Total Capital Investment (TCI = FCI + WC)</b>		<b>\$220.1</b>

#### 5.4.2 Citric Acid Biorefinery CAPEX

The top 4 plant sections with the highest contribution to CA capital equipment cost in descending order include the CHP plant (Area 500), WWT plant (Area 600), enzymatic hydrolysis (Area 200), pretreatment sections (Area 100) each contributing 48%, 23%, 15% and 5% respectively to a total equipment capital cost of \$157.4 million (see Figure 44 and Table 59).



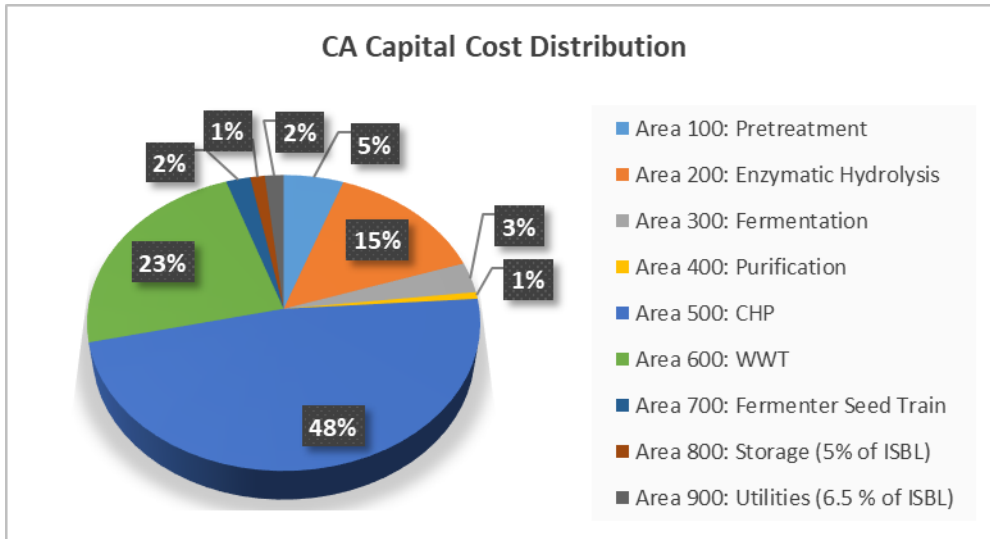


Figure 44 - CA Capital Cost Distribution

As seen in Table 58 the majority of capital expenditure in the CA scenario is attributed to the CHP plant boiler, enzymatic hydrolysis tank and the membrane bioreactor that is integrated into the aerobic activated sludge process. When considering the choice of feedstock, biorefineries using 2<sup>nd</sup> generation biomass for fermentation would naturally have higher capital costs due to the requirement of accessing monomer sugars that make up the lignocellulosic structure for fermentation.

Table 58 - Installed Cost Distribution (CA-Electricity Co-production Scenario)

Equipment ID	Equipment Description	Installed Equipment Cost (\$ million)	%
H-501	CHP Boiler (Water Heater)	\$ 46.98	30.8%
R-202	Enzymatic Hydrolysis Tank (2-stage)	\$ 23.46	15.4%
F-601	Membrane Bioreactor Filtration	\$ 18.65	12.2%
E-504	Flue Gas Cooler 2	\$ 11.17	7.3%
R-601	Anaerobic Digester	\$ 10.16	6.7%
C-501	Boiler Air Blowers	\$ 6.80	4.5%
R-101	Dilute Acid Tank	\$ 2.73	1.8%
E-503	Flue Gas Cooler 1	\$ 2.50	1.6%
R-301	Citric Acid Fermenter (x3)	\$ 2.45	1.6%
Other	Other	\$ 28	18.0%

The CA scenario's TDC and TIDC as seen in Table 59, amount to \$173.3 million and \$104.0 million respectively, resulting in a FCI of \$277.3 million and a TCI of \$291.2 million.

Table 59 - CA Capital Expenditure Summary

<b>Equipment Summary</b>		
<b>Area</b>	<b>Area</b>	<b>Installed Cost (\$ million)</b>
Area 100: Pretreatment	100	\$9.0
Area 200: Enzymatic Hydrolysis	200	\$23.9
Area 300: Fermentation	300	\$5.5
Area 400: Purification	400	\$1.2
Area 500: CHP	500	\$78.7
Area 600: WWT	600	\$38.8
Area 700: Fermenter Seed Train	700	\$3.7
Area 800: On-site Enzyme Production (not modelled)	800	-
Area 900: Storage (5% of ISBL)	900	\$2.2
Area 1000: Utilities (6.5 % of ISBL)	1000	\$2.8
<b>ISBL Total</b>		<b>\$43.3</b>
<b>Total Equipment Cost</b>		<b>\$165.7</b>
Warehouse (4 % of ISBL)		\$1.7
Site Development (9 % of ISBL)		\$3.9
Additional Piping (4.5 % of ISBL)		\$1.9
<b>Total Direct Costs (TDC)</b>	<b>% of TDC</b>	<b>\$173.3</b>
Prorate able Expenses	10	\$17.3
Field Expenses	10	\$17.3
Home Office and Construction	20	\$34.7
Project Contingency	10	\$17.3
Other Costs incl. start-up, permits etc.	10	\$17.3
<b>Total Indirect Costs (TIDC)</b>		<b>\$104.0</b>
<b>Fixed Capital Investment (FCI = TDC + TIDC)</b>		<b>\$277.3</b>
Working Capital (WC 5% of FCI)		\$13.9
<b>Total Capital Investment (TCI = FCI + WC)</b>		<b>\$291.2</b>

#### 5.4.3 Glutamic Acid Biorefinery CAPEX

Figure 45 shows the equipment capital cost distribution by area for the GA biorefinery scenario. The CHP plant (Area 500) is the highest contributor to equipment capital cost (35%), followed by WWT (Area 600) (17%), GA fermentation (Area 300) (15%), enzymatic hydrolysis (Area 200) (12%) and purification (Area 400) (7%). The total equipment capital cost amounts to \$208.8 million (see Table 61).

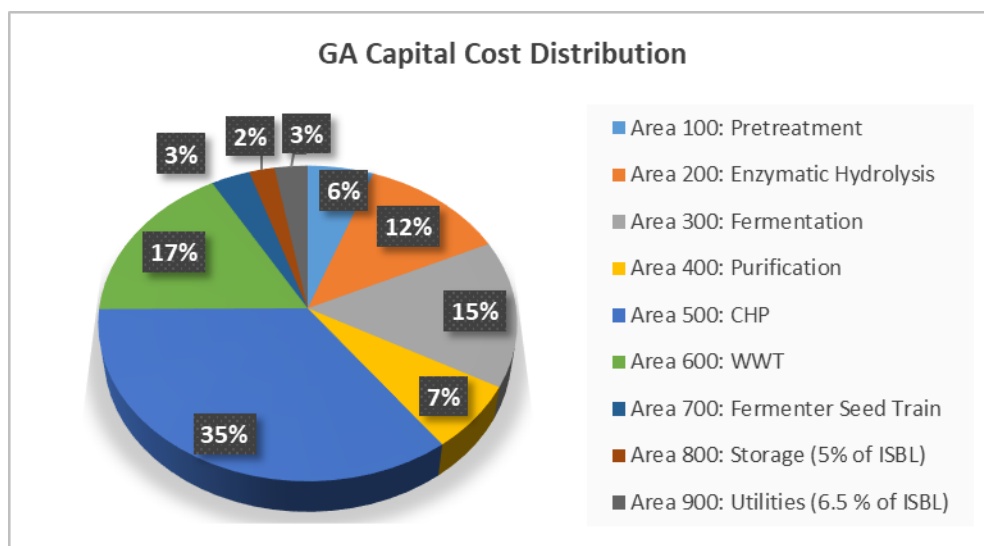


Figure 45 - GA Capital Cost Distribution

As is the case with the CA biorefinery scenario, the GA scenario attributes most of its capital expenditure to the CHP plant's boiler, the GA fermenters and the enzymatic hydrolysis tank (see Table 60). GA fermenters seem to contribute higher amounts of capital expenditure than those utilised for CA. This is mostly due to the addition of integrated (both micro and nanofiltration) membranes.

Table 60 - Installed Cost Distribution (GA-Electricity Co-production Scenario)

Equipment ID	Equipment Description	Installed Equipment Cost (\$ million)	%
H-501	CHP Boiler (Water Heater)	\$46.29	20.9%
R-301	Glutamic Acid Fermenter (x8)	\$30.09	13.6%
R-202	Enzymatic Hydrolysis Tank (2-stage)	\$28.73	12.9%
F-601	Membrane Bioreactor Filter Modules	\$23.55	10.6%
E-501	Air Preheater	\$22.26	10.0%
TURB-501	Condensing Steam Turbine (Stage 1)	\$8.99	4.0%
F-401	Microfiltration Module Train	\$7.26	3.3%
F-402	Nano filter Module Train NF20	\$5.57	2.5%
R-701	Seed Fermenter Train	\$5.14	2.3%
Other (GA)	Other (GA)	\$44.03	19.8%

The GA scenario's TDC and TIDC as seen in Table 61, amount to \$251.1 million and \$150.7 million respectively, resulting in a FCI of \$401.8 million and a TCI of \$421.9 million.

Table 61 - GA Capital Expenditure Summary

<b>Equipment Summary</b>		
<b>Area</b>	<b>Area</b>	<b>Installed Cost (\$ million)</b>
Area 100: Pretreatment	100	\$13.0
Area 200: Enzymatic Hydrolysis	200	\$28.9
Area 300: Fermentation	300	\$35.1
Area 400: Purification	400	\$16.0
Area 500: CHP	500	\$81.9
Area 600: WWT	600	\$39.3
Area 700: Fermenter Seed Train	700	\$7.7
Area 800: Storage (5% of ISBL)	800	\$5.0
Area 900: Utilities (6.5 % of ISBL)	900	\$6.5
<b>ISBL Total</b>		<b>\$100.8</b>
<b>Total Equipment Cost</b>		<b>\$233.5</b>
Warehouse (4 % of ISBL)		\$4.0
Site Development (9 % of ISBL)		\$9.1
Additional Piping (4.5 % of ISBL)		\$4.5
<b>Total Direct Costs (TDC)</b>	<b>% of TDC</b>	<b>\$251.1</b>
Prorate able Expenses	10	\$25.1
Field Expenses	10	\$25.1
Home Office and Construction	20	\$50.2
Project Contingency	10	\$25.1
Other Costs incl. start-up, permits etc.	10	\$25.1
<b>Total Indirect Costs (TIDC)</b>		<b>\$150.7</b>
<b>Fixed Capital Investment (FCI = TDC + TIDC)</b>		<b>\$401.8</b>
Working Capital (WC 5% of FCI)		\$20.1
<b>Total Capital Investment (TCI = FCI + WC)</b>		<b>\$421.9</b>

#### 5.4.4 CAPEX Comparison (Including CHP Baseline Scenario)

Table 62 - Comparison of Capital Investment (\$ million)

<b>Biorefinery</b>	<b>Bypass to CHP (%)</b>	<b>FCI (\$ million)</b>	<b>TCI (\$ million)</b>
Xylitol	46	\$ 209.6	\$ 220.1
Citric Acid	45	\$ 277.3	\$ 291.2
Glutamic Acid	35	\$ 401.8	\$ 421.9
CHP Baseline	100	\$ 132	\$ 138.6

The GA biorefinery has the highest FCI (\$402 million), followed by CA (\$277 million) and xylitol scenarios (\$210 million) (see Table 62). Conversely the CHP baseline has the lowest FCI (\$132 million) as seen in Table 62, due solely to its lack of processing complexity as all biomass is bypassed for combustion. The GA scenario bypasses the most lignocellulose to its biorefinery (65% of biomass goes through to the biorefinery). This results in higher operating as well as capital costs compared to xylitol and CA scenarios. Therefore, scenarios that bypass more of their biomass to the CHP plant for energy (as is the case with xylitol, 46% biomass bypass), will have a lower throughput of lignocellulose to the product biorefinery and therefore also have lower equipment costs. Other reasons for the low xylitol scenario capital expenditure include utilising hemicellulosic sugars that do not require the need for additional enzymatic hydrolysis and further down-stream separation units as is the case for CA and GA scenarios.

## 5.5 Financial Performance of Biorefinery Scenarios

This section characterises the economic performance of each biorefinery scenario in terms of key financial indicators. Indicators used to determine profitability include the internal rate of return (IRR), net present value (NPV), total capital investment (TCI) and pay-back period (PBP). Performance is provided for a range of market prices (maximum, minimum and average). In addition to the review of individual scenario financial performance, a comparison of each scenarios investment and profitability to the CHP baseline scenario is also outlined (average product price used), along with a detailed look at the viability of the results with regards to world demand.

### 5.5.1 Profitability Indicators (NPV, IRR, TCI, PBP)

The following section characterises the economic performance of the scenarios investigated in terms economic indicators. Indicators were reported for different selling prices (maximum, average and minimum product price) to account for uncertainty and show market conditions under which a product scenario would be able or unable to generate profit.

#### 5.5.1.1 Xylitol 46 % bypass (High, Low and Average Product Price)

At a 46 % bagasse and trash bypass to the CHP plant and for an average xylitol market price of \$4538/ton, the xylitol biorefinery scenario produces an IRR of 30% and a NPV of \$405 million for a TCI of \$220.1 million proving that the scenario is profitable under average market price conditions (refer to Figure 46). However, when the lowest market price of \$2500/ton is used the scenario's IRR and NPV falls to unprofitable conditions (4.02%) and conversely when the maximum market price of

\$11 985/ton is used, profitability swings to overly optimistic conditions with an IRR of 91% and an NPV of \$2.15 billion. Profitability is therefore highly dependent on the price of xylitol.

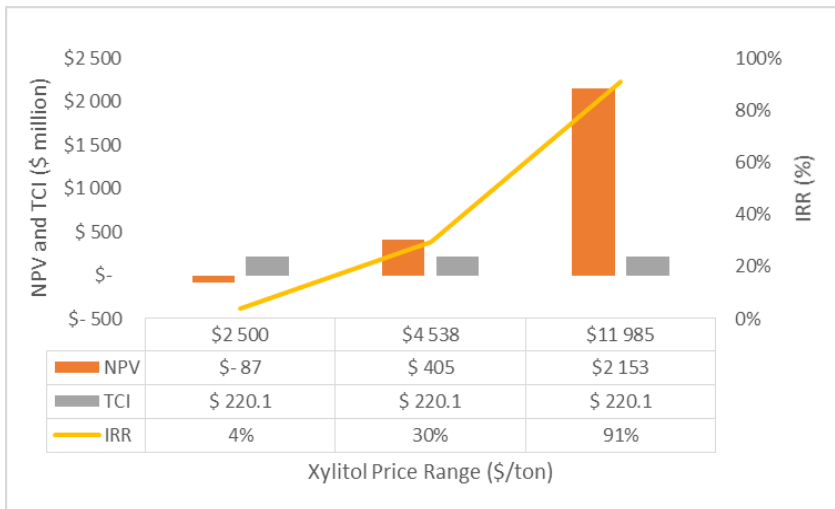


Figure 46 - Xylitol Profitability Range for Max, Ave and Min Xylitol Prices

For a 46% biomass bypass to the CHP plant’s boiler over a 25 year period, under average market price conditions (\$4538/ton) the plant is able to recover the initial investment within a PBP of 3.8 years (see appendix A9, B9 and C9 for PBP).

**5.5.1.2 CA 45% bypass (High, Low and Average Product Price)**

From it can be seen that the CA scenario under average market prices of \$1102/ton for a 45 % biomass bypass to the CHP plant provides an IRR of 13% and NPV of \$77 million and therefore the plant is profitable under average and maximum market price conditions. When the lowest market price of \$750/ton is used the plant becomes unprofitable with an IRR of 2% and an NPV of -\$153 million. Therefore the profitability of the CA scenario is highly sensitive to the market price (as was the case for the xylitol scenario mentioned previously). This is discussed further in section 5.6.

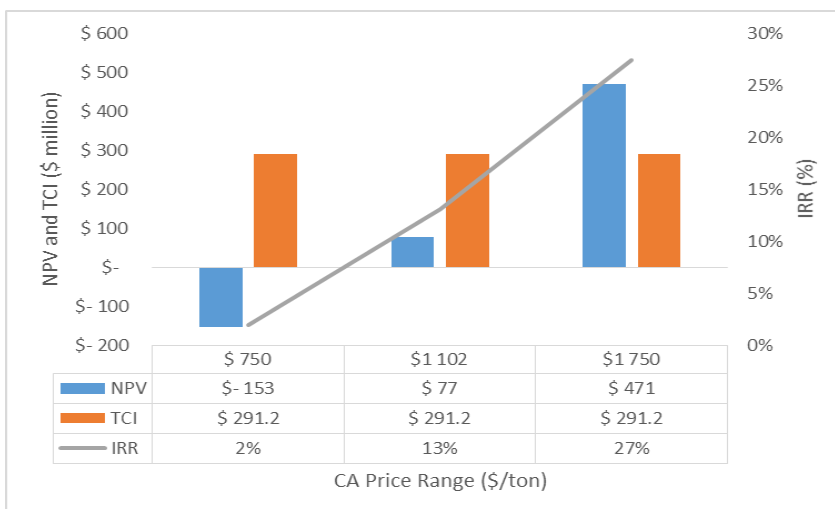


Figure 47 - CA Profitability for different CA Prices (Max, Ave and Min)

Over a 25 year project life span for a 46% biomass bypass and under average market conditions the scenario is only able to return the initial investment after 11.1 years. The CA biorefinery is therefore further reiterated as being unprofitable and a bad investment.

#### 5.5.1.3 GA 35% bypass (High, Low and Average Product Price)

It can be seen that for a 35% biomass bypass that the GA biorefinery is able to produce a positive IRR and NPV under all market price conditions (min., max and average GA prices). For an average market price of \$3625/ton the scenario produces an NPV of \$866 million for a TCI of \$422 million with IRR of 32% (> 9.7% hurdle rate) and therefore the scenario can be considered profitable and a good investment. Even under minimum market prices the biorefinery is still able to provide an IRR of 18%. A maximum market price of \$6500/ton results in a sharp increase in the IRR and NPV (60% and \$2.33 billion), therefore the GA scenario's profitability seems to be highly dependent on market price. The same was proven for xylitol and CA scenarios previously mentioned. Further analysis of profitability with regard to selling price is detailed further in section 5.6.

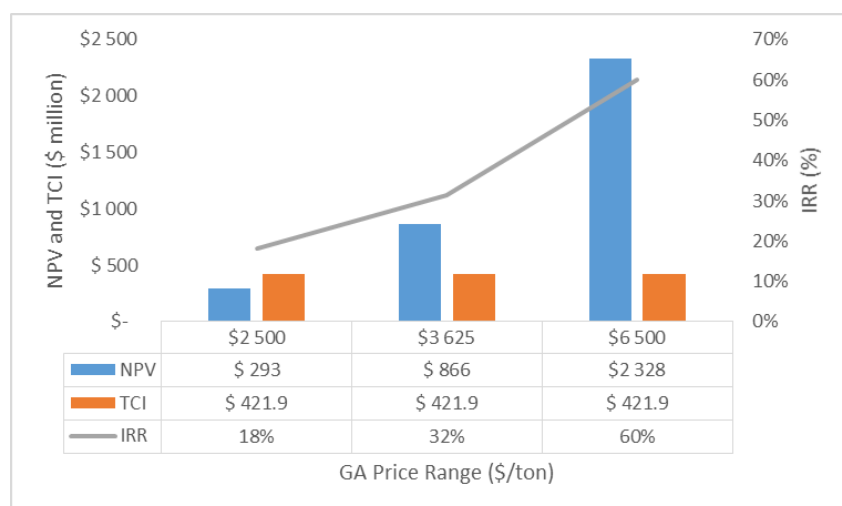


Figure 48 - GA Profitability Range for Max, Ave and Min GA Prices

For a 25 year project life-span, the scenario has a PBP of 3.5 years before the TCI of \$422 million is recovered. The scenario can therefore be reiterated as being a sound investment with a low pay-back time and high profit.

#### 5.5.2 Economic Comparison between different Biorefinery Scenarios

The investigated biorefinery scenarios are further compared to each other as well as the CHP plant baseline as shown in Figure 49. As previously mentioned, when comparing capital expenditure (refer to section 5.4.4), the relative profitability of scenarios can be attributed to a number of key factors, the bypass ratio, the sugar pathway utilized for the product biorefinery (hemicellulosic versus cellulosic) and the technology used for reaction and separation, as well as the product price.

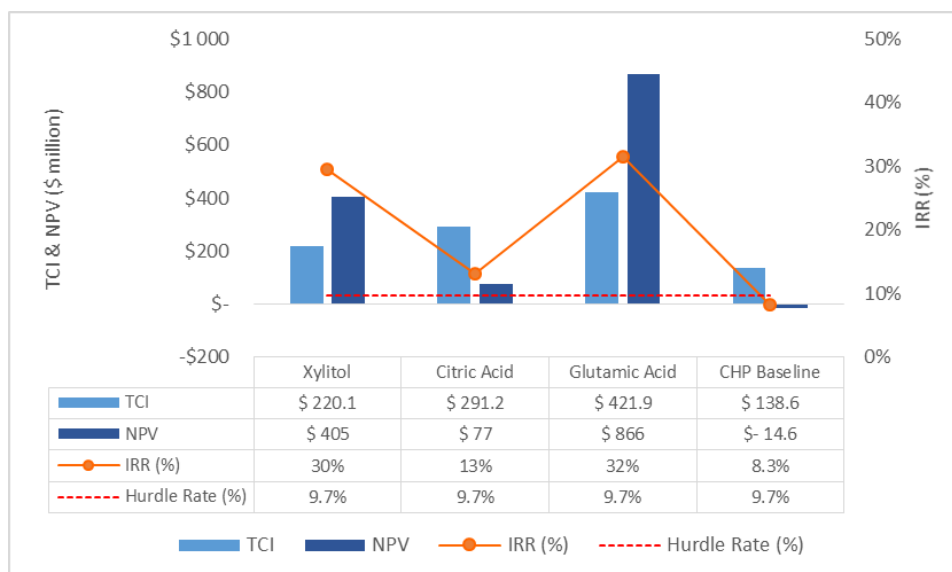


Figure 49 - Comparison of Key Profitability and Investment Indicators among Biorefinery Scenarios

From Figure 49, it can be observed that both the xylitol (IRR of 30% and NPV of \$423 million) and GA (IRR of 32% and NPV of \$866 million) biorefinery scenarios are the most profitable when compared to the CHP plant baseline (IRR of 8.3% and the NPV of -\$14.6 million). The CA biorefinery when compared to the CHP baseline as well as the other product scenarios is not as profitable (IRR of 13% and NPV of \$77 million). It is therefore more profitable to valorise bagasse to bioproducts than to invest in a standalone CHP plant and combust bagasse and trash for steam and electricity revenue. In terms of capital investment, the GA scenario requires the highest capital investment (\$422 million), whereas xylitol requires a lower TCI (\$220 million) than CA (\$291 million) and GA scenarios. The CHP baseline has a lower investment barrier of \$138.6 million than all the product scenarios but exceeds the NPV (-14.6 million) after a 25 year project life span. Therefore from an investment entry point of view (TCI) the xylitol scenario is more favourable, whereas from a profitability perspective (IRR and NPV) the GA is a scenario more enticing investment opportunity.

## 5.6 Cost Sensitivity Analysis of Scenarios

The following section highlights the sensitivity of the biorefinery profitability in terms of IRR to changes in key parameters. Parameters are adjusted by 30% (increased and decreased) and the resulting percentage change to the IRR is observed. In this manner, parameters of key importance can be identified and recommendations can be made to further improve biorefinery profitability.

### 5.6.1 Xylitol Biorefinery Cost Sensitivity Analysis

From Figure 50 it can be observed that the change in IRR for the xylitol biorefinery scenario is most sensitive to changes in xylitol selling price, followed by FCI, TOC, activated carbon (AC), income tax



and catalyst cost. It can be viewed, that the higher the xylitol selling price, the greater the change in IRR whereas conversely, the lower the xylitol selling price, the lower the IRR. Increasing xylitol selling price by 30 %, provides a 43.1% increase in IRR. Naturally the opposite is true for FCI, TOC, activated carbon, income tax and catalyst cost, where higher increases in both CAPEX and OPEX result in lower profitability. If the Raney-Ni catalyst can be partially recycled or a cheaper alternative catalyst of the same efficacy can be used, a significant portion of operating costs can be reduced and allow a more profitable investment scenario.

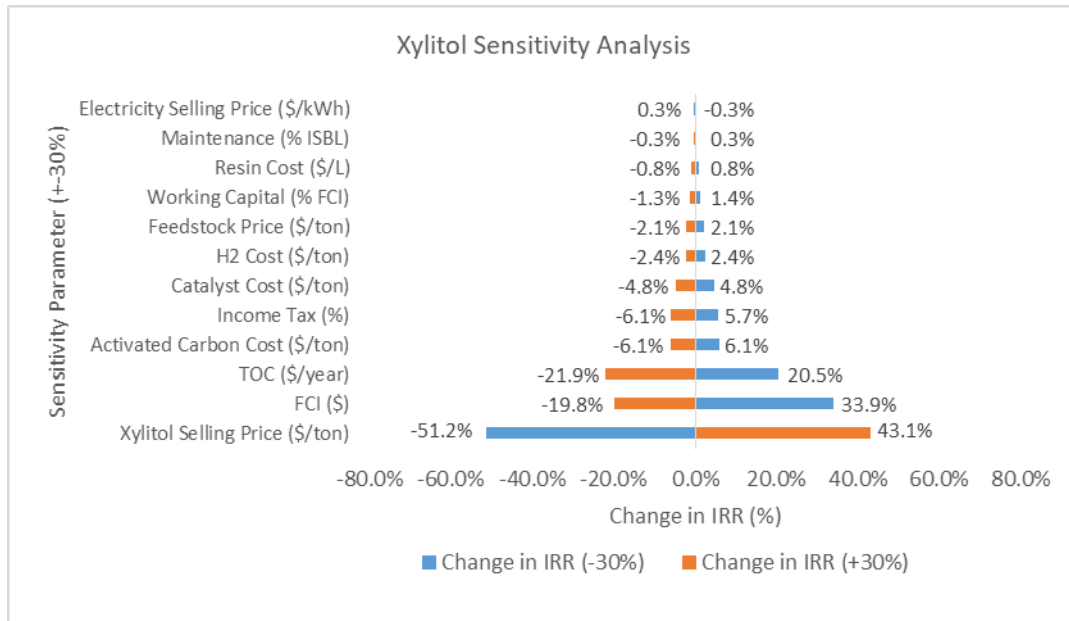


Figure 50 - Xylitol Sensitivity Analysis

### 5.6.2 CA Biorefinery Cost Sensitivity Analysis

Figure 51 summarises the sensitivity of the CA scenario’s profitability (IRR) to adjustments in key parameters. The CA scenario via the solvent extraction route is not profitable and therefore provides very high percentage changes to IRR and some changes are undefinable when profitability is lowered further and therefore a change in IRR (%) from the maximum market price case is provided. In descending order, the highest change in IRR resulted from adjustments to citric acid selling price, FCI, TOC and solvent cost.

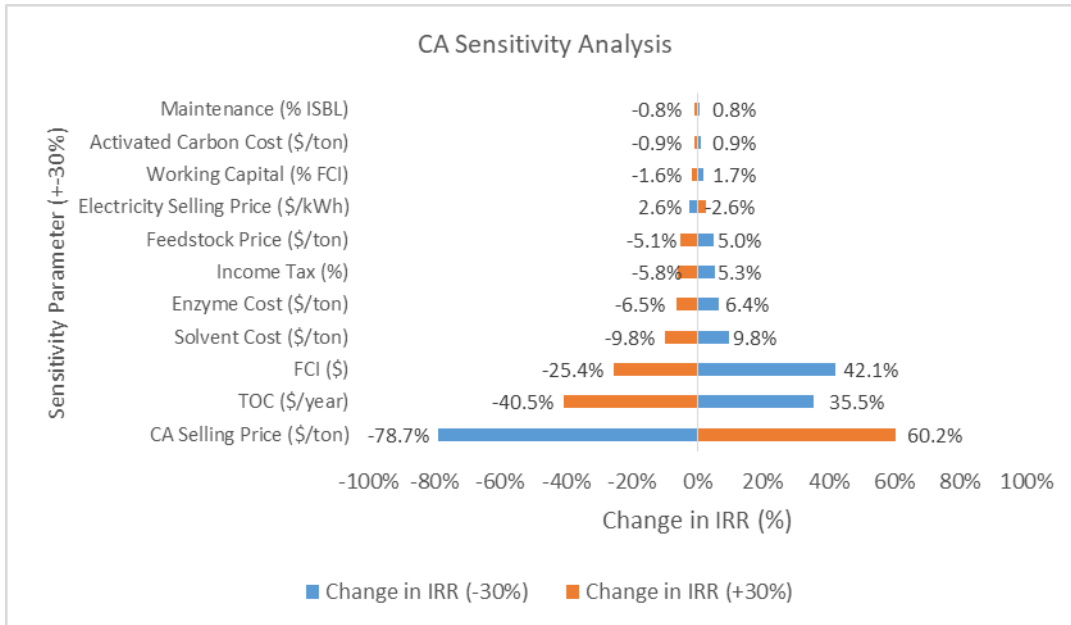


Figure 51 - CA Sensitivity Analysis

### 5.6.3 GA Biorefinery Cost Sensitivity Analysis

The sensitivity of the GA scenario’s profitability in terms IRR to changes in key parameters is displayed in Figure 52. The IRR changes most drastically when altering the glutamic selling price. A 30% increase in glutamic selling price results in a 36.1% increase in IRR. Other notable variables inducing significant change to the IRR include FCI, TOC and enzyme cost. When each of these mentioned variables is increased by 30%, GA IRR conversely reduces by 19.7, 14.2 and 7.3% respectively.

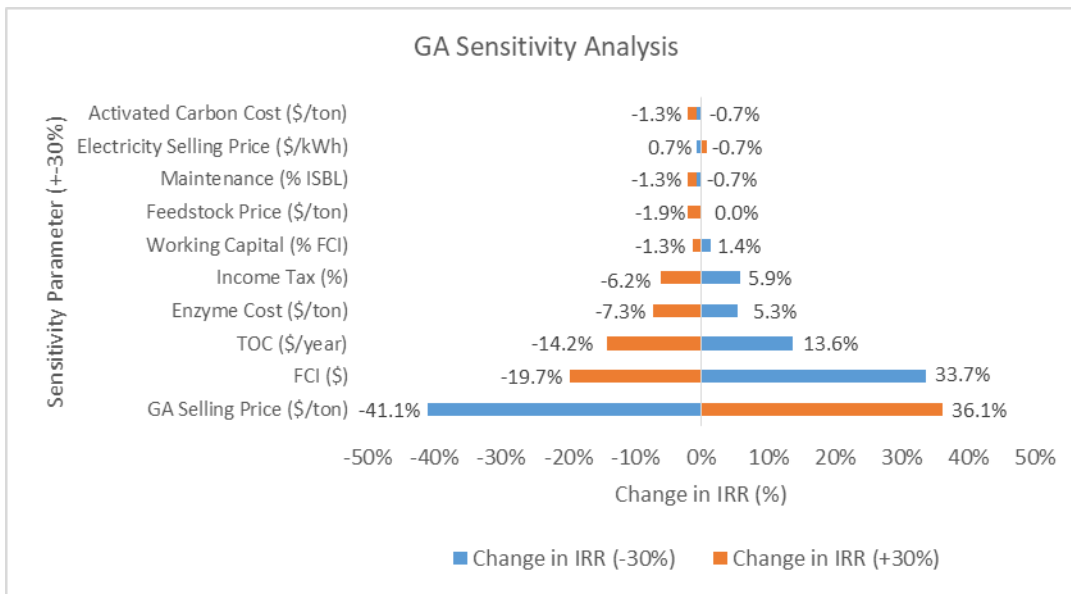


Figure 52 - GA Sensitivity Analysis (% Change in IRR)

#### 5.6.4 Summary

Xylitol, CA and GA scenarios are all most sensitive to adjustments in product selling price, followed by adjustments in FCI and TOC. With regards to xylitol and GA profitability, catalyst and enzyme cost also impart a considerable difference to profitability respectively and therefore reductions in cost through increased catalyst recycling measures or onsite enzyme production can further improve IRR. CA profitability can be improved further by ensuring most of the solvent during extraction and stripping is recycled back to the system. Other recommendations are discussed in Chapter 6.

### 5.7 Environmental Impact of Biorefinery Scenarios

Figure 53 compares the environmental impact of the investigated biorefinery scenarios. It can be seen that the GA biorefinery overall is the least environmentally friendly product scenario, consuming the highest quantity of water (4.3 million ton/annum) and producing the most non-biogenic CO<sub>2</sub> emissions (0.097 million tons/annum) and solid waste (3612 tons/year), out of all scenarios investigated. In contrast, the CHP baseline is the most environmentally friendly scenario from a biogenic carbon emissions perspective, producing zero non-biogenic CO<sub>2</sub> emissions per annum with the lowest water consumption (2.1 million tons/year). This is due the higher quantity of carbon containing biomass being sent to the boiler as opposed to the product biorefineries who convert some of the biomass fed to the resulting products. Therefore the greater the biomass bypass to the CHP boiler the more environmentally friendly the scenario becomes. The xylitol biorefinery has the lowest non-biogenic emissions out of all product biorefineries (0.003 million tons/year). This can be explained by acknowledging that this scenario has the highest biomass bypass (46%) to the CHP plant boiler than other product scenarios. In addition, this is further augmented by the cellulignin combusted within the CHP plant as opposed to the lignin combusted in the CA and GA scenarios (lower input carbon, results in lower output carbon emissions). Therefore one can deduce that biorefineries deriving value from hemicellulose portions of lignocellulose, in comparison to those deriving value from cellulose are able to produce more carbon neutral emissions and therefore more readily meet environmental regulations.

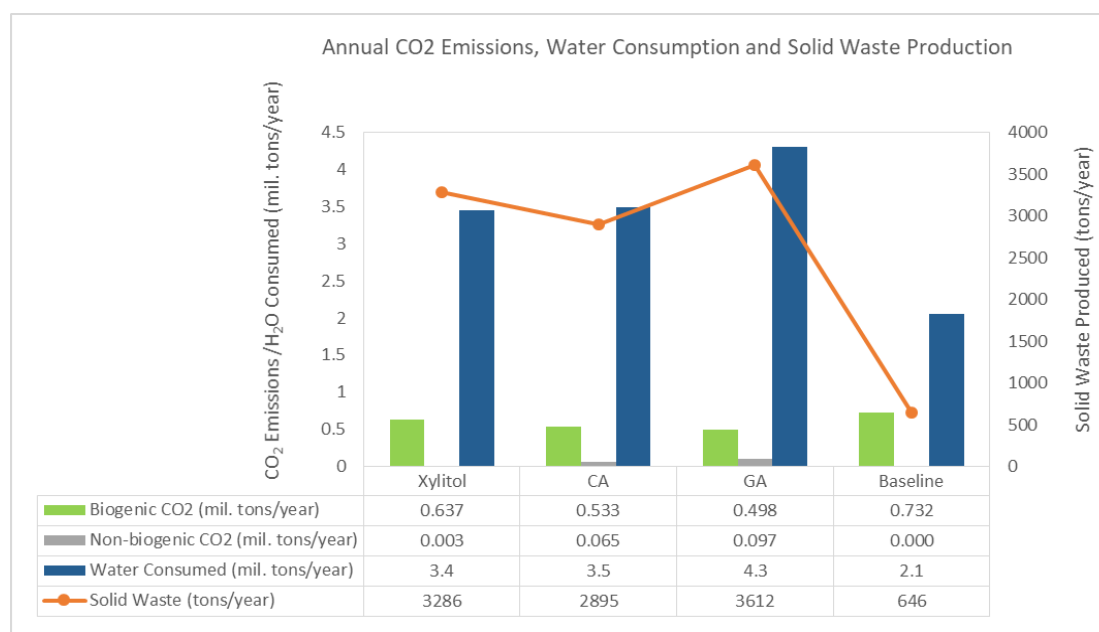


Figure 53 - Environmental Impact of Biorefinery Scenarios (Carbon Emissions, Solid Waste and Water Consumption)

When comparing scenario profitability in terms of IRR versus environmental impact as is defined in this present work, it can be seen that high profitability is correlated with high environmental impact as is the case for the GA and electricity co-production scenario. This could be attributed to the lower bypass of biomass to the CHP boiler and the greater availability for GA production in comparison to the other biorefinery scenarios. However, the greater the amount of feedstock treated, the greater the chemical dosage and expenses. The xylitol scenario achieves a good balance between high profitability (IRR of 30%) and low environmental impact (lowest water consumption and non-biogenic emissions, excluding the CHP baseline). The same conclusion is evident when reviewing the CA scenario, which achieves the second lowest profitability (IRR, 13%) and second highest environmental impact. A profitability-environmental composite score was calculated by combining weighted normalised data points for profitability (IRR %) and environmental factors as was done previously in the present work when determining desirability scores to help select the most suitable microorganisms for CA and GA production (see Table 63). Solid waste, water consumption and non-biogenic emissions were normalised and weighted to determine the environmental impact score. 80% and 20% weighting was used for profitability and environmental impact when determining the composite score respectively.

Table 63 – Weighted profitability and environmental composite score

Scenario	Profitability (IRR, 0-1)	Environmental Impact Score (0-1)	Weighted Composite Score (0-1)
Xylitol	0,91	0,49	0,83
CA	0,19	0,33	0,22
GA	1,00	0,00	0,80
CHP Baseline	0,00	1,00	0,20

It can be observed that GA and xylitol achieve the highest composite scores indicating that these scenarios achieve good profitability without compromising too heavily on environmental impact. CA conversely achieves the lowest score as it is both unprofitable and has the second highest environmental impact.

## 5.8 Viability of Results in Comparison to Existing Facilities

### 5.8.1 Biorefinery Capacity and Market Representation

Although results for most of the investigated scenarios look profitable, the viability needs to be assessed and compared to current production facilities around the world. Currently, data on typical production facility sizes for GA and xylitol is to the best of the author's knowledge unavailable, however a few of the largest CA production facilities are listed in Table 64. When comparing world production to the currently investigated 97 893 tons of CA per annum facility, it can be seen that the largest facility owned by Jungbunzlauer in Austria is able to produce 136 000 tons CA/year (SRI Consulting, 2000). The smallest facility listed in Table 64, is in Ireland (45 000 tons/year). The present CA scenario's capacity falls within this range and the facility can therefore be stated to be within realistic bounds but within a high economy of scale capacity.

Table 64 - Citric Acid Production Facilities from Different Locations ( (SRI Consulting, 2000) (Ciriminna, et al., 2017))

Country	Company	Location	Capacity ('000s tons/year)
Austria	Jungbunzlauer	Perhofen, Austria	136
United States	Archer Daniels Midland	Southport, NC, US	98
Ireland	Archer Daniels Midland	Ringaskiddy, Co Cork, Ireland	45
Belgium	Hoffmann-La Roche	Tienen, Belgium	90
United States	Cargill	Eddyville, Iowa, US	72.6
United States	Tate & Lyle	Dayton, Ohio, US	60
South Africa	Present work	KZN, SA	97.9

The biorefinery scenario production capacities are compared to total world demand per annum in Figure 54 and Table 65. It can be seen that the current xylitol facility functions within a very high operating capacity of 38 789 tons xylitol per year, representing 20% of current world demand (Grand View Research, 2017). This could mean that the xylitol scenario investigated, although profitable, may only be profitable due to the higher economy of scale achieved by processing a larger quantity of feed stock, whereas CA and GA scenarios in contrast represent a smaller 5% and 3% of current world demand respectively. As previously mentioned for the CA scenario, the scale is reiterated as being very large but within realistic values (refer to Table 65).

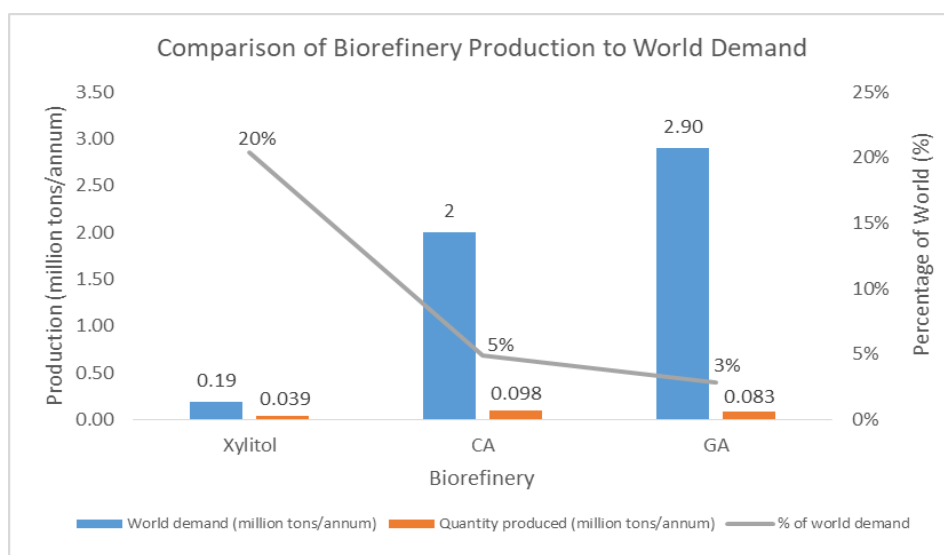


Figure 54 - Representation of Current Production to Total World Demand

Table 65 - Biorefinery Market Representation

Biorefinery	World demand (tons/annum)	Present Work (tons/annum)	% of world demand
<b>Xylitol</b>	190000 <sup>a</sup>	38789	20%
<b>CA</b>	2000000 <sup>b &amp; c</sup>	97893	5%
<b>GA</b>	2900000 <sup>d</sup>	83005	3%

a - (Grand View Research, 2017), b - (Expert Market Research, 2017), c - (Ciriminna, et al., 2017), d - (Global Market Insights, Inc., 2016)

### 5.8.2 Techno-economic Comparison to Previous Studies

Previous techno-economic studies have utilised similar pretreatment technologies and property methods as the currently investigated xylitol biorefinery scenario as seen in Table 66 (predominantly DA pretreatment and the NRTL property method respectively). All feedstocks comprise one or other form of lignocellulosic biomass, however operate under vastly different scales of capacity and different production routes (catalytic and fermentative). When comparing xylitol productivity and cost per ton of xylitol, one can see that the current results fit between the works for fermentative production conducted by Hernández, et al. (2014) with a maximum production cost of \$ 3.61 per ton of xylitol produced and Mussatto, et al. (2013) with a minimum of 1.36 \$/ton xylitol produced. Furthermore Mountraki et al. (2017) showed that the catalytic hydrogenation pathway using a Raney-nickel catalyst was far more profitable than the fermentative production route (higher NPV and lower PBP). One can therefore say that fermentative technology has the potential to become profitable in the future however catalytic production will continue to remain superior in terms of profitability. However, in order to confidently conclude with certainty that catalytic production is more favourable, further techno-economic studies would need to be conducted on different feedstocks as each lignocellulosic biomass has varying quantities of hemicellulose which can further cause variations in profitability. Unfortunately to the best of the authors knowledge no other conclusive studies that

incorporate detailed techno-economic analysis utilising discounted cash flow (DCF) analysis are present in academic literature and therefore a thorough comparison for different feedstocks utilising standardised pretreatment and economic calculations is a topic for discussion in future work.

Table 66 - Economic Comparison of Present Xylitol Scenario to Past Studies

Parameter	Present Work	(Mussatto, et al., 2013)	(Hernández, et al., 2014)	(Mountraki, et al., 2017)	(Mountraki, et al., 2017)
<b>Feedstock</b>	Sugarcane Bagasse and Trash	BSG	Olive Stone	C5 sugar syrup	C5 sugar syrup
<b>Pretreatment Method</b>	DA	DA	DA	Organosolv	Organosolv
<b>Process Type</b>	Hydrolysis(H <sub>2</sub> SO <sub>4</sub> ) Catalytic Hydrogenation	Hydrolysis(H <sub>2</sub> SO <sub>4</sub> ) Fermentation	Hydrolysis(H <sub>2</sub> SO <sub>4</sub> ) Fermentation	Catalytic Hydrogenation	Fermentation
<b>Aspen Thermodynamic Model</b>	ELEC-NRTL with STEAM NBS	NRTL	NRTL with Hayden-O'Connell	NRTL	NRTL
<b>Feedstock Capacity (tons/h)</b>	113.5 (61.3 after bypass)	100	10	4.13	4.13
<b>Moisture Content (%)</b>	43	45	7.71	30	30
<b>IRR (%)</b>	30	N/A	N/A	N/A	N/A
<b>PBP</b>	5.3	N/A	N/A	1.1	13.9
<b>NPV (\$ mil./year)</b>	405.4	N/A	N/A	77.6	-59.8
<b>TCI (\$ million)</b>	220.1	N/A	N/A	17.8	83.88
<b>Price Used (\$/ton)</b>	4538	2950	Not mentioned	2990	2990
<b>Xylitol Production Cost (\$/kg)</b>	2.04	1.36	3.61	2.33	3.07
<b>Xylitol Productivity (ton/h)</b>	5.90	10.38	0.5	3.972	3.414
<b>ton xylitol/ton feedstock</b>	0.05	0.104	0.050	0.962	0.827
<b>TOC (\$ mil./year)</b>	79.2	92.74	11.86	60.86	68.92

When comparing GA production costs utilising membrane integrated reactor systems from Pal, et al., (2016; 2015) to the present study, it can be seen that the present work displays more than 3 times lower production costs (1.249 \$/kg GA compared to 3.936 \$/kg GA) (refer to Table 67). This could be attributed to the higher production capacities investigated in the present work. In addition, 2<sup>nd</sup> generation bagasse and trash are available in large quantities and in close proximity to the biorefinery site allowing cost-effective access to feed and therefore favourable operating costs. The present GA-electricity co-production scenario is able to meet its energy requirements by diverting a portion of the raw feed to the CHP plant's boiler for electricity and steam production. In contrast Pal, et al. (2016) reported \$40 000/annum in electricity costs. Therefore GA production using 2<sup>nd</sup> generation feedstock in the present study shows an operating cost advantage. There is however a trade-off between total operating cost and total capital investment. In order for the GA-electricity scenario to exhibit lower operating costs, high amounts of capital is required. For the present work in addition to the expense

for fermentation, costs are distributed towards pretreatment, enzymatic hydrolysis and the CHP plant for lignocellulose valorisation. Capital investment costs are more than 10 times greater for the present work compared to that conducted by Pal et al. (2016) for 1<sup>st</sup> generation sugarcane juice feedstock. Therefore the production of GA from 2<sup>nd</sup> generation feedstock has the benefit of being produced at lower operating costs when compared to 1<sup>st</sup> generation feedstock, but requires high initial capital investment.

Table 67 - GA Economic Comparison to Past Studies (Pal, et al., 2015)

Parameter	Present Work	(Pal, et al., 2016)
<b>Feedstock</b>	Sugarcane Bagasse and Trash	Cane sugar juice
<b>Feedstock type</b>	2 <sup>nd</sup> - generation	1 <sup>st</sup> - generation
<b>Feedstock cost (\$/ton)</b>	10.77	156.25
<b>Plant Capacity (tons/year)</b>	83005	1550
<b>TOC (\$ mil./year)</b>	109.1	6.1
<b>Production Cost (\$/kg GA)</b>	421.93	0.66
<b>TCI (\$ million)</b>	1.31	3.94
<b>Investment Cost (\$/kg GA/year)</b>	5.08	0.426



## 6. Conclusion

A techno-economic analysis of xylitol, CA and GA biorefineries, utilising sugarcane lignocellulosic feedstock with combined electricity and steam production, annexed to existing sugar mills was conducted to determine the profitability of these scenarios. The profitability of each scenario was characterised using financial indicators (IRR, NPV, pay-back period and TCI) and compared to a baseline combined heat and power (CHP) plant to investigate whether more value could be gained merely burning bagasse in boilers solely for electricity production or for the production of bioproducts.

It was determined that feedstock, reaction and separation technology, biomass bypass to the CHP boiler and the reaction pathways for different sugar pathways (hemicellulosic and cellulosic utilisation) had a high impact on profitability and costs. Higher bypass of biomass to the CHP boiler would result in greater throughput to the biorefineries resulting in increased operating and capital expenditure but also increased revenue. GA production using 2<sup>nd</sup> generation feedstock in the present study shows an operating cost advantage but with a higher compromised TCI.

Xylitol, CA and GA product scenarios all gave IRR's in excess of the CHP baseline scenario as well as the hurdle rate of 9.7% (8.3% IRR for 100% biomass bypass). It is therefore more economically profitable to convert 2<sup>nd</sup> generation biomass to bioproducts than to combust biomass in boilers solely for electricity production. The profitability of all biorefinery scenarios are highly dependent on 3 main variables, namely the product selling price, FCI and the TOC. In addition to these variables, the xylitol biorefinery profitability was significantly affected by catalyst price, whereas GA and CA profitability were sensitive to enzyme and solvent costs. Increasing the catalyst recovery or producing enzymes onsite would significantly lower costs.

In terms of environmental impact, the GA scenario consumed the most process water and produced the highest quantity of solid waste and non-biogenic carbon emissions. Whereas the CHP baseline scenario produced no non-biogenic carbon dioxide emissions. Although investigated xylitol and GA scenarios were deemed profitable, the current production scale of the xylitol biorefinery (38 789 tons/annum) was analysed and found to represent 20% of the world's total xylitol demand, indicating that the profitability of this scenario could be attributed to high production capacities. The same was found true for GA and CA production (however to a lesser extent), with each scenario representing 3% and 5% of global demand.

## 7. Recommendations and Future Scope

When considering the operation hours for the 9 month harvest-storage cycle for which biorefinery scenarios are investigated (6480 hours), an extension in operation to a full 12 months could dramatically improve profitability. This could be achieved by supplementing biorefineries with another widely available feedstock produced from the paper and pulp industry, namely wood chips. This feedstock could be sourced from saw mills in surrounding wood plantations that are typically in the vicinity of traditional sugar cane growing regions in South Africa (KwaZulu-Natal and Mpumalanga provinces) (AgriSA, 2016). However supplementation of feedstock with wood chips does warrant a few considerations. The most prominent consideration is the polysaccharide content and structure in hardwoods versus agricultural residues. The hemicellulose and lignin content is typically higher in this form of lignocellulosic feedstock compared to sugarcane bagasse (30+-5% versus the 27.1% hemicellulose and 20+-4% versus 21.9% lignin respectively as used in the present work) (Petersen, et al., 2014); (Saini, et al., 2015)). This has 2 main effects on production, a higher quantity of lignin for energy and steam in the CHP boiler and a lower quantity of cellulosic glucose available for conversion to CA and GA respectively. Therefore the supplementation of wood chips would be more applicable to the xylitol-electricity co-production scenario, where the hemicellulosic chain sugars are valorised and where pretreatment conditions would not need to be dramatically altered. If operation was extended to a full year for the xylitol biorefinery (8760 versus 6480 hours), profitability would improve by an estimated 48% (from IRR of 30% to 44.5%).

Future scope should focus on the following areas:

- The use of different detoxification techniques in biorefinery scenarios (e.g. activated carbon adsorption, ion-exchange) should be investigated in more detail to determine its effect on profitability and determine which detoxification routes induce low operating and capital expense, while minimising sugar losses.
- The profitability of different purification techniques for CA production e.g. precipitative purification routes compared to solvent extraction routes should be conducted to assess the relative costs of each and determine which is more financially viable while imparting a low environmental impact
- Compare the profitability of precipitative purification routes to membrane purification routes for GA production to determine the cost and environmental benefit
- The effect of the solid to liquid ratio on the profitability of integrated biorefineries should be investigated, as a large portion of the steam and electricity that is produced was used to remove water

- The use of excess electricity for hydrogen production via electrolysis or steam methane reformation for the xylitol scenario should be investigated to potentially further improve the economic outlook of this biorefinery. A portion of the hydrogen produced could be used to offset the chemical operating costs of the biorefinery or in the case of excess hydrogen production, a part of the gas could be sold to other industries for additional revenue at a higher cost than the from the sale of electricity.
- Utilising alternative by-products for revenue streams such as arabinitol and oxalic acid for xylitol and CA scenarios should be assessed and market maturity should be accounted for

## 8. References

- [1] Agbor, V. B. et al., 2011. Biomass pretreatment: Fundamentals toward application. *Biotechnology Advances*, December, 29(6), pp. 675-685.
- [2] AgriSA, 2016. *A Raindrop in The Drought: Agri SA's status report on the current drought crisis*, s.l.: AgriSA.
- [3] Akobi, C., Hafez, H. & Nakhla, G., 2017. Impact of furfural on biological hydrogen production kinetics from synthetic lignocellulosic hydrolysate using mesophilic and thermophilic mixed cultures. *International Journal of Hydrogen Energy*, 27 April, 42(17), pp. 12159-12172.
- [4] Alberts, A. & Rothenberg, G., 2012. *Process for preparing foamed polymer*. Netherlands, Patent No. WO 2012052385.
- [5] Alibaba, 2016. *actory direct supply New product Ruthenium powder price*. [Online] Available at: [http://www.alibaba.com/product-detail/Factory-direct-supply-New-product-Ruthenium\\_1833251833.html?spm=a2700.7724838.0.0.GXVAXN&s=p](http://www.alibaba.com/product-detail/Factory-direct-supply-New-product-Ruthenium_1833251833.html?spm=a2700.7724838.0.0.GXVAXN&s=p) [Accessed May 2016].
- [6] Alibaba, 2016. *High quality raney nickel catalyst*. [Online] Available at: [http://www.alibaba.com/product-detail/High-quality-raney-nickel-catalyst\\_60132154494.html?spm=a2700.7724838.0.0.9IH6q](http://www.alibaba.com/product-detail/High-quality-raney-nickel-catalyst_60132154494.html?spm=a2700.7724838.0.0.9IH6q) [Accessed May 2016].
- [7] Alvira, P., Tomás-Pejó, E., Ballesteros, M. & Negro, M., 2010. Pretreatment technologies for an efficient bioethanol production process based on enzymatic hydrolysis: A review. *Bioresource Technology*, July, 101(13), pp. 4851-4861.
- [8] Amin, G. & Al-Talhi, A., 2007. Production of L-glutamic Acid by Immobilized Cell Reactor of the Bacterium *Corynebacterium glutamicum* Entrapped into Carrageenan Gel Beads. *World Applied Sciences Journal*, 2(1), pp. 62-67.
- [9] Anastassiadis, S., Aivasidis, A. & Wandrey, C., 1993. *Process for the continuous production of citric acid by fermentation*. Germany, Patent No. P 43174884-09.
- [10] Anastassiadis, S. & Rehm, H.-J., 2006. Citric acid production from glucose by yeast *Candida oleophila* ATCC 20177 under batch, continuous and repeated batch cultivation. *Electronic Journal of Biotechnology*, 15 January, 9(1), pp. 27-39.
- [11] Angumeenal, A. & Venkappayya, D., 2013. An overview of citric acid production. *LWT - Food Science and Technology*, March , 50(2), pp. 367-370.
- [12] Avellar, B. & Glasser, W., 1998. Steam-assisted biomass fractionation. I. Process considerations and economic evaluation.. *Biomass Bioenergy*, Volume 14, p. 205–218.
- [13] Baek, S.-C. & Kwon, Y.-J., 2007. Optimization of the pretreatment of rice straw hemicellulosic hydrolyzates for microbial production of xylitol. *Biotechnology Bioprocess Engineering*, Volume 12, p. 404–419.
- [14] Bahndorf, D. & Kienle, U., 2004. *World Market of Sugar and Sweeteners*, Leinfelden-Echterdingen: International Association for Stevia Research e.V..

- [15] Baniel, A., Vitner, A., Gonen, D. & Heidel, D., 2003. *Process for the recovery of citric acid*. United States of America, Patent No. US7411090 B2.
- [16] Bao, L., Bai, P. & Guo, X., 2014. Renewable Chemicals: Production or Purification by Adsorption Process. *Dispersion Science and Technology*, 15 October, 36(9), pp. 1297-1308.
- [17] Barakat, A., Monlau, F., Steyer, J. & Carrere, H., 2012. Effect of lignin-derived and furan compounds found in lignocellulosic hydrolysates on biomethane production. *Bioresource Technology*, January, Volume 104, pp. 90-99.
- [18] Baudel, H., de Abreu, C. & Zaror, C., 2005. Xylitol production via catalytic hydrogenation of sugarcane bagasse dissolving pulp liquid effluents over Ru/C catalyst. *Journal of Chemical Technology Biotechnology*, Volume 80, pp. 230-233.
- [19] Benjamin, Y., Cheng, H. & Görgens, J., 2013. Evaluation of bagasse from different varieties of sugarcane by dilute acid pretreatment and enzymatic hydrolysis. *Industrial Crops and Products*, November, Volume 51, pp. 7-18.
- [20] Benjamin, Y., Garcia-Aparicio, M. P. & Gorgens, J. F., 2014. Impact of cultivar selection and process optimization on ethanol yield from different varieties of sugarcane. *Biotechnology for Biofuels*, 12 April.7(60).
- [21] Bensah, E. C. & Mensah, M., 2013. Chemical Pretreatment Methods for the Production of Cellulosic Ethanol: Technologies and Innovations. *International Journal of Chemical Engineering*, Volume 2013, p. 21.
- [22] Bensah, E. C. & Mensah, M., 2013. Chemical Pretreatment Methods for the Production of Cellulosic Ethanol: Technologies and Innovations. *International Journal of Chemical Engineering*, 27 August. p. 21.
- [23] Berovic, M. & Legisa, M., 2007. Citric acid production. *Biotechnology annual review*, February, Volume 13, pp. 303-343.
- [24] Bhayani, B. V. & Ramarao, B. V., 2013. Filtration-Based Separations in the Biorefinery. In: *Separation and Purification Technologies in Biorefineries*. Chichester: John Wiley & Sons, Ltd, pp. 327-349.
- [25] Bidy, M. J., Scarlata, C. & Kinchin, C., 2016. *Chemicals from Biomass: A Market Assessment of Bioproducts with Near-Term Potential*, Denver: SciTech Connect.
- [26] Bilitewski, B., Härdtle, G. & Marek, K., 1997. *Waste Management*. Berlin: Springer.
- [27] Binder, S., 2013. In: *Rapid Development of Small-Molecule producing Microorganisms based on Metabolite Sensors*. s.l.:s.n.
- [28] Blunk, S. & Jenkins, B., 2000. *Combustion Properties of Lignin Residue from Lignocellulose Fermentation*, Davis: National Renewable Energy Laboratory.
- [29] Boerjan, W., Ralph, J. & Baucher, M., 2003. Lignin biosynthesis. *Annu. Rev. Plant Biol.* 54, p. 519–549.
- [30] Bowtell, J. et al., 1999. Effect of oral glutamine on whole body carbohydrate storage during recovery from exhaustive exercise. *Journal of Applied Physiology*, June, 86(6), pp. 1770-1777.
- [31] Briffaud, J. a. E. M., 1979. Citric acid production from glucose | Growth and excretion kinetics in a stirred fermentor. *Biotechnology and Bioengineering*, 21(11), pp. 2082-2092.

- [32] Canilha, L., Carvalho, W., Felipe, M. & Silva, J., 2008. Xylitol production from wheat straw hemicellulosic hydrolysate: hydrolysate detoxification and carbon source used for inoculum preparation. *Brazilian Journal of Microbiology*, Volume 39, pp. 333-336.
- [33] Canilha, L. et al., 2012. Bioconversion of Sugarcane Biomass into Ethanol: An Overview about Composition, Pretreatment Methods, Detoxification of Hydrolysates, Enzymatic Saccharification, and Ethanol Fermentation. *Journal of Biomedicine and Biotechnology*, October, 2012(989572), p. 15.
- [34] Canteri-Schemin, M., Ramos, F. H., Waszczynskyj, N. & Wosiacki, G., 2005. Extraction of pectin from apple pomace. *Brazilian Archives of Biology and Technology*, March, 48(2), pp. 259-266.
- [35] Cao, G. et al., 2009. Acid hydrolysis of cornstover for biohydrogen production using *Thermoanaerobacterium thermosaccharolyticum*. *International Journal of Hydrogen Energy*, Volume 34, p. 7182–7188.
- [36] Carvalheiro, F., Duarte, L. & Gírio, F., 2008. Hemicellulose biorefineries: a review on biomass pretreatments. *Journal of Scientific & Industrial Research*, Volume 67, pp. 849-864.
- [37] Carvalheiro, F. et al., 2005. Evaluation of the detoxification of brewery's spent grain hydrolysate for xylitol production by *Debaryomyces hansenii* CCM1 941. *Process Biochemistry*, March, 40(3-4), pp. 1215-1223.
- [38] Carvalho, R. N. L. d., 2009. Dilute acid and enzymatic hydrolysis of sugarcane bagasse for biogas production. *Department of Biotechnology of Lund's Faculty of Engineering*.
- [39] Chandel, A. K. et al., 2012. Dilute Acid Hydrolysis of Agro-Residues for the Depolymerization of Hemicellulose: State-of-the-Art. In: S. S. d. Silva & A. K. Chandel, eds. *D-Xylitol - Fermentative Production, Application and Commercialization*. 1 ed. São Paulo: Springer Science + Business Media, pp. 39-61.
- [40] Chandel, A., Kapoor, R., Singh, A. & Kuhad, R., 2007. Detoxification of sugarcane bagasse hydrolysate improves ethanol production by *Candida shehatae* NCIM 3501. *Bioresources Technology*, Volume 98, p. 1947–1950.
- [41] Chandel, A. K., Kapoor, R. K., Singh, A. & Kuhad, R. C., 2007. Detoxification of sugarcane bagasse hydrolysate improves ethanol production by *Candida shehatae* NCIM 3501. *Bioresource Technology*, July, 98(10), pp. 1947-1950.
- [42] Chandel, A. K. & Silva, S. S. d., 2013. In: A. K. Chandel & S. S. d. Silva, eds. *Sustainable Degradation of Lignocellulosic Biomass - Techniques, Applications and Commercialization*. Rijeka, Croatia: InTech, pp. 15-47.
- [43] Chandel, A. K., Silva, S. S. & Singh, O. V., 2011. Detoxification of lignocellulosic hydrolysates for improved bioethanol production. In: M. A. S. Bernardes, ed. *Biofuel Production-Recent Developments and Prospects*. Rijeka, Croatia: In Tech, p. 225–246.
- [44] Ciriminna, R., Meneguzzo, F., Delisi, R. & Pagliaro, M., 2017. Citric acid: emerging applications of key biotechnology industrial product. *Chemistry Central Journal*, 8 March.11(22).
- [45] Clauser, N. et al., 2016. Small-sized biorefineries as strategy to add value to sugarcane bagasse. *Chemical Engineering Research and Design*, Volume 107, pp. 137-146.
- [46] Cooke, P., 2013. *Growth Cultures: The Global Bioeconomy and its Bioregions*. London and New York: Routledge.

- [47] Currie, J. N., 1917. The citric acid fermentation of *A. niger*. *J. Biol. Chem.* 1917 vol.31:5.
- [48] Datta, R. & Bergemann, E. P., 1996. *Process for producing of citric acid and monovalent citrate salts*. United States of America, Patent No. US 5532148 A.
- [49] Davis, S., Foxon, K. & Booyesen, K., 2015. *Overview of the SMRI New Products Strategy: a staged screening approach*. Durban, Sugar Milling Research Institute NPC.
- [50] Delaunay, S. et al., 2002. *Microbiol. Bio-technol.* 28. pp. 333-337.
- [51] Dias, M. O. C. M. P. et al., 2011. Second generation ethanol in Brazil: Can it compete with electricity production?. *Bioresource Technology*, October, 102(19), pp. 8964-8971.
- [52] Diedericks, D., van Rensburg, E. & Gorgens, J. F., 2013. Enhancing sugar recovery from sugarcane bagasse by kinetic analysis of a two-step dilute acid pretreatment process. *Biomass and Bioenergy*, 1 August, Volume 57, pp. 149-160.
- [53] Dinesh Kumar Mishra, A. A. D. J.-S. H., 2013. Ruthenium nanoparticles supported on zeolite Y as an efficient catalyst for selective hydrogenation of xylose to xylitol. *Journal of Molecular Catalysis A: Chemical*, Volume 376, pp. 63-70.
- [54] Dussán, K. J. et al., 2014. Dilute-acid Hydrolysis of Cellulose to Glucose from Sugarcane Bagasse. *Chemical Engineering Transactions*, Volume 38, pp. 433-438.
- [55] Eggeman, T. & Elander, R., 2005. Process and economic analysis of pretreatment technologies. *Bioresources Technology*, 96(18), p. 2019–2025.
- [56] Eichman, J., Townsend, A. & Melaina, M., 2016. *Economic Assessment of Hydrogen Technologies Participating in California Electricity Markets*, Golden: NREL.
- [57] Enzminger, J. a. A. J., 1986. Use of cell recycle in the aerobic fermentative production of citric acid by yeast. *Biotechnology Letters*, 8(1), pp. 7-12.
- [58] Esser, K. & Bennett, J., 2002. *The Mycota: A Comprehensive Treatise on Fungi as Experimental Systems for Basic and Applied Research*. In: New York: Springer-Verlag Berlin Heidelberg, pp. 213-224.
- [59] Etzkorn, B., 2012. *Data Normalization and Standardization*. [Online] Available at: <http://themodelmill.com/wp-content/uploads/2011/11/Data-Normalization-and-Standardization.pdf> [Accessed July 2016].
- [60] Expert Market Research, 2017. *Global Citric Acid Market Share, Size, Price, Trends, Industry Report and Outlook 2017-2022*, New York: Expert Market Research.
- [61] Eyer, K., Oeggerli, A. & Heinzle, E., 1995. On-line gas analysis in animal cell cultivation:II. Methods of oxygen uptake rate estimation and its application to controlled feeding of glutamine.. *Biotechnol.Bioeng.*, Volume 45, pp. 54-62.
- [62] Farzad, S. et al., 2017. Multi-product biorefineries from lignocelluloses: a pathway to revitalisation of the sugar industry?. *Biotechnology for Biofuels*, 11 April.10(87).
- [63] Fisher, A. B. & Fong, S. S., 2014. Lignin biodegradation and industrial implications. *AIMS Bioengineering*, 1(2), pp. 92-112.
- [64] Foster, K.-C. C. a. J. W., 1959. A glutamic acid producing bacillus. *Journal of Bacteriology*, pp. 715-725.

- [65] Gírio, F. et al., 2010. Hemicelluloses for fuel ethanol: a review. *Bioresource Technology*, July, 101(13), pp. 4775-4800.
- [66] Global Market Insights, Inc., 2016. *Glutamic Acid and Monosodium Glutamate (MSG) Market Size, Potential, Industry Outlook, Regional Analysis Application Development, Competitive Landscape & Forecast, 2016 - 2023*, Delaware : s.n.
- [67] globalsweet, 2010. *A Brief History of Xylitol*. [Online]  
Available at: <https://www.globalsweet.com/merchant2/merchant.mvc?Screen=XYHI>
- [68] Görgens, J. et al., 2015. *A Biorefinery Approach to Improve The Sustainability of The South African Sugar Industry*, Stellenbosch: Green Fund.
- [69] Görgens, J. et al., 2015. *A Biorefinery Approach to Improve The Sustainability of The South African Sugar Industry*, Stellenbosch: Green Fund.
- [70] Grand View Research, 2017. *Xylitol Market Analysis By Application (Chewing Gum, Confectionery, Bakery, Oral Care), By Region (North America, Europe, Asia Pacific, Latin America, MEA), And Segment Forecast, 2014 - 2025*, s.l.: Grand View Research.
- [71] Grewal, H. & Kalra, .., 1995. Fungal production of citric acid. *Biotechnology Advances*, 13(2), pp. 209-234.
- [72] Grumezescu, A. M. & Holban, A. M. eds., 2017. *Food Bioconversion*. 2nd ed. London: Academic Press.
- [73] Grzenia, D. L., Schell, D. J. & Wickramasinghe, S. R., 2012. Membrane extraction for detoxification of biomass hydrolysates. *Bioresource Technolog*, Volume 111, pp. 248-254.
- [74] Grzenia, D. L., Schell, D. J. & Wickramasinghe, S. R., 2008. Membrane extraction for removal of acetic acid from biomass hydrolysates. *Journal of Membrane Science*, Volume 322, pp. 189-195.
- [75] Haar, L., Gallagher, J. & Kell, J., 1984. NBS/NRC Steam Tables.
- [76] Harmsen, P., Huijgen, W., López, L. B. & Bakker, R., 2010. *Literature Review of Physical and Chemical Pretreatment Processes for Lignocellulosic Biomass*, Wageningen: Energy Research Centre of the Netherlands.
- [77] Harrison, M. D. et al., 2013. Effect of pretreatment on saccharification of sugar cane bagasse by complex and simple enzyme mixtures. *Bioresource Technology*, November, Volume 148, pp. 105-113.
- [78] Heikkila, H., Nurmi, J., Rahkila, L. & Toyryla, M., 1992. *Method for the production of xylitol*. s.l. Patent No. US5081026.
- [79] Heinonen, J., 2013. *CHROMATOGRAPHIC RECOVERY OF CHEMICALS FROM ACIDIC BIOMASS HYDROLYSATES*, Lapeenranta: s.n.
- [80] Heinzle, E., Biber, A. P. & Cooney, C. L., 2006. Citric Acid - Alternative Process using Starch. In: *Development of Sustainable Bioprocesses: Modeling and Assessment*. Chichester: John Wiley & Sons, pp. 125-160.
- [81] Hendriks, A. & Zeeman, G., 2009. Pretreatments to enhance the digestibility of lignocellulosic biomass. *Bioresource Technology*, January , 100(1), pp. 10-18.
- [82] Hermann, T., 2003. Industrial production of amino acids by coryneform bacteria. *J.Biotechnol*, April, Volume 104, pp. 155-172.



- [83] Hernández, V. et al., 2014. Techno-economic and environmental assessment of an olive stone based biorefinery. *Resources, Conservation and Recycling*, November, Volume 92, pp. 145-150.
- [84] Heys, S. D. & Ashkanani, F., 1999. Glutamine. *British Journal of Surgery*, 1 March, pp. 289-290.
- [85] Hiden, A. et al., 2009. Wet disk milling pretreatment without sulfuric acid for enzymatic hydrolysis of rice straw. *Bioresource Technology*, May, 100(10), pp. 2706-2711.
- [86] Humbird, D. et al., 2011. *Process Design and Economics for Biochemical Conversion of Lignocellulosic Biomass to Ethanol*, Golden; Seattle; Atlanta: NREL.
- [87] ICIS Chemical Business, 2006. *Indicative Chemical Prices A-Z: Citric Acid*, s.l.: ICIS.
- [88] Industry Experts, 2017. *Xylitol - A Global Market Overview, report ID: 4033114*, s.l.: Research and Markets.
- [89] Jain, A., Wei, Y. & Tietje, A., 2016. Biochemical conversion of sugarcane bagasse into bioproducts. *Biomass and Bioenergy*, 8 July, Volume 93, pp. 227-242.
- [90] James, B. D., DeSantis, D. A. & Saur, G., 2016. *Hydrogen Production Pathways Cost Analysis (2013 - 2016)*, Arlington: Strategic Analysis Inc..
- [91] Jones, A., 2016. *Fertilizer Prices and Input Cost Movements*, s.l.: Market Realist.
- [92] Jönsson, L. J., Alriksson, B. & Nilvebrant, N.-O., 2013. Bioconversion of lignocellulose: inhibitors and detoxification. *Biotechnology for Biofuels*, January, 6(16), pp. 1-10.
- [93] Kaar, W. E. & Holtzapfel, M. T., 2000. Using lime pretreatment to facilitate the enzymic hydrolysis of corn stover. *Biomass and Bioenergy. Volume 18, Issue 3*, pp. 189-199.
- [94] Kalyuzhnyi, S., 1997. Batch Anaerobic Digestion of Glucose and its Mathematical Modelling. II Description Verification and Application of Model. *Bioresource Technology*, 7 August, Volume 59, pp. 249-258.
- [95] Kamal, S. M. M., Mohamad, N. L., Abdullah, A. G. L. & Abdullah, N., 2011. Detoxification of sago trunk hydrolysate using activated charcoal for xylitol production. *Procedia Food Science*, 23 December, 1(3), pp. 908-913.
- [96] Kannan, B. & Agastian, P., 2016. Xylitol: Production, Optimization and Industrial Application. *International Journal of Current Microbiology and Applied Sciences*, 27 September, 5(9), pp. 324-339.
- [97] Kapanji, K., 2016. *Modelling of Biorefineries Using Sugarcane Lignocellulose at Existing South African Sugar Mills*, Stellenbosch: s.n.
- [98] Kautola, H., Rymowicz, W., Linko, Y.-Y. & Linko, P., 1992. The utilization of beet molasses in citric acid production with yeast. *Science des Aliments*, Volume 12, pp. 383-392.
- [99] Kazi, F. K., Fortman, J. & Anex, R., 2010. *Techno-Economic analysis of biochemical scenarios for production of cellulosic ethanol*, Golden: Technical Report NREL/TP-6A2-46588.
- [100] Khosravi-Darani, K. & Zoghi, A., 2008. Comparison of pretreatment strategies of sugarcane bagasse: Experimental design for citric acid production. *Bioresource Technology*, 10 January, Volume 99, p. 6986-6993.
- [101] Kinoshita, S. e. a., 1958. Glutamic Acid Fermentation. *Proc. Int. Symp. Enzyme Chem.* 2, pp. 464-468.
- [102] Kristiansen, B., Linden, J. & Matthey, M., 2002. *Citric Acid Biotechnology*. London: Taylor & Francis.
- [103] Kumar, A. & Jain, V., 2008. Solid state fermentation studies of citric acid production. *African Journal of Biotechnology Vol. 7 (5)*, pp. 644-650.

- [104] Kumar, A. K. & Sharma, S., 2017. Recent updates on different methods of pretreatment of lignocellulosic feedstocks: a review. *Bioresources and Bioprocessing*, 18 January, 4(1), p. 7.
- [105] Kumar, P., Barrett, D. M., Delwiche, M. J. & Stroeve, P., 2009. Methods for Pretreatment of Lignocellulosic Biomass for Efficient Hydrolysis and Biofuel Production. *Industrial & Engineering Chemistry Research*, March, 48(8), p. 3713–3729.
- [106] Kumar, R., Vikramachakravarthi, D. & Pal, P., 2014. Production and purification of glutamic acid: A critical review towards process intensification. *Chemical Engineering and Processing* 81, p. 59–71.
- [107] Kundu, P. et al., 1984. Pretreatment of Indian cane molasses for increased production of citric acid. *Biotechnology and Bioengineering*, 7 February, 26(9), pp. 1114-1121.
- [108] Lakshmi, P. V. & Mangala, D. S., 2011. Fermentative Production of L-Glutamic Acid. *International Journal of Advanced Biotechnology and Research*, 2(3), pp. 376-381.
- [109] Lakshmi, P. V. & Mangala, D. S., 2011. Fermentative Production of L-Glutamic Acid. *International Journal of Advanced Biotechnology and Research*, 2(3), pp. 376-381.
- [110] Larissa Canilha, A. K. C. T. S. d. S. M. F. A. F. A. W. L. d. C. F. M. d. G. A. F. S. S. d. S., 2012. Bioconversion of Sugarcane Biomass into Ethanol: An Overview about Composition, Pretreatment Methods, Detoxification of Hydrolysates, Enzymatic Saccharification, and Ethanol Fermentation. *Journal of Biomedicine and Biotechnology*, pp. 1-15.
- [111] Larsson, S., Reimann, A., Nilvebrant, N.-O. & Jönsson, L. J., 1999. Comparison of Different Methods for the Detoxification of Lignocellulose Hydrolyzates of Spruce. *Applied Biochemistry and Biotechnology*, January, 77-79(1), pp. 91-103.
- [112] Lavarack, B., Griffin, G. & Rodman, D., 2002. The acid hydrolysis of sugarcane bagasse hemicellulose to produce xylose, arabinose, glucose and other products. *Biomass and Bioenergy*, November, 23(5), pp. 367-380.
- [113] Leal, M. R. L. et al., 2013. Sugarcane straw availability, quality, recovery and energy use: A literature review. *Biomass and Bioenergy*, June, Volume 53, pp. 11-19.
- [114] Lederberg, J., 2000. *Encyclopedia of Microbiology*. 2nd ed. San Diego: Academic Press.
- [115] Lee, Y.-J., 2005. *Oxidation of Sugarcane Bagasse Using a Combination of Hypochlorite and Peroxide*, Chonnam: Department of Food Science, Chonnam National University.
- [116] Lengon, S., Brooks, J. & Maddox, I., 1999. Influence of the glycolytic rate on production of citric acid and oxalic acid by *Aspergillus niger* in solid state fermentation. *World J. Microbiology. Biotechnology*, Volume 15, pp. 493-495.
- [117] Lewis, C. J. & Boyton, R. S., 2000. *Acid Neutralization With Lime for Environmental Control and Manufacturing Processes*, s.l.: National Lime Association.
- [118] Linzmeyera, P. et al., 2014. Comparison of Different Separation Methods for Solids Removal in an Ethanol Fermentation Broth from Banana Culture Waste. *Chemical Engineering Transactions*, Volume 37, pp. 355-360.
- [119] Li, T.-H., 1965. *The production of glutamic acid by fermentation*, Masters Theses, s.l.: Missouri University of Science and Technology.

- [120] Liu, H. et al., 2014. Purification of Hemicelluloses in Pre-Hydrolysis Liquor of Kraft-Based Dissolving Pulp Production Process Using Activated Carbon and Ion-Exchange Resin Adsorption Followed by Nanofiltration. *Journal of Biobased Materials and Bioenergy*, June, 8(3), pp. 325-330.
- [121] Liu, W. et al., 2017. Analysis of the Global Warming Potential of Biogenic CO<sub>2</sub> Emission in Life Cycle Assessments. 3 January, Volume 7, p. 39857 .
- [122] Liyakathali, N. A. M., 2014. *Ultrasonic Pretreatment of Energy Vane Bagasse For Biofuel Production*, s.l.: s.n.
- [123] Maddox, I. et al., 1985. Production of citric acid from sugars present in wood hemicellulose using *Aspergillus niger* and *Saccharomycopsis lipolytica*. *Biotechnology Letters*, November, 7(11), pp. 815-818.
- [124] Madhusudan, C., Manoj, S. & Rishi, C. M., 2010. Citric acid production from cane molasses using submerged fermentation by *Aspergillus ATCC9142*. *Journal of Pharmacy Research*, 13 May, 3(6), pp. 1215-1222.
- [125] Makinen, K., 2016. *Understanding Xylitol*. [Online] Available at: <http://xylitol.org/xylitol-video-presentations/understanding-xylitol>
- [126] Mala, R. & Anish, R., 2008. Bioethanol from lignocellulosic biomass part III hydrolysis and fermentation. In: A. Pandey, ed. *Handbook of plant-based biofuels*. s.l.:CRC Press, pp. 159-173.
- [127] Mamman, A. S. et al., 2008. Furfural: Hemicellulose/xylose-derived biochemical. *Biofuels, Bioproducts. Biorefining.*, 12 September, Volume 2, pp. 438-454.
- [128] Manonmani, H. K. & Sreekantiah, K. R., 1987. Studies on the conversion of cellulose hydrolysate into citric acid by *Aspergillus niger*. *Process Biochem.*, Volume 22, pp. 92-94.
- [129] Martin, C. & Thomsen, A. B., 2007. Wet oxidation pretreatment of lignocellulosic residues of sugarcane, rice, cassava and peanuts for ethanol production. *Journal of Chemical Technology & Biotechnology*, v82 n2, pp. 174-181.
- [130] Mason, T. J. & Lorimer, J. P., 2003. *Applied Sonochemistry*. s.l.:Wiley-VCH Verlag GmbH & Co.
- [131] Max, B. et al., 2010. Biotechnological production of citric acid. *Brazilian Journal of Microbiology* v.41(4), p. 862–875.
- [132] McGinnis, G., Wilson, W. & Mullen, C., 1983. Biomass pretreated with water and high pressure oxygen The WO process. *Ind Eng Chem Prod Res Dev*, Volume 22, p. 352–357.
- [133] Melaja, A. J. & Hamalainen, L., 1977. *Process for Making Xylitol*. United States of America, Patent No. US4008285.
- [134] Miesher, G. M. & Haute, T., 1975. *Method of Producing Glutamic Acid By Fermentation*. United States of America, Patent No. 3,929,575.
- [135] Mikkola, J. & Salmi, T., 2001. Three-phase catalytic hydrogenation of xylose to xylitol - prolonging the catalyst activity by means of on-line ultrasonic treatment. *Catalysis Today*, Volume 64, p. 271–277.
- [136] Mikkola, J.-P., Salmi, T. & Sjöholm, R., 1999. Modelling of kinetics and mass transfer in the hydrogenation of xylose over Raney nickel catalyst. *Journal of Chemical Technology and Biotechnology*.
- [137] Mikkola, J.-P. et al., 2003. Hydrogenation of xylose to xylitol on sponge nickel catalyst – a study of the process and catalyst deactivation kinetics. *Brazilian Journal of Chemical Engineering*, 20(3).

- [138] Mikkola, J.-P. et al., 2000. Deactivation kinetics of Mo-supported Raney Ni catalyst in the hydrogenation of xylose to xylitol. *Applied Catalysis A: General*, Volume 196, p. 143–155.
- [139] Mikkola, J.-P. et al., 2000. Deactivation kinetics of Mo-supported Raney Ni catalyst in the hydrogenation of xylose to xylitol. *Applied Catalysis A: General*, 196(1), pp. 143-155.
- [140] Mishra, D. K., Dabbawala, A. A. & Hwang, J.-S., 2013. Ruthenium nanoparticles supported on zeolite Y as an efficient catalyst for selective hydrogenation of xylose to xylitol. *Journal of Molecular Catalysis A*, Volume Chemical 376, p. 63–70.
- [141] Moeller, L. et al., 2013. Citric acid production from sucrose by recombinant *Yarrowia lipolytica* using semicontinuous fermentation. *Engineering in Life Sciences*, March, 13(2), p. 163–171.
- [142] Mohamad, N. L., Kamal, S. M. M. & Mokhtar, M. N., 2015. Xylitol Biological Production: A Review of Recent Studies. *Food Reviews International*, 34(1), pp. 74-89.
- [143] Monlau, F. et al., 2014. Do furanic and phenolic compounds of lignocellulosic and algae biomass hydrolyzate inhibit anaerobic mixed cultures? A comprehensive review. *Biotechnology Advances*, April, 32(5), pp. 934-951.
- [144] Mosier, N. et al., 2005. Features of promising technologies for pretreatment of lignocellulosic biomass. *Bioresource Technology*, Volume 96, pp. 673-686.
- [145] Mountraki, A., Koutsospyros, K. & Mlayah, B., 2017. Selection of Biorefinery Routes: The Case of Xylitol and its Integration with an Organosolv Process. *Waste and Biomass Valorization*, 27 January. pp. 1-18.
- [146] Moutta, R. et al., 2011. Statistical optimization of sugarcane leaves hydrolysis into simple sugars by dilute sulfuric acid catalyzed process. *Sugar Technology*.
- [147] Moutta, R. et al., 2011. Statistical optimization of sugarcane leaves hydrolysis into simple sugars by dilute sulfuric acid catalyzed process. *Sugar Technology*.
- [148] Muscat, D., Adhikari, B., Adhikari, R. & Chaudhary, D., 2012. Comparative study of film forming behaviour of low and high amylose starches using glycerol and xylitol as plasticizers. *J. Food Eng.*, Volume 109, pp. 189-201.
- [149] Mussatto, S. I., Moncada, J., Roberto, I. C. & Cardona, C. A., 2013. Techno-economic analysis for brewer's spent grains use on a biorefinery concept: The Brazilian case. *Bioresource Technology*, November, Volume 148, pp. 302-310.
- [150] Mussatto, S. & Teixeira, J., 2010. Lignocellulose as raw material in fermentation processes. *Applied Microbiology*, pp. 897-907.
- [151] Naleli, K., 2016. *Process Modelling In Production of Biobutanol from Lignocellulosic Biomass via ABE Fermentation*, Stellenbosch: Stellenbosch University.
- [152] Nampoothiri, M. & Pandey, A., 1996. Solid state fermentation for L-glutamic acid production using *Brevibacterium sp.* *Biotechnol. Lett.* 18 (2), pp. 199-204.
- [153] Nguyen, T. K., Martinkova, L., Seichert, L. & Machek, F., 1992. Citric acid production by *Aspergillus niger* using media containing low concentrations of glucose or corn starch. *Folia microbiologica*, December, 37(6), pp. 433-441.

- [154] Ogawa, T. a. F. A., 1976. Additive effect of ferrocyanide treatment and step change of pH on citric acid production from Iranian beet molasses with *Aspergillus niger*. *Journal of Fermentation Technology*, Volume 54, pp. 63-66.
- [155] Osiewacz, H., 2002. *The Mycota - A Comprehensive Treatise on Fungi as Experimental Systems for Basic and Applied Research, Industrial Applications Vol. 10.*. Berlin: Springer Science & Business Media.
- [156] Ostrem, K., 2004. *Greening Waste: Anaerobic Digestion For Treating The Organic Fraction Of Municipal Solid Wastes*, Columbia : s.n.
- [157] Ostrinskaya, A., Kelley, J. A. & Kunz, R. R., 2016. Characterization of nitrated sugar alcohols by atmospheric pressure chemical-ionization mass spectrometry. *Rapid Communications in Mass Spectrometry*, 28 November, 31(4), pp. 333-343.
- [158] Pallares, J., Rodriguez, S. & Sanroman, A., 1996. Citric acid production in submerged and solid state culture of *Aspergillus niger*. *Bioprocess Engineering*, June, 15(1), pp. 31-33.
- [159] Palmqvist, E. & Hahn-Hagerdal, B., 2000. Fermentation of lignocellulosic hydrolysates. II: inhibitors and mechanisms of inhibition. *Bioresource Technology*, August, 74(1), pp. 25-33.
- [160] Pal, P., Dekonda, V. C. & Kumar, R., 2015. Fermentative production of glutamic acid from renewable carbon source: Process intensification through membrane-integrated hybrid bio-reactor system. *Chemical Engineering and Processing: Process Intensification*, June, Volume 92, p. 7–17.
- [161] Pal, P., Kumar, R., Vikrama Chakravarthi, D. & Chakraborty, S., 2016. Modeling and simulation of continuous production of L (+) glutamic acid in a membrane-integrated bioreactor. *Biochemical Engineering Journal*, 15 February, Volume 106, pp. 68-86.
- [162] Pandey, A. et al., 2015. *Industrial Biorefineries & White Biotechnology*. Amsterdam: Elsevier.
- [163] Panneerselvam, A., 2013. *Ozonolysis – A Novel Pretreatment Technology For Energy Grasses*, Raleigh: s.n.
- [164] Papagianni, M., 2007. Advances in citric acid fermentation by *Aspergillus niger*: Biochemical aspects, membrane transport and modeling. *Biotechnology Advances*, June, 25(3), pp. 244-263.
- [165] 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*, 10 Jul, 7(105), pp. 1-19.
- [166] Peterson, M. E., 2013. Xylitol. *Topics in Companion Animal Medicine*, Volume 28, Issue 1, pp. 18-20.
- [167] Petrides, D., 2000. *Bioprocess Design*. Scotch Plains: INTELLIGEN, INC.
- [168] Qingdao Greaty International Trade Co., L., 2017. *Food grade L-Glutamic Acid powder amino acid*. [Online] Available at: [https://www.alibaba.com/product-detail/Food-grade-L-Glutamic-Acid-powder\\_60554105641.html?spm=a2700.7724838.0.0.JC6Pvd](https://www.alibaba.com/product-detail/Food-grade-L-Glutamic-Acid-powder_60554105641.html?spm=a2700.7724838.0.0.JC6Pvd) [Accessed June 2017].
- [169] Quintero, J., Moncada, J. & Cardona, C., 2013. Techno-economic analysis of bioethanol production from lignocellulosic residues in Colombia: a process simulation approach. *Bioresource Technology*, July, Volume 139, pp. 300-307.

- [170] Quiroz-Castañeda, R. E. & Folch-Mallol, J. L., 2013. Hydrolysis of Biomass Mediated by Cellulases for the Production of Sugars. In: *Sustainable Degradation of Lignocellulosic Biomass - Techniques, Applications and Commercialization*. s.l.:Intech, pp. 133-134.
- [171] Rafiqul, I. & Sakinah, A. M. M., 2013. Processes for the Production of Xylitol - A Review. *Food Reviews International*, 29(2), pp. 127-156.
- [172] Rajendran, K., Kankanala, H. R., Lundin, M. & Taherzadeh, M. J., 2014. A Novel Process Simulation Model (PSM) for Anaerobic Digestion Using Aspen Plus. *Bioresource Technology*, September, Volume 168, pp. 7-13.
- [173] Rao, R., Jyothi, P. & Rao, V., 2007. Biotechnological production of xylitol from hemicelulose materials. *Lignocellulose biotechnology*.
- [174] Ravella, S. R., Gallagher, J., Fish, S. & Prakasham, R. S., 2012. Overview on Commercial Production of Xylitol, Economic Analysis and Market Trends. In: S. da Silva & A. Chandel, eds. Berlin, Heidelberg: Springer-Verlag, pp. 291-306.
- [175] regyp, 2017. *Calculating landed cost of gypsum*. [Online] Available at: <https://www.regyp.com.au/calculating-landed-cost-of-gypsum-per-tonne/256/> [Accessed March 2017].
- [176] Renon, H. & Prausnitz, J., 1968. Local Compositions in Thermodynamic Excess Functions for Liquid Mixtures. *AIChE J.*, 14(1), pp. 135-144.
- [177] Ritthausen, K. H. L., n.d. Hypothalamic morphology following ingestion of aspartame or MSG in the neonatal rodent and primate: A preliminary report. *Journal of Toxicology*, 2(2), pp. 471-480.
- [178] Roche, C. et al., 2009. Particle concentration and yield stress of biomass slurries during enzymatic hydrolysis at high-solids loadings. *Biotechnology Bioenergy*, 27 April, 104(2), pp. 290-300.
- [179] S. Kinoshita, S. U. M. S., 1957. Studies on the amino acid fermentation. Part 1. Production of l-glutamic acid by various microorganisms. *Journal of General and Applied Microbiology*, pp. 193-205.
- [180] S.U. Choi, T. N. T. Y. J., 2004. Enhanced glutamic acid production of *Brevibacterium* sp. with temperature shift-up cultivation. *Journal of Bioscience and Bioengineering*, pp. 211-213.
- [181] SA National Department of Health, 2013. *Strategic Plan for the Prevention and Control of Non-Communicable Diseases 2013–17*, Johannesburg: Department of Health.
- [182] Saini, J. K., Saini, R. & Tewari, L., 2015. Lignocellulosic agriculture wastes as biomass feedstocks for second-generation bioethanol production: concepts and recent developments. *3 Biotech*, August, 5(4), pp. 337-353.
- [183] Sanchez, S. & Demain, A., 2014. Production of Amino Acids. *Encyclopedia of Food Microbiology*, Volume 1.
- [184] Sano, C., 2009. History of glutamate production. *American Journal of Clinical Nutrition*, Volume 90, pp. 728-732.
- [185] Sapolsky, R., 2005. *Biology and Human Behavior: The Neurological Origins of Individuality*, 2nd edition. s.l.:The Teaching Company.

- [186] Sarkar, N., Ghosh, S. K., Bannerjee, S. & Aikat, K., 2012. Bioethanol production from agricultural wastes: an overview. *International Journal of Chemical Engineering: Renewable Energy*, vol. 37, no. 1, p. 19–27.
- [187] Sepúlveda-Huerta E, T.-L. S. B.-G. V. R. J. V. M., 2006. Production of detoxified sorghum straw hydrolysates for fermentative purposes. *Journal of Science Food Agriculture*, Volume 86, p. 2579–2586.
- [188] Shankaranand, V. S. & Lonsane, B. K., 1993. Sugarcane-pressmud as a novel substrate for production of citric acid by solid-state fermentation. *World J. Microbiol. Biotechnol.*, 9, May, 9(3), pp. 377-380.
- [189] Shetty, P. R., Ravella, S. R. & Hobbs, P., 2009. Current trends in biotechnological production of xylitol and future prospects. *Biotechnology and Pharmacy*, January, 3(1), pp. 8-36.
- [190] Shuler, M. L. & Kargi, F., 2002. *Bioprocess Engineering: Basic Concepts*. 2nd ed. s.l.:Prentice Hall.
- [191] Sian, B. H., 2013. *Production of Citric Acid*. [Online]  
Available at: <https://prezi.com/ycx0ssuf8o4p/production-of-citric-acid>
- [192] Sierra, R., Granda, C. & Holtzaple, M., 2009. Lime pretreatment. *Methods in Molecular Biology*. Volume 581, pp. 115-124.
- [193] Silva, A. S. d. et al., 2010. Milling pretreatment of sugarcane bagasse and straw for enzymatic hydrolysis and ethanol fermentation. *Bioresource Technology*, October, 101(19), pp. 7402-7409.
- [194] Singh, G. M. et al., 2015. Estimated global, regional, and national disease burdens related to sugar-sweetened beverage consumption in 2010. *Circulation*, June, 132(8), pp. 639-666.
- [195] Smithers, J., 2014. Review of sugarcane trash recovery systems for energy cogeneration in South Africa. *Renewable and Sustainable Energy Reviews*, April, Volume 32, pp. 915-925.
- [196] Soccol, C. R., Vandenberghe, L. P., Rodrigues, C. & Pandey, A., 2006. New Perspectives for Citric Acid Production and Application. *Food Technology. Biotechnology*, 44(2), pp. 141-149.
- [197] SRI Consulting, 2000. *Citric acid: Supply/demand*, Zurich: ICIS Chemical Business.
- [198] Stephan, J. et al., 2004. Diffusion of glutamic acid in relation to growth of *Geotrichum candidum* and *Penicillium camembertii* at the surface of a solid medium. *J. Chem. Technol. Biotechnol.*, Volume 79, pp. 234-239.
- [199] Sunitha, I., Rao, M. & Ayyanna, C., 1998. Optimization of medium constituents and fermentation conditions for the production of l-glutamic acid by the coim-mobilized whole cells of *Micrococcus glutamicus* and *Pseudomonas reptilivora*. *Bioprocess. Eng.* 18, pp. 353-359.
- [200] Sun, R.-C., 2008. Detoxification and Separation of Lignocellulosic Biomass Prior to Fermentation for Bioethanol Production by Removal of Lignin and Hemicelluloses. *bioresources*, pp. 452-455.
- [201] Sun, Y. & Cheng, J., 2002. Hydrolysis of lignocellulosic materials for ethanol production: a review. *Bioresource Technology*, May, 83(1), pp. 1-11.
- [202] Tao, L. et al., 2011. Process and techno-economic analysis of leading pretreatment technologies for lignocellulosic ethanol production using switchgrass. *Bioresource Technology*, 4 August, 102(24), p. 11105–11114.
- [203] Tchobanoglous, G., Burton, F. L. & Stensel, D. H., 2004. *Wastewater Engineering, Treatment and Reuse*. 4th ed. New York: McGraw-Hill Education.

- [204] Todaro, C. C., 1996. Centrifugation. In: H. C. Vogel & C. L. Todaro, eds. *Fermentation and Biochemical Engineering Handbook (Second Edition): Principles, Process Design, and Equipment*. 2nd ed. Westwood(NJ): William Andrew Publishing, p. 558–589.
- [205] Tomishige, K. et al., 2016. Selective transformation of hemicellulose (xylan) into n-pentane, pentanols or xylitol over a rhenium-modified iridium catalyst combined with acids. *Green Chemistry*, 18(1), pp. 165-175.
- [206] Towler, G. & Sinnott, R., 2012. *Chemical Engineering Design: Principles, Practice and Economics of Plant and Process Design*. 2nd ed. s.l.:Elsevier.
- [207] Travaini, R., Martín-Juárez, J., Lorenzo-Hernando, A. & Bolado-Rodríguez, S., 2016. Ozonolysis: An advantageous pretreatment for lignocellulosic biomass revisited. *Bioresource Technology Volume 199*, p. 2–12.
- [208] Tutt, M., Kikas, T. & Olt, J., 2012. *Comparison of Different Pretreatment Methods on Degradation of Rye Straw*, Jelgava: Engineering for Rural Development.
- [209] USDA, Economic Research Service, 2017. *Tighter global sugar supplies support recent price increases*. [Online] Available at: <https://www.ers.usda.gov/data-products/chart-gallery/gallery/chart-detail/?chartId=81896> [Accessed 3 August 2017].
- [210] Ushakov, V., 2005. Principles of the Development of Piezoelectric Transducers for Ultrasonic Testing of Articles with Flat and Curvilinear Surfaces. *Russian Journal of Nondestructive Testing*, 41(6), pp. 355-361.
- [211] van Zyl, C., Prior, B. A. & Preez, J. C. d., 1991. Acetic acid inhibition of d-xylose fermentation by *Pichia stipitis*. *Enzyme Microbiol Technology*, 13(1), pp. 82-86.
- [212] Vandenberghe, L. P., Soccol, C. R., Pandey, A. & Lebeault, J.-M., 1999. Microbial production of citric acid. *Brazilian Archives of Biology and Technology vol.42 no.3*.
- [213] Vandenberghe, L. P., Soccol, C. R., Pandey, A. & Lebeault, J.-M., 1999. Microbial Production of Citric Acid. *Brazilian Archives of Biology and Technology*, 42(3).
- [214] Villarreal, M., Prata, A., Felipe, M. & Silva, J. A. E., 2006. Detoxification procedures of eucalyptus hemicellulose hydrolysate for xylitol production by *Candida guilliermondii*. *Enzyme and Microbial Technology*, 6 December, 40(1), pp. 17-24.
- [215] Vishwanath, B. A., 2006. *Electroporation Studies on Brevibacterium spp with Respect to Glutamic Acid Fermentation*, s.l.: U.G.C.
- [216] Watson, N., Prior, B., Lategan, P. & Lussi, M., 1984. Factors in acid treated bagasse inhibiting ethanol production from d-xylose by *Pachysolen tannophilus*. *Enzyme Microbiology Technology*, 6(10), pp. 451-456.
- [217] Werpy, T. et al., 2004. Top Value Added Chemicals From Biomass. Volume 1 - Results of Screening for Potential Candidates From Sugars and Synthesis Gas. August, Volume 1, pp. 63-64.
- [218] Wright, I. & Hayward, L., 1960. The Pentitol Pentanitrates. *Canadian Journal of Chemistry*, Volume 38, pp. 316-319.



- [219] Wyman, C. E. & Yang, B., 2009. Cellulosic biomass could help meet California's transportation fuel needs. *California Agriculture*, 1 October, 63(4), pp. 185-190.
- [220] Yachmenev, V., Condon, B., Klasson, T. & Lambert, A., 2009. Acceleration of the Enzymatic Hydrolysis of Corn Stover and Sugar Cane Bagasse Celluloses by Low Intensity Uniform Ultrasound. *Journal of Biobased Materials and Bioenergy*, 3(1), pp. 25-31.
- [221] Yadav, M., Mishra, D. K. & Hwang, J.-S., 2011. Catalytic hydrogenation of xylose to xylitol using ruthenium catalyst on NiO. *Applied Catalysis A: General*, pp. 305-600.
- [222] Yalcin, S. K., Bozdemir, M. T. & Ozbas, Z. Y., 2010. *Citric acid production by yeasts: Fermentation conditions, process optimization and strain improvement*, Ankara: s.n.
- [223] Yamaguchi, A., Sato, O., Mimura, N. & Shiraia, M., 2016. Catalytic production of sugar alcohols from lignocellulosic biomass. *Catalysis Today*, 1 May, Volume 265, pp. 199-202.
- [224] Yoshimura, M., Takenaka, Y., Ikeda, S. & Yoshii, H., 1982. US, Patent No. 4,347,317.
- [225] Zauba, 2016. *Detailed Export Data of raney nickel catalyst under HS Code 3815*. [Online] Available at: <https://www.zauba.com/export-RANEY+NICKEL+CATALYST/hs-code-3815-hs-code.html> [Accessed May 2016].
- [226] Zhao, D. & Li, Y.-R., 2015. Climate Change and Sugarcane Production: Potential Impact and Mitigation Strategies. *International Journal of Agronomy*, 17 September, Issue Article ID 547386, p. 10.
- [227] Zhaojiang Wang, J. Z. X. W. Z. L. Y. F. M. Q., 2016. Limited adsorption selectivity of active carbon toward non-saccharide compounds in lignocellulose hydrolysate. *Bioresource Technology*, May, Volume 208, pp. 195-199.
- [228] Zheng, Y., Lin, H. & Tsao, G. T., 1998. Pretreatment for Cellulose Hydrolysis by Carbon Dioxide Explosion. *Biotechnology progress Vol 14.*, pp. 890-896.
- [229] Zheng, Y., Pan, Z. & Zhang, R., 2009. Overview of biomass pretreatment for cellulosic production. *International Journal of Agricultural and Biological Engineering*, September, 2(3), pp. 51-68.
- [230] Zhou, P.-P., Meng, J. & Bao, J., 2017. Fermentative production of high titer citric acid from corn stover feedstock after dry dilute acid pretreatment and biodetoxification. *Bioresource Technology*, January, Volume 224, pp. 563-572.

## Appendix A. Xylitol Techno-economic Calculation Summary

The following appendix specifies the calculations that were used to conduct the techno-economic analyses for the xylitol-electricity scenario. Calculations include the specification of components (Table 68) in Aspen Plus<sup>®</sup>, the configuration of process flow sheets (Appendix A2), mass and energy balance flows (Appendix A3), utility consumption (Appendix A4), as well as equipment sizing and costing to determine profitability indicators (Appendices A5-A9).

### A1. Xylitol Component Summary

Table 68 - Xylitol Aspen Component Summary

Component ID	Type	Component name	Alias
GLUCAN	Solid	Di-lactic-Acid	C6H10O5
XYLAN	Solid	Glutaric-Acid	C5H8O4
ARABINAN	Solid	Glutaric-Acid	C5H8O4
ACETATE	Solid	Acetic-Acid	C2H4O2-1
ASH	Solid	Calcium-Oxide	CAO
LIG-SOL	Conventional	Vanillin	C8H8O3-D1
LIGNIN	Solid	Vanillin	C8H8O3-D1
CELLULOS	Solid	Di-lactic-Acid	C6H10O5
EXTRACT	Conventional	Dextrose	C6H12O6
H2SO4	Conventional	Sulphuric-Acid	H2SO4
H2O	Conventional	Water	H2O
O2	Conventional	Oxygen	O2
N2	Conventional	Nitrogen	N2
CO2	Conventional	Carbon-Dioxide	CO2
H2	Conventional	Hydrogen	H2
GLUCOSE	Solid	Dextrose	C6H12O6
XYLOSE	Solid	Xylose	C5H10O5-D2
ARABINOS	Solid	Arabinose	C5H10O5-D1
CELLOBIO	Solid	Cellobiose	C12H22O11-D1
GLUCOLIG	Conventional	Di-lactic-Acid	C6H10O5
XYLOLIG	Conventional	Glutaric-Acid	C5H8O4
ARABOLIG	Conventional	Glutaric-Acid	C5H8O4
FURFURAL	Conventional	Furfural	C5H4O2
HMF	Conventional	5-Hydroxymethylfurfural	C6H6O3-N5
AACID	Conventional	Acetic-Acid	C2H4O2-1
NH3	Conventional	Ammonia	H3N
AMM-SULF	Solid	Ammonium-Sulphate	(NH4)2SO4
LIME	Solid	Calcium-Oxide	CAO
ETHANOL	Conventional	Ethanol	C2H6O-2
AIR	Conventional	Air	AIR
XYLITOL	Conventional	Xylitol	C5H12O5
ARABITOL	Conventional	D-Arabinitol	C5H12O5-N2
SORBITOL	Conventional	Sorbitol	C6H14O6
CASO4	Solid	Calcium-Sulphate	CASO4

## A2. Xylitol Aspen Simulation Process Flow Diagrams

### A2.1. Overall Xylitol Biorefinery

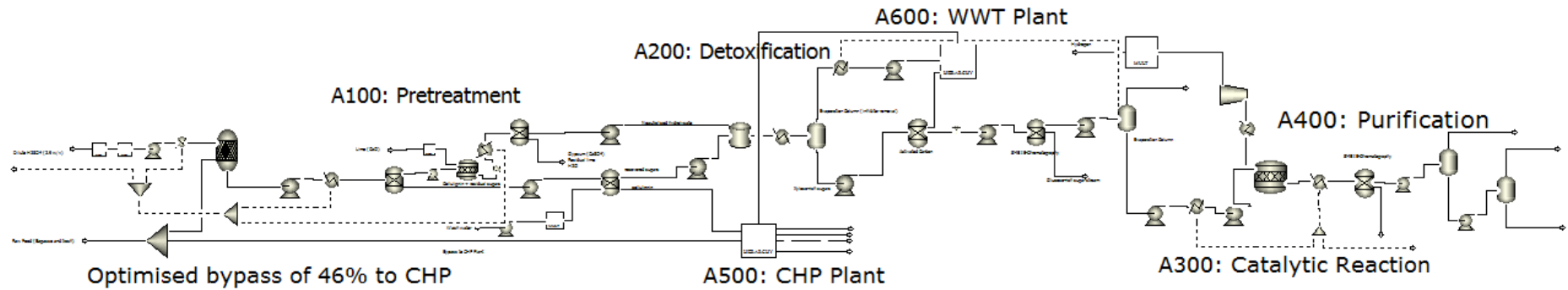


Figure 55 - Aspen Xylitol Biorefinery Flowsheet

### A2.2. Area 100 Pretreatment (DA, Neutralization, Washing)

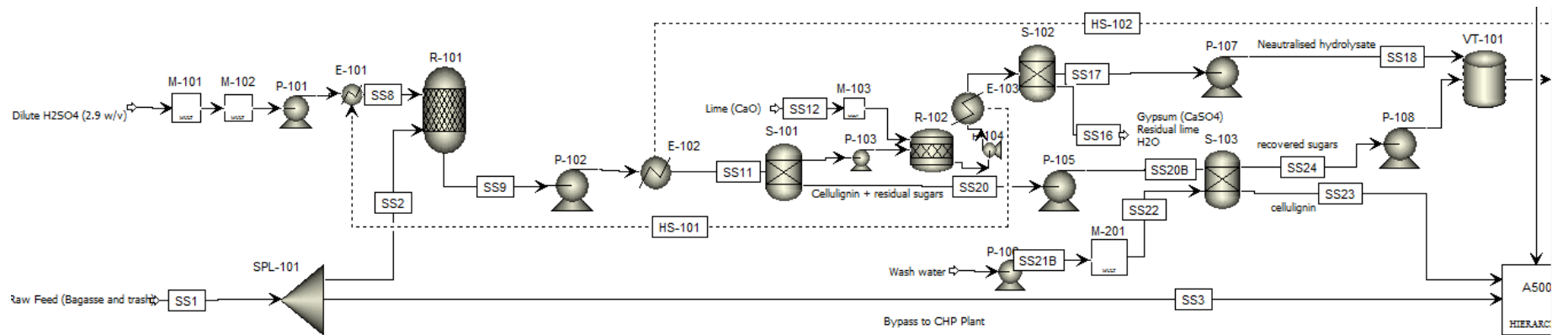


Figure 56 - Aspen Xylitol Pretreatment Flowsheet (Area 100)

### A2.3. Area 200 Detoxification (Solids Separation and Sugar Washing Recovery)

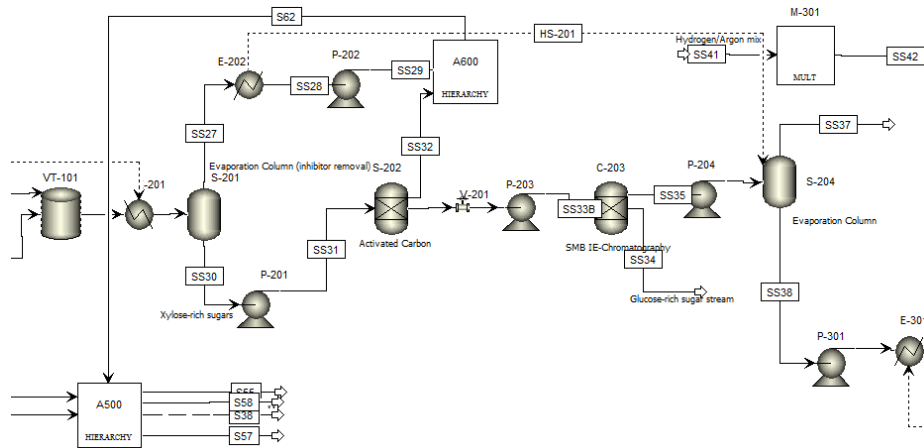


Figure 57 - Aspen Xylitol Detoxification Flowsheet (Area 200)

### A2.4. Area 300 Catalytic Reaction

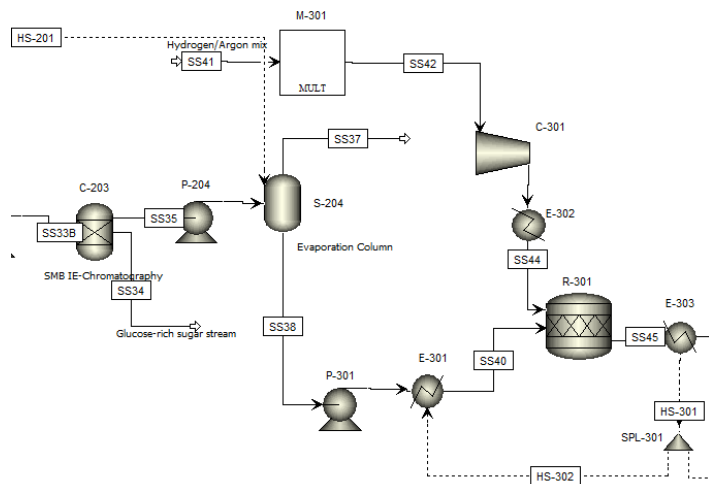


Figure 58 - Aspen Xylitol Catalytic Reactor Flowsheet (Area 300)

### A2.5. Area 400 Purification and Drying

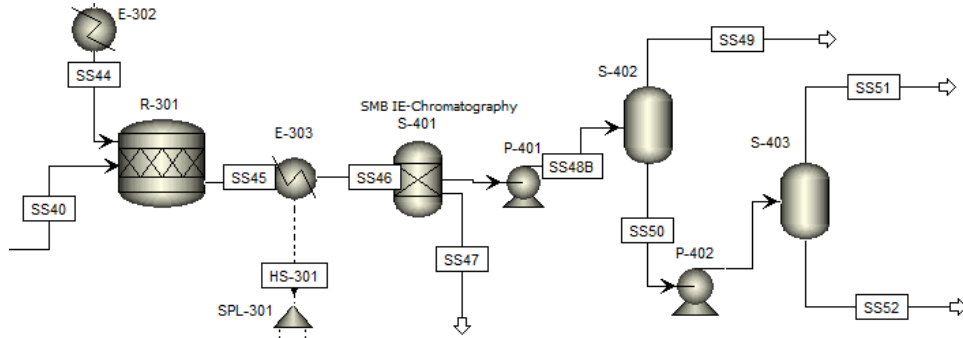


Figure 59 - Aspen Xylitol Purification and Drying Flowsheet (Area 400)

### A2.6. Area 500 Combined Heat and Power Plant

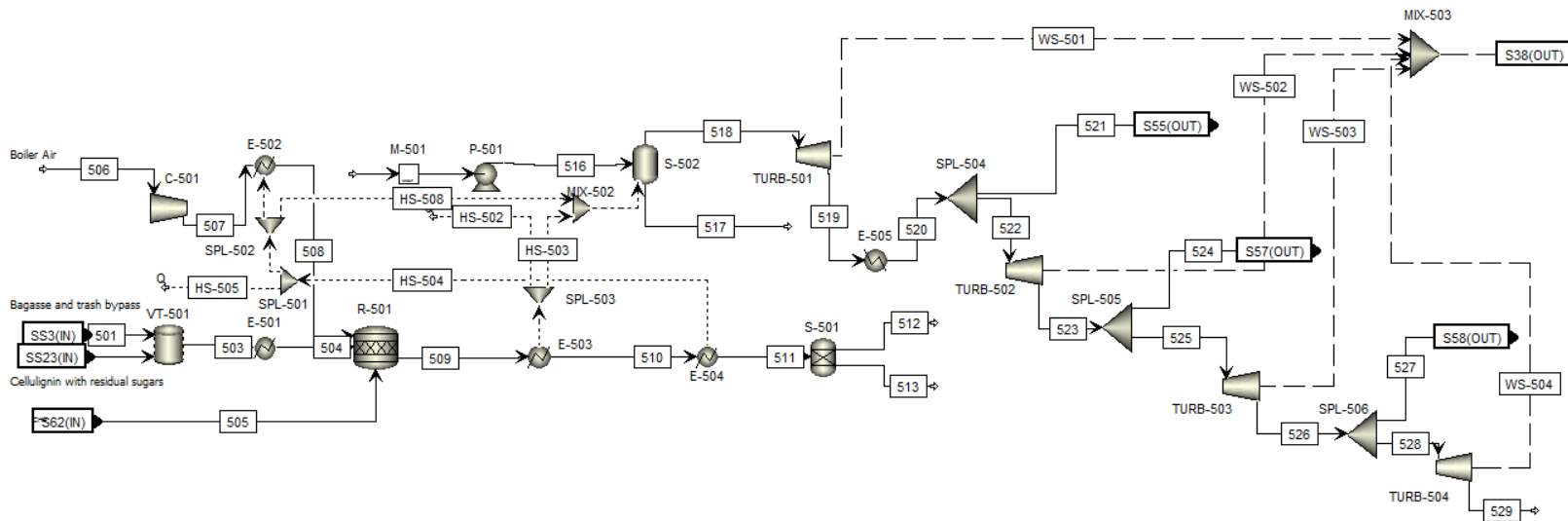


Figure 60 - Aspen Xylitol CHP Plant Flowsheet (Area 500)

A2.7. Area 600 Wastewater Treatment Plant

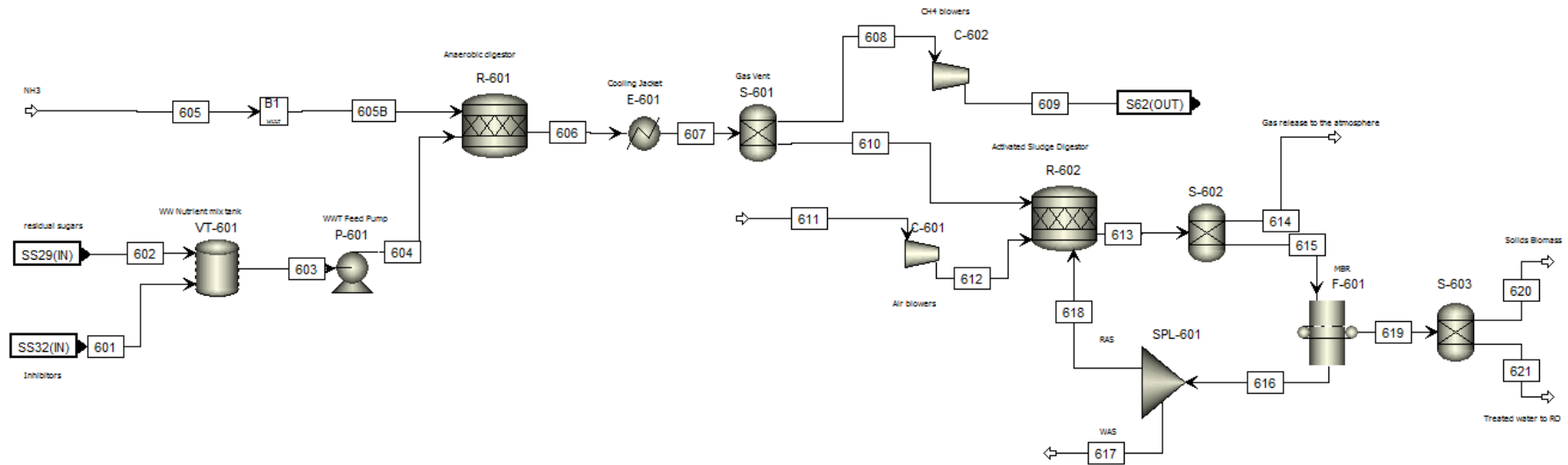


Figure 61 - Aspen Xylitol WWT Flowsheet (Area 600)

### A3. Xylitol Mass and Energy Balance Streams

Streams	S4	S5	S55	S57	S58	S62	SS1	SS2	SS3	SS4	SS5	SS6	SS7	SS8	SS9	SS10	SS11	SS12	SS13	SS14
Temperature C	137.1	60	340		219.2	55.6	25	25	25	40	40	40	40	117.6	130.6	130.6	60	25	25	78.5
Pressure atm	40	2.2	28		9.5	1.2	1	1	1	1	1	1	1.8	1.8	2.714	3	3	1	1	1
Vapor Frac	0	0	1		1	1	0	0	0	0	0	0	0	0.061	0	0	0	0	0	0
Mole Flow kmol/hr	4371.507	12996.89	6042.769	0	5139.828	48.231	2862.32	1545.653	1316.667	15992.76	14521.42	12924.07	12924.07	12924.07	14422	14422	14422	100	89	13085.89
Mass Flow tons/hr	92.942	271.791	120	0	102.069	1.118	113.529	61.306	52.223	325.292	295.365	262.875	262.875	262.875	324.181	324.181	324.181	6.181	5.502	277.293
Volume Flow cum/hr	87.022	243.107	10282.33	0	21138.58	1079.309	89.501	48.33	41.17	293.48	266.48	237.167	237.159	14137.85	299.497	299.494	285.097	1.701	1.513	246.755
Enthalpy Gcal/hr	-296.697	-896.094	-334.287		-289.149	-1.262	-247.59	-133.699	-113.891	-1098.07	-997.047	-887.372	-887.366	-861.706	-994.854	-994.851	-1013.96	-15.169	-13.5	-909.594
Mass Flow tons/hr																				
GLUCAN	0	0	0	0	0	0	2.498	1.349	1.149	0	0	0	0	0	1.215	1.215	1.215	0	0	0
XYLAN	0	0	0	0	0	0	11.58	6.253	5.327	0	0	0	0	0	0.625	0.625	0.625	0	0	0
ARABINAN	0	0	0	0	0	0	2.157	1.165	0.992	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	1.362	0.736	0.627	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	2.271	1.226	1.044	0	0	0	0	0	1.226	1.226	1.226	0	0	0
LIG-SOL	0	0.106	0	0	0	0	0	0	0	0	0	0	0	0	0.116	0.116	0.116	0	0	0.106
LIGNIN	0	0	0	0	0	0	14.305	7.725	6.58	0	0	0	0	0	7.609	7.609	7.609	0	0	0
CELLULOS	0	0	0	0	0	0	26.452	14.284	12.168	0	0	0	0	0	12.831	12.831	12.831	0	0	0
EXTRACT	0	0	0	0	0	0	4.428	2.391	2.037	0	0	0	0	0	2.391	2.391	2.391	0	0	0
H2SO4	0	6.947	0	0	0	0	0	0	0	9.433	8.566	7.623	7.623	7.623	7.623	7.623	7.623	0	0	0
H2O	85.975	255.608	120	0	102.069	0	48.477	26.178	22.299	315.859	286.8	255.252	255.252	255.252	280.484	280.484	280.484	0	0	256.884
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0.374	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	1.482	0	0	0	0	0	0	0	0	0	0	0	0	1.626	1.626	1.626	0	0	1.482
XYLOSE	5.936	5.507	0	0	0	0	0	0	0	0	0	0	0	0	6.043	6.043	6.043	0	0	5.507
ARABINOS	1.031	1.206	0	0	0	0	0	0	0	0	0	0	0	0	1.324	1.324	1.324	0	0	1.206
CELLOBIO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.041	0.041	0.041	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0	0.206	0	0	0	0	0	0	0	0	0	0	0	0	0.226	0.226	0.226	0	0	0.206
HMF	0	0.06	0	0	0	0	0	0	0	0	0	0	0	0	0.065	0.065	0.065	0	0	0.06
AACID	0	0.67	0	0	0	0	0	0	0	0	0	0	0	0	0.736	0.736	0.736	0	0	0.67
NH3	0	0	0	0	0	0.471	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	6.181	5.502	1.529
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SORBITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CASO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	9.643
CH4	0	0	0	0	0	0.273	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CACO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO3-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PRO-ACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BUT-ACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Streams	SS14B	SS15	SS16	SS17	SS18	SS19	SS20	SS20B	SS21	SS21B	SS22	SS23	SS23B	SS24	SS25	SS26	SS27	SS28	SS29	SS30	SS31	SS32	SS33
Temperature C	78.5	50	50	50	50	60	60	60	25	25	25	52.3	52.3	52.3	50.3	105	79.9	40	40	79.9	60.9	60.9	60.9
Pressure atm	1.2	1.2	1.2	1.2	3	2	2	2.5	1	1.2	1.2	1.2	1.2	1.2	1.2	1.2	0.425	0.425	1.5	0.425	3.5	3.5	3.5
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0
Mole Flow kmol/hr	13085.89	13085.89	283.036	12802.86	12802.86	12996.89	1425.11	1425.11	555.084	555.084	494.025	687.58	1231.555	1231.555	14034.41	14034.41	1684.129	1684.129	1684.129	12350.28	12350.28	189.894	12160.39
Mass Flow tons/hr	277.293	277.293	15.026	262.267	262.267	271.791	52.39	52.39	11.023	11.023	9.811	36.763	25.437	25.437	287.703	287.703	34.166	34.166	34.166	253.537	253.537	4.269	249.268
Volume Flow cum/hr	246.753	242.917	6.913	236.003	235.987	243.109	42.052	42.052	10.028	10.028	8.925	28.156	22.832	22.832	258.842	267.953	114304	31.142	31.141	240.742	228.768	3.716	225.054
Enthalpy Gcal/hr	-909.593	-916.402	-38.877	-877.525	-877.511	-896.095	-117.865	-117.865	-37.895	-37.895	-33.726	-66.903	-84.688	-84.688	-962.199	-948.175	-96.982	-114.928	-114.927	-845.196	-845.173	-13.402	-831.771
Mass Flow tons/hr																							
GLUCAN	0	0	0	0	0	0	1.215	1.215	0	0	0	1.215	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0.625	0.625	0	0	0	0.625	0	0	0	0	0	0	0	0	0	0	0
ARABINAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	1.226	1.226	0	0	0	1.226	0	0	0	0	0	0	0	0	0	0	0
LIG-SOL	0.106	0.106	0	0.106	0.106	0.106	0.01	0.01	0	0	0	0.003	0.007	0.007	0.113	0.113	0.076	0.076	0.076	0.036	0.036	0.036	0
LIGNIN	0	0	0	0	0	0	7.609	7.609	0	0	0	7.609	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	12.831	12.831	0	0	0	12.831	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	2.391	2.391	0	0	0	2.391	0	0	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	6.947	0.676	0.676	0	0	0	0.203	0.473	0.473	0.473	0.473	0	0	0	0.473	0.473	0.473	0
H2O	256.884	256.884	3.853	253.03	253.03	255.608	24.876	24.876	11.023	11.023	9.811	10.406	24.281	24.281	277.311	277.311	33.188	33.188	33.188	244.123	244.123	3.662	240.461
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	1.482	1.482	0	1.482	1.482	1.482	0.144	0.144	0	0	0	0.043	0.101	0.101	1.583	1.583	0	0	0	1.583	1.583	0	1.583
XYLOSE	5.507	5.507	0	5.507	5.507	5.507	0.536	0.536	0	0	0	0.107	0.429	0.429	5.936	5.936	0	0	0	5.936	5.936	0	5.936
ARABINOS	1.206	1.206	0	1.206	1.206	1.206	0.117	0.117	0	0	0	0.035	0.082	0.082	1.288	1.288	0	0	0	1.288	1.288	0	1.288
CELLOBIO	0	0	0	0	0	0	0.041	0.041	0	0	0	0.041	0	0	0	0	0	0	0	0	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0.206	0.206	0	0.206	0.206	0.206	0.02	0.02	0	0	0	0.006	0.014	0.014	0.22	0.22	0.207	0.207	0.207	0.013	0.013	0.013	0
HMF	0.06	0.06	0	0.06	0.06	0.06	0.006	0.006	0	0	0	0.002	0.004	0.004	0.064	0.064	0.001	0.001	0.001	0.063	0.063	0.063	0
AACID	0.67	0.67	0	0.67	0.67	0.67	0.065	0.065	0	0	0	0.02	0.046	0.046	0.716	0.716	0.694	0.694	0.694	0.022	0.022	0.022	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	1.529	1.529	1.529	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
SORBITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CASO4	9.643	9.643	9.643	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CACO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO3-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PRO-ACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BUT-ACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0



Streams	SS33A	SS33B	SS34	SS35	SS36	SS37	SS38	SS39	SS40	SS41	SS42	SS43	SS44	SS45	SS46	SS47	SS48	SS48B	SS49	SS50	SS50B	SS51	SS52
Temperature C	60.9	61	61	61	61	80.2	80.2	63.5	136.9	25	25	757.8	130	142.6	65	64.9	64.9	64.9	100	100	117.6	100	100
Pressure atm	2.5	3.5	3.5	3.5	0.425	0.425	0.425	27	27	1	1	40	39.9	39.9	5	3.5	3.5	5	0.918	0.918	1.2	0.001	0.001
Vapor Frac	0	0	0	0	0	1	0	0	0	1	1	1	1	0.013	0.013	0.042	0	0	1	0	0	1	0
Mole Flow kmol/hr	12160.39	12160.39	6063.911	6096.478	6096.478	1724.971	4371.507	4371.507	4371.507	98.84	93.898	93.898	93.898	4424.365	4424.365	1358.674	3065.691	3065.691	2452.553	613.138	613.138	578.189	34.949
Mass Flow tons/hr	249.268	249.268	122.071	127.197	127.197	34.255	92.942	92.942	92.942	0.22	0.209	0.209	0.209	93.15	93.15	27.08	66.07	66.07	48.704	17.366	17.366	11.51	5.856
Volume Flow cum/hr	225.066	225.058	111.142	114.031	114.046	117176.9	86.712	82.439	87.075	2419.593	2298.613	200.252	79.373	138.85	392.396	475.477	58.823	58.82	81174.43	14.655	14.266	1.67E+07	3.195
Enthalpy Gcal/hr	-831.771	-831.764	-412.182	-419.582	-419.59	-98.876	-302.767	-302.697	-296.726	0	0	0.486	0.07	-296.628	-303.204	-89.394	-213.81	-213.807	-140.207	-47.36	-47.36	-33.076	-8.892
Mass Flow tons/hr																							
GLUCAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIG-SOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	240.461	240.461	120.231	120.231	120.231	34.255	85.975	85.975	85.975	0	0	0	0	85.975	85.975	25.793	60.183	60.183	48.704	11.479	11.479	11.478	0.001
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0.22	0.209	0.209	0.209	0.117	0.117	0.117	0	0	0	0	0	0	0
GLUCOSE	1.583	1.583	1.583	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOSE	5.936	5.936	0	5.936	5.936	0	5.936	5.936	5.936	0	0	0	0	0.175	0.175	0.175	0	0	0	0	0	0	0
ARABINOS	1.288	1.288	0.258	1.031	1.031	0	1.031	1.031	1.031	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLOBIO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HMF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ETHANOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AIR	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	5.835	5.835	0	5.835	5.835	0	5.834	5.834	0.032	5.803
ARABITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	1.048	1.048	0.996	0.052	0.052	0	0.052	0.052	0	0.052
SORBITOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CASO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CACO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NO3-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OH-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
PRO-ACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BUT-ACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

## A4. Xylitol Utility Consumption Summary

Table 69 - Xylitol Utility Summary

Area Name	Area	Utility	Equipment ID	Equipment Description	Equipment classification	Quantity (tons/h or kW)
Pretreatment	100	LPS	R-101	Dilute Acid Tank Reactor	Reactor	57.07
Purification	400	LPS	S-402	Evaporation Column	Separator	40.95
CHP Plant	500	Electricity	P-501	Boiler Feed Water Pump	Pump	480.54
CHP Plant	500	Electricity	C-501	Boiler Feed Air Blower	Compressor	329.40
WWT	600	Electricity	P-601	WW Nutrient Feed Pump	Pump	0.49
WWT	600	Electricity	C-601	Activated Sludge Air Blowers	Compressor	57.60
WWT	600	Electricity	C-602	Natural Gas Blower	Compressor	8.01
Pretreatment	100	Electricity	P-101	Dilute Acid Feed Pump	Pump	9.49
Pretreatment	100	Electricity	P-102	Mixed Hydrolysate	Pump	4.29
Pretreatment	100	Electricity	P-103	Mixed Sugar Hydrolysate Pump	Pump	2.43
Pretreatment	100	Electricity	P-104	Hydrolysate Filter Press Feed Pump	Pump	2.47
Pretreatment	100	Electricity	P-105	Cellulignin Slurry Pump	Pump	1.05
Pretreatment	100	Electricity	P-106	Wash Water Feed Pump	Pump	1.00
Pretreatment	100	Electricity	P-107	Neutralised Hydrolysate Pump	Pump	0.00
Detoxification	200	Electricity	P-201	Activated Carbon Feed Pump	Pump	35.46
Detoxification	200	Electricity	P-202	Wastewater Works Feed Pump	Pump	0.94
Detoxification	200	Electricity	P-203	SMB Chromatography Feed Pumps 1	Pump	11.34
Detoxification	200	Electricity	P-204	Evaporation Column Feed Pump	Pump	9.44
Catalytic Reaction	300	Electricity	C-301	Hydrogen Gas Feed Compressor	Compressor	473.07
Purification	400	Electricity	P-401	Vacuum Evaporator Feed Pump	Pump	4.55
Purification	400	Electricity	P-402	Evaporative Crystalliser Feed Pump	Pump	0.21
Purification	400	Electricity	S-403	Crystallisation and Drying Unit	Separator	6475.47
Pretreatment	100	Electricity	P-108	Recovered Sugars Pump	Pump	0.47
Catalytic Reaction	300	Electricity	P-301	Reactor Feed Pump	Pump	113.05
Pretreatment	100	Electricity	A-101	Acid and Water Tank Mixer (VT-102 & VT-103)	Agitator	8.30

Area Name	Area	Utility	Equipment ID	Equipment Description	Equipment classification	Quantity (tons/h or kW)
Pretreatment	100	Electricity	A-102	DA Tank Reactor Agitator (R-101)	Agitator	10.11
Pretreatment	100	Electricity	A-103	Lime Neutralisation Agitator (R-102)	Agitator	4.28
Pretreatment	100	Electricity	A-104	Wash water Tank Mixer (VT-105)	Agitator	3.57
Pretreatment	100	Electricity	A-105	Recover Sugar Tank Mixer (VT-106)	Agitator	45.30
Catalytic Reaction	300	Electricity	A-301	Catalytic Reactor Mixers (x3) (R-301)	Agitator	14.86
WWT	600	Electricity	A-601	Activated Sludge Mixers (x4) (R-602)	Agitator	108.91
Pretreatment	100	Electricity	CONV-101	Bagasse and Trash Conveyor	Conveyor	0.37
Pretreatment	100	Electricity	CONV-102	Bypassed Biomass Conveyor	Conveyor	0.42
Pretreatment	100	Electricity	CONV-103	Lime Conveyor	Conveyor	0.01
Pretreatment	100	Electricity	CONV-104	Cellulignin Conveyor	Conveyor	0.21
Pretreatment	100	Electricity	CONV-105	Gypsum Conveyor	Conveyor	0.05
CHP Plant	500	Electricity	CONV-501	Combined Boiler Feed Conveyor	Conveyor	0.67
WWT	600	Electricity	CONV-601	Solids Belt press	Conveyor	0.27
WWT	600	Cooling Water	E-601	Anaerobic Digester Cooling Jacket	Exchanger	48.57
CHP Plant	500	Cooling Air	E-505	Turbine Outlet Steam Cooler	Exchanger	10.53
Catalytic Reaction	300	Cooling Air	E-302	Pre-feed Hydrogen Gas Cooler	Exchanger	311.64

## A5. Xylitol Equipment Sizing Summaries

### A5.1 Pumps

Equipment ID	Equipment Type	Description	Quantity	Units of capacity	Duty (kW)	Inlet Flow (m <sup>3</sup> /h)	Inlet Flow (L/s)
P-101	Pump	Sulphuric Acid Pump	1	L/s	0.28	7	2
P-102	Pump	Dilution Water Pump	1	L/s	9.22	234	65
P-103	Pump	Dilute Acid Feed Pump	2	L/s	9.49	241	67
P-104	Pump	Dilute Acid Hydrolysate Pump	2	L/s	4.29	299	83
P-105	Pump	Hemicellulosic Hydrolysate Pump	2	L/s	2.43	243	68
P-106	Pump	Lime Tank Exit Pump	2	L/s	2.47	247	69
P-107	Pump	Cellulignin Slurry Pump	2	L/s	1.05	42	12
P-108	Pump	Wash Water Pump	2	L/s	1.00	10	3
P-109	Pump	Neutralised Hydrolysate Pump	2	L/s	0.00	236	66
P-110	Pump	Recovered Sugars Pump	2	L/s	0.47	23	6
P-201	Pump	Activated Carbon Feed Pump	1	L/s	35.46	230	64
P-202	Pump	Wastewater Works Feed Pump	1	L/s	0.94	30	8
P-203	Pump	SMB Chromatography Feed Pumps 1	1	L/s	11.34	227	63
P-204	Pump	Evaporation Column Feed Pump	1	L/s	9.44	115	32
P-301	Pump	Reactor Feed Pump	2	L/s	113.05	87	24
P-401	Pump	Crystalliser Feed Pump	2	L/s	4.53	60	17
P-402	Pump	Dryer Feed Pump	2	L/s	0.21	15	4
P-501	Pump	Boiler Feed Water Pump	2	L/s	480.54	203	56
P-601	Pump	Anaerobic Digester Feed Pump	2	L/s	0.49	33	9
P-602	Pump	Aerobic Digester Feed Pump	2	L/s	N/A	33	9
P-603	Pump	Solids Slurry Recycle Pump	2	L/s	N/A	12	3
P-604	Pump	RAS Pump	2	L/s	N/A	12	3
P-605	Pump	WAS Pump	2	L/s	N/A	0	0

## A5.2. Conveyors

Conveyor										
Streams	Area	Equipment ID	Name	Vol. flow (m <sup>3</sup> /h)	Mass flow (ton/h)	Number	Material	Residence Time (h)	P (kW) = m x kW/m	Length (m)
SS1	100	CONV-101	Bagasse and Trash Conveyor	90	113.5	2	304 SS	0.067	0.37	11.93
SS3	100	CONV-102	Bypassed Biomass Conveyor	48	52.2	2	304 SS	0.167	0.42	16.11
SS13	100	CONV-103	Lime Conveyor	2	5.5	2	304 SS	0.083	0.01	0.25
SS23	100	CONV-104	Cellulignin Conveyor	28	10.7	2	304 SS	0.167	0.21	9.39
SS16	100	CONV-105	Gypsum Conveyor	7	15.0	2	304 SS	0.167	0.05	2.30
503	500	CONV-501	Combined Boiler Feed Conveyor	69	89.0	2	304 SS	0.167	0.67	23.11
620+621	600	CONV-601	Solids Belt press	32	35.9	2	304 SS	0.167	0.27	10.78

## A5.3. Compressors, Blowers and Turbines

Equipment ID	Equipment Description	Classification	Quantity	Units of capacity	Duty (kW)	Inlet Flow (m <sup>3</sup> /h)	Inlet Flow (L/s)
C-301	Hydrogen-Argon Gas Feed Compressor	Compressor	1	kW	473.07	1281	356
C-501	Boiler Feed Air Blower	Blower	2	m <sup>3</sup> /h	57.60	627389	174275
C-502	Natural Gas Blower	Blower	1	m <sup>3</sup> /h	8.01	1157	321
C-601	Activated Sludge Blowers	Blower	2	m <sup>3</sup> /h	329.40	3388.719	941
C-602	Ammonia Compressor	Compressor	1	kW		614.5182	171
TURB-501	Condensing Steam Turbine Stage 1	Turbine	1	kW	-5494		
TURB-502	Condensing Steam Turbine Stage 2	Turbine		kW	-3966		
TURB-503	Condensing Steam Turbine Stage 3	Turbine		kW	-1317		
TURB-504	Condensing Steam Turbine Stage 4	Turbine		kW	-20		

## A5.4. Heat Exchangers

Equipment	Description	Type	Duty Q (kW)	U (W/m <sup>2</sup> K)	T1 (°C) hot fluid, in	T2 (°C) hot fluid, out	t1 (°C) cold fluid, in	t2 (°C) cold fluid, out	Tlm (°C)	A (m <sup>2</sup> ) = Q/(U x Ft x ΔTlm)
E-101	DA Preheater	Shell and Tube	7919	1150	80	50	40	69	10.5	364.6
E-102	DA Hydrolysate Cooler	Shell and Tube	-22225	1150	131	60	50	105	16.7	641.2
E-103	Lime Tank Cooling Jacket	Jacketed Vessel	Included							
E-201	Evaporator Heater	Shell and Tube	22225	1150	131	60	50	105	16.7	641.2
E-301	Catalytic Feed Heater	Shell and Tube	6486	1150	135	65	63	130	3.3	957.0
E-302	Pre-feed Hydrogen Gas Cooler	Air-cooled Exchanger	-393	200	641	130	30	35	280.8	3.9

Equipment	Description	Type	Duty Q (kW)	U (W/m <sup>2</sup> K)	T1 (°C) hot fluid, in	T2 (°C) hot fluid, out	t1 (°C) cold fluid, in	t2 (°C) cold fluid, out	Tlm (°C)	A (m <sup>2</sup> ) = Q/(U x Ft x ΔTlm)
E-303	Catalytic Product Cooler	Shell and Tube	393	1150	135	65	63	130	3.3	57.9
E-501	Biomass Feed Preheater	Shell and Tube	256	2750	233	170	35	40	162.3	0.3
E-502	Boiler Air Preheater	Shell and Tube	2753	30	270	149	27	40	170.3	299.3
E-503	Flue gas heat exchanger 1	Shell and Tube	-164888	30	870	270	30	35	477.2	6398.4
E-504	Flue gas heat exchanger 2	Shell and Tube	-30590	200	270	149	27	40	170.3	498.9
E-505	Turbine Outlet Steam Cooler	Shell and Tube	-157	160	341	340	30	35	308.0	1.8
E-601	Anaerobic Digester Cooling Jacket	Shell and Tube	-510	1150	52	40	20	30	21.0	11.7

#### A5.5. Separators and Filters

Flash drums											
Equipment	Description	Flow (m3/h)	Configuration (horizontal/vertical)	Material	Vessel diameter, D <sub>c</sub> (m)	Vessel length, L <sub>c</sub> (m)	wall thickness, t <sub>w</sub> (m)	SS Density, ρ (kg/m <sup>3</sup> )	Shell mass (kg) = π x D <sub>c</sub> x L <sub>c</sub> x t <sub>w</sub> x ρ		
S-201	Inhibitor Evaporator	274.8574997	vertical	304 SS	4.44	17.76	0.0187	8000	37049		
S-204	Water Evaporator	114.8508	vertical	304 SS	3.32	13.28	0.0187	8000	20708		
S-403	Xylitol Dryer/Evaporator	14.62164	vertical	304 SS	1.67	6.68	0.0187	8000	5240		
Columns											
S-202	Activated Carbon Adsorption Column	230.4054	vertical	304 SS	4.19	16.74	0.0187	8000	32938		
S-203	WBA SMB Chromatography	114.8508	vertical	304 SS	1.66	6.64	0.0187	8000	5177		
S-401	SAC SMB Chromatography Unit	386.9636	vertical	304 SS	2.49	9.95	0.0187	8000	11635		
Separators											
Equipment	Equipment Name	tons/h	m <sup>3</sup> /h	#	Material	Purchased Cost	Cost Year	Purchase Cost in Base Year, C1	Scaling Exponent	kg/h, S1	kg/h, S2
S-101	Cellulignin-Sugars Filter Press	324.18	285	18	304 SS	\$178 520	2009	\$20 000	0.8	28630	324180
S-102	Gypsum Rotary Filter Drum	277.29	243	35	Carbon Steel	\$1 091 840	2016				
S-103	Wash Water Tank Filter	63.41	52	3	304 SS	\$32 611	2009	\$20 000	0.8	28630	63413
S-402	Vacuum Crystalliser	67.97	60	N/A	304 SS	\$200 000	2016	\$200 000	N/A		
S-501	Particulate Bag Filter (Vent Scrubber)	1.15	624681	N/A	304 SS		1998	\$99 000	0.78	18523	1147
F-601	Membrane Bioreactor Filter Modules	53.99	3860.9 6	N/A	PA		2010	\$5 248 750	0.6	393100	53993
S-603	Belt press	35.94	32.33	1	304 SS	\$50 000	2016	\$50 000			



## A5.6. Reactors and Vessels

Vessels								
Equipment ID	Type	Description	Classification	Quantity	Units of capacity	Inlet Flow (m <sup>3</sup> /h)	Residence Time (h)	Volume (m <sup>3</sup> )
VT-101	Vessel	Bagasse and Trash Holding Tank	Vertical Tank	1	m <sup>3</sup>	89.5	0.5	44.8
VT-102	Vessel	Sulphuric Acid Storage Tank	Vertical Tank	1	m <sup>3</sup>	6.9	0.5	3.4
VT-103	Vessel	Dilution Water Storage Tank	Vertical Tank	1	m <sup>3</sup>	38.8	0.5	19.4
VT-104	Vessel	Wash Water Storage Tank	Vertical Tank	1	m <sup>3</sup>	8.9	1	8.9
VT-105	Vessel	Wash Water Filtration Tank	Vertical Tank	3	m <sup>3</sup>	51.0	1	51.0
VT-106	Vessel	Hydrolysate Sugar Holding Tank	Vertical Tank	1	m <sup>3</sup>	258.8	2.5	647.1
VT-501	Vessel	Bagasse and Cellulignin Holding Tank	Vertical Tank	1	m <sup>3</sup>	69.3	1	69.3
VT-502	Vessel	Condensate Storage Tank	Vertical Tank	1	m <sup>3</sup>	72.1	1	72.1
VT-601	Vessel	Wastewater Nutrient Mix Tank	Vertical Tank	1	m <sup>3</sup>	33.3	1	33.3
VT-602	Vessel	Ammonia Pressure Vessel	Vertical Tank	1	m <sup>3</sup>	-	1	0.0
VT-603	Vessel	Solids Slurry Holding Tank	Vertical Tank	1	m <sup>3</sup>	44.0	8	351.8
VT-604	Vessel	WAS Holding Tank	Vertical Tank	1	m <sup>3</sup>	32.3	1	32.3

Reactors									
Equipment ID	Type	Description	Quantity	Units of capacity	Inlet Flow (m <sup>3</sup> /h)	Temp (°C)	Pressure (atm)	Residence Time (h)	Volume (m <sup>3</sup> )
R-101	Reactor	Dilute Acid Tank Reactor	1	m <sup>3</sup> /reactor	289.0	130	-	0.5	216.7
R-102	Reactor	Lime Neutralising Tank	1	m <sup>3</sup> /reactor	244.6		-	0.25	61.2
R-301	Reactor	Catalytic Xylitol Reactor	3	m <sup>3</sup> /reactor	89.2	135	-	2.5	334.6
R-501	Reactor	CHP Boiler	1	kg/h	-	870	-	N/A	N/A
R-601	Reactor	Anaerobic Digester	1	m <sup>3</sup> /reactor	33.3	52	1	360	17960.1
R-602	Reactor	Activated Sludge Reactor	1	m <sup>3</sup> /reactor	32.4	30	1	48	1555.9

## A5.7. Agitators

Agitators										
Type	Stream	Area	Equipment ID	Name	Vol. flow (m <sup>3</sup> /h)	Mass flow (ton/h)	#	Material	Residence Time (h)	P (kW) = 0.04-0.1 kW/m <sup>3</sup>
Vessel	SS7	100	A-101	Acid and Water Tank Mixer (VT-102 & VT-103)	237	263	1	316 SS	0.5	8.30
Reactor	SS2 + SS8	100	A-102	DA Tank Reactor Agitator (R-101)	289	324	1	316 SS	0.5	10.11
Reactor	S5+SS13	100	A-103	Lime Neutralisation Agitator (R-102)	245	277	1	316 SS	0.25	4.28
Vessel	SS20B + SS22	100	A-104	Wash water Tank Mixer (VT-105)	51	62	3	316 SS	1	3.57
Vessel	SS18 + SS23B	100	A-105	Recover Sugar Tank Mixer (VT-106)	259	288	1	316 SS	2.5	45.30
Reactor	SS39	300	A-301	Catalytic Reactor Mixers (x3) (R-301)	85	96	3	316 SS	2.5	14.86
Reactor	610	600	A-601	Activated Sludge Mixers (x4) (R-602)	32	36	4	316 SS	48	108.91

## A6. Xylitol Equipment Cost Summary

Table 70 - Xylitol Equipment Sizing and Cost Summary

Area	Equipment ID	Equipment Description	Specialised Classification	Quantity	Units of ca	Duty (kW)	Inlet Flow (L/s)	Equipment Cost, Ce (\$)	Installed Cost in 200	Installed Cost in 2016 (\$)	Residence Time (h)	Volume (m3)
100	P-101	Sulphuric Acid Pump	Pump	1	L/s	0.28	1.94	\$ 7 273.73	\$ 16 729.58	\$ 17 080.10	N/A	N/A
100	P-102	Dilution Water Pump	Pump	1	L/s	9.22	64.90	\$ 15 708.39	\$ 36 129.29	\$ 36 886.28	N/A	N/A
100	P-103	Dilute Acid Feed Pump	Pump	2	L/s	9.49	66.84	\$ 31 889.60	\$ 73 346.09	\$ 74 882.86	N/A	N/A
100	P-104	Dilute Acid Hydrolysate Pump	Pump	2	L/s	4.29	83.19	\$ 35 828.06	\$ 82 404.54	\$ 84 131.11	N/A	N/A
100	P-105	Hemicellulosic Hydrolysate Pump	Pump	2	L/s	2.43	67.53	\$ 32 057.63	\$ 73 732.54	\$ 75 277.41	N/A	N/A
100	P-106	Lime Tank Exit Pump	Pump	2	L/s	2.47	68.54	\$ 32 303.88	\$ 74 298.92	\$ 75 855.66	N/A	N/A
100	P-107	Cellulignin Slurry Pump	Pump	2	L/s	1.05	11.68	\$ 17 563.86	\$ 40 396.87	\$ 41 243.28	N/A	N/A
100	P-108	Wash Water Pump	Pump	2	L/s	1.00	2.79	\$ 14 835.90	\$ 34 122.57	\$ 34 837.52	N/A	N/A
100	P-109	Neutralised Hydrolysate Pump	Pump	2	L/s	0.00	65.56	\$ 31 576.68	\$ 72 626.37	\$ 74 148.07	N/A	N/A
100	P-110	Recovered Sugars Pump	Pump	2	L/s	0.47	6.34	\$ 15 972.23	\$ 36 736.13	\$ 37 505.84	N/A	N/A
200	P-201	Activated Carbon Feed Pump	Pump	1	L/s	35.46	64.00	\$ 15 598.37	\$ 35 876.24	\$ 36 627.93	N/A	N/A
200	P-202	Wastewater Works Feed Pump	Pump	1	L/s	0.94	8.20	\$ 8 268.57	\$ 19 017.71	\$ 19 416.17	N/A	N/A
200	P-203	SMB Chromatography Feed Pumps 1	Pump	1	L/s	11.34	62.96	\$ 15 471.44	\$ 35 054.32	\$ 36 329.89	N/A	N/A
200	P-204	Evaporation Column Feed Pump	Pump	1	L/s	9.44	31.90	\$ 11 548.26	\$ 26 561.01	\$ 27 117.52	N/A	N/A
300	P-301	Reactor Feed Pump	Pump	2	L/s	113.05	24.20	\$ 21 050.89	\$ 48 417.06	\$ 49 431.51	N/A	N/A
400	P-401	Crystalliser Feed Pump	Pump	2	L/s	4.53	16.78	\$ 19 014.89	\$ 43 734.26	\$ 44 650.60	N/A	N/A
400	P-402	Dryer Feed Pump	Pump	2	L/s	0.21	4.12	\$ 15 272.94	\$ 35 127.77	\$ 35 863.78	N/A	N/A
500	P-501	Boiler Feed Water Pump	Pump	2	L/s	480.54	56.39	\$ 29 322.10	\$ 67 440.83	\$ 68 853.88	N/A	N/A
600	P-601	Anaerobic Digester Feed Pump	Pump	2	L/s	0.49	9.24	\$ 16 847.56	\$ 38 749.39	\$ 39 561.29	N/A	N/A
600	P-602	Aerobic Digester Feed Pump	Pump	2	L/s	N/A	9.24	\$ 16 847.53	\$ 38 749.33	\$ 39 561.22	N/A	N/A
600	P-603	Solids Slurry Recycle Pump	Pump	2	L/s	N/A	3.23	\$ 14 984.71	\$ 34 464.83	\$ 35 186.95	N/A	N/A
600	P-604	RAS Pump	Pump	2	L/s	N/A	3.23	\$ 14 984.71	\$ 34 464.83	\$ 35 186.95	N/A	N/A
600	P-605	WAS Pump	Pump	2	L/s	N/A	0.00	\$ 13 800.79	\$ 31 741.81	\$ 32 406.88	N/A	N/A
300	C-301	Hydrogen-Argon Gas Feed Compressor	Compressor	1	kW	473.07	355.72	\$ 1 166 496.13	\$ 1 866 393.81	\$ 1 905 499.21	N/A	N/A
500	C-501	Boiler Feed Air Blower	Blower	2	m3/h	57.60	174274.72	\$ 4 266 089.65	\$ 6 825 743.43	\$ 6 968 759.01	N/A	N/A
500	C-502	Natural Gas Blower	Blower	1	m3/h	8.01	321.30	\$ 17 628.24	\$ 28 205.18	\$ 28 796.14	N/A	N/A
600	C-601	Activated Sludge Blowers	Blower	2	m3/h	329.40	941.31	\$ 72 951.34	\$ 116 722.15	\$ 119 167.76	N/A	N/A
600	C-602	Ammonia Compressor	Compressor	1	kW		170.70	\$ 490 000.00	\$ 784 000.00	\$ 800 426.67	N/A	N/A
500	TURB-501	Condensing Steam Turbine Stage 1	Turbine	1	kW	-5494.12		\$ 1 681 511.44	\$ 3 026 720.59	\$ 3 090 137.59	N/A	N/A
500	TURB-502	Condensing Steam Turbine Stage 2	Turbine	1	kW	-3966.07				\$ 6 557 451.80	N/A	N/A
500	TURB-503	Condensing Steam Turbine Stage 3	Turbine	1	kW	-1316.98					N/A	N/A
500	TURB-504	Condensing Steam Turbine Stage 4	Turbine	1	kW	-20.40					N/A	N/A
100	E-101	DA Preheater	Heater	1	m2	7918.97		\$ 78 571.12	\$ 172 856.47	\$ 176 478.23	N/A	N/A
100	E-102	DA Hydrolysate Cooler	Cooler	1	m2	-22224.83		\$ 131 433.26	\$ 289 153.17	\$ 295 211.62	N/A	N/A
100	E-103	Lime Tank Cooling Jacket	Cooler	1	m2						N/A	N/A
200	E-201	Evaporator Heater	Heater	1	m2	22224.83		\$ 131 433.26	\$ 289 153.17	\$ 295 211.62	N/A	N/A
300	E-301	Catalytic Feed Heater	Heater	1	m2	6486.07		\$ 197 726.50	\$ 434 998.29	\$ 444 112.54	N/A	N/A
300	E-302	Prefeed Hydrogen Gas Cooler	Cooler	1	m2	-392.66		\$ 24 234.35	\$ 53 315.56	\$ 54 432.65	N/A	N/A
300	E-303	Catalytic Product Cooler	Cooler	1	m2	392.66		\$ 30 002.16	\$ 66 004.76	\$ 67 387.72	N/A	N/A
500	E-501	Biomass Feed Preheater	Heater	1	m2			\$ 24 011.67	\$ 52 825.66	\$ 53 932.49	N/A	N/A
500	E-502	Boiler Air Preheater	Heater	1	m2	2753.14		\$ 67 065.18	\$ 147 543.40	\$ 150 634.78	N/A	N/A
500	E-503	Flue gas heat exchanger 1	Cooler	1	m2	-164888.06		\$ 1 722 420.24	\$ 3 789 324.54	\$ 3 868 719.91	N/A	N/A
500	E-504	Flue gas heat exchanger 2	Cooler	1	m2	-30590.47		\$ 103 495.91	\$ 227 691.00	\$ 232 461.66	N/A	N/A
500	E-505	Turbine Outlet Steam Cooler	Cooler	1	m2	-157.22		\$ 24 091.42	\$ 53 001.13	\$ 54 111.63	N/A	N/A
500	E-506	Steam Condenser	Cooler	1	m2						N/A	N/A
600	E-601	Anaerobic Digester Cooling Jacket	Cooler	1	m2	-509.76		\$ 24 883.41	\$ 54 743.51	\$ 55 890.52	N/A	N/A

Area	Equipment ID	Equipment Description	Specialised Classification	Quantity	Units of ca	Duty (kW)	Inlet Flow	(Equipment Cos'	Installed Cost in	Installed Cost in	Residence Time (h)	Volume (m3)
100	VT-101	Bagasse and Trash Holding Tank	Vertical Tank	1	m3	N/A	25	25030	45054	45998	0.5	44.8
100	VT-102	Sulphuric Acid Storage Tank	Vertical Tank	1	m3	N/A	2	8324	14983	15296	0.5	3.4
100	VT-103	Dilution Water Storage Tank	Vertical Tank	1	m3	N/A	11	16168	29102	29712	0.5	19.4
100	VT-104	Wash Water Storage Tank	Vertical Tank	1	m3	N/A	2	11479	20663	21096	1	8.9
100	VT-105	Wash Water Filtration Tank	Vertical Tank	3	m3	N/A	14	80827	145489	148538	1	51.0
100	VT-106	Hydrolysate Sugar Holding Tank	Vertical Tank	1	m3	N/A	72	134958	242924	248014	2.5	647.1
500	VT-501	Bagasse and Cellulignin Holding Tank	Vertical Tank	1	m3	N/A	19	32212	57981	59196	1	69.3
500	VT-502	Condensate Storage Tank	Vertical Tank	1	m3	N/A	20	32973	59352	60595	1	72.1
600	VT-601	Wastewater Nutrient Mix Tank	Vertical Tank	1	m3	N/A	9	21273	38292	39094	1	33.3
600	VT-602	Ammonia Pressure Vessel	Vertical Tank	1	m3	N/A	0	5000	9000	9189	1	0.0
600	VT-603	Solids Slurry Holding Tank	Vertical Tank	1	m3	N/A	12	89819	161675	165062	8	351.8
600	VT-604	WAS Holding Tank	Vertical Tank	1	m3	N/A	9	20953	37715	38505	1	32.3
100	CONV-601	Bagasse and Trash Conveyor	Conveyer	2	m		25	107686	183065	186901	0.07	6.0
100	CONV-602	Bypassed Biomass Conveyor	Conveyer	2	m		13	117375	199538	203719	0.17	8.1
100	CONV-603	Lime Conveyor	Conveyer	2	m		0	80585	136995	139865	0.08	0.1
100	CONV-604	Cellulignin Conveyor	Conveyer	2	m		8	101774	173015	176641	0.17	4.7
100	CONV-105	Gypsum Conveyor	Conveyer	2	m		2	85346	145089	148129	0.17	1.2
500	CONV-501	Combined Boiler Feed Conveyor	Conveyer	2	m		19	133612	227141	231900	0.17	11.6
600	CONV-601	Solids Beltpress	Conveyer	2	m		9	105001	178501	182241	0.17	5.4
100	HOP-101	Bagasse and Trash Hopper	Hopper	1	m3	N/A	25	37539	67571	68986	1	89.5
100	HOP-102	Lime Hopper	Hopper	1	m3	N/A	0	6871	12368	12627	1	1.5
500	CYC-501	Bag House Filter	Filter	1		N/A					N/A	N/A
100	F-101	Cellulignin-Sugars Filter Press	Filter Press	18	m3	189.6779	79	140182	238309	243302	N/A	285.1
100	F-102	Gypsum Rotary Filter Drum	Filter Press	35	m3			1091840	1856128	1856128	N/A	N/A
100	F-103	Cellulignin Filter Press	Filter Press	3	m3						N/A	N/A
200	S-201	Inhibitor Evaporator	Flash Drum	1	shell mass, kg		76	534962	1069924	1092342	N/A	N/A
200	S-202	Activated Carbon Adsorption Column	Column	1						970993	N/A	N/A
200	S-203	WBA SMB Chromatography	Column	8	shell mass, kg					1801670	N/A	N/A
200	S-204	Water Evaporator	Flash Drum	1	shell mass, kg		32	332117	664235	678152	N/A	N/A
400	S-401	SAC SMB Chromatography Unit	Column	8	shell mass, kg					3348324	N/A	N/A
400	S-402	Xylitol Crystalliser	Crystalliser	1	m					227245	N/A	N/A
400	S-403	Xylitol Dryer/Evaporator	Column	1	shell mass, kg		4	113622	227245	232006	N/A	N/A
600	S-601	H2S Scrubber	Column	1							N/A	N/A
600	F-601	MBR Filtration Module	Filter	1	kg/h		1072	1519722	1519722	1551564	N/A	N/A
600	F-602	Air Filter	Filter	1							N/A	N/A
100	R-101	Dilute Acid Tank Reactor	Reactor	1	m3/reacto	N/A	80	2122521	3183781	3250489	0.5	216.7
100	R-102	Lime Neutralising Tank	Reactor	1	m3/reacto	N/A	68	805154	1207732	1233036	0.25	61.2
300	R-301	Catalytic Xylitol Reactor	Reactor	3	m3/reacto	N/A	25	8947375	17894750	18269687	2.5	334.6
500	R-501	CHP Boiler	Reactor	1	kg/h	N/A		26122660	47020787	48005985	N/A	N/A
600	R-601	Anaerobic Digester	Reactor	1	m3/reacto	N/A	9	3647000	6199899	6329802	360	17960.1
600	R-602	Activated Sludge Reactor	Reactor	1	m3/reacto	N/A	9	1083036	1841162	1879738	48	1555.9
100	A-101	Acid and Water Tank Mixer (VT-102 & VT-103)	Agitator	1	kW	8.300579	66	24135	36202	36961	0.5	118.6
100	A-102	DA Tank Reactor Agitator (R-101)	Agitator	1	kW	10.11338	80				0.5	144.5
100	A-103	Lime Neutralisation Agitator (R-102)	Agitator	1	kW	4.28086	68				0.25	61.2
100	A-104	Washwater Tank Mixer (VT-105)	Agitator	3	kW	3.568331	14	56294	84441	86210	1	51.0
100	A-105	Recover Sugar Tank Mixer (VT-106)	Agitator	1	kW	45.29623	72	69263	103894	106071	2.5	647.1
300	A-301	Catalytic Reactor Mixers (x3) (R-301)	Agitator	3	kW	14.86381	24				2.5	212.3
600	A-601	Activated Sludge Mixers (x4) (R-602)	Agitator	4	kW	108.9098	9	605275	907912	926935	48	1555.9

## A7. Xylitol Pricing from Literature and Industry

Table 71 - Xylitol Pricing According to Literature

Year	Demand (tons/annum)	Average Cost Price (\$/ton)	Original Reference
1993	10000		(Bahndorf & Kienle, 2004)
1995		7200	(Rao, et al., 2007) (Canilha, et al., 2012)
2000		4500	(Rao, et al., 2007) (Canilha, et al., 2012)
2003		3800	(Rao, et al., 2007) (Canilha, et al., 2012)
2005		3500	(Rao, et al., 2007) (Canilha, et al., 2012)
2007		3300	(Rao, et al., 2007) (Canilha, et al., 2012)
2009		4500	(Shetty, et al., 2009)
2012	125000	5000	(S.R., et al., 2012)
2013	160000		
2014	116100		(Grand View Research, 2017)
2015	122300		(Grand View Research, 2017)
2016	190900		Xylitol - A Global Market Overview   Sugar Substitute   Flavor, 2017
2016	190000	4500	(Biddu, et al., 2016)
2017		4500	Food Bioconversion edited by Alexandru Mihai Grumezescu, Alina
2020	242000		Food Bioconversion edited by Alexandru Mihai Grumezescu, Alina Pg 429
2022	266500		Xylitol - A Global Market Overview   Sugar Substitute   Flavor, 2017
2017		9056	
2017		10440	Danisco Sweetener
2017		9656	

Table 72 - Xylitol Pricing by Industry and Country

<b>Manufacturers by Country</b>	<b>Price (\$/ton)</b>
<b>Canada</b>	<b>\$14 985</b>
Honson Pharmatech Group Ltd.	\$3 000
Xylitol Canada Inc.	\$11 985
<b>China</b>	<b>\$60 076</b>
Anhui BBKA Pharmaceutical Co., Ltd.	\$3 500
Awell Ingredients Co., Ltd.	\$2 750
Ever Exceed Imp. & Exp. Co., Ltd.	\$3 250
Foodchem International Corporation	\$2 500
Health Chemical Co. Ltd.	\$3 750
Henan Boom Gelatin Co., Ltd.	\$3 000
Jiachem Dentbio Co., Ltd.	\$4 000
Jiaxing Sunlong Industrial & Trading Co., Ltd.	\$3 675
Linyi Kemele Co., Ltd.	\$4 051
Sanhe Sweet Food Co., Ltd.	\$10 000
SanHerb BioScience Co., Ltd.	\$8 000
Soyoung Biotech. Inc.	\$4 200
Synchem International Co., Ltd.	\$3 500
Xi'an Lyphar Biotech Co., Ltd.	\$3 900
<b>Finland</b>	<b>\$0</b>
Danisco Sweeteners Oy	\$0
<b>Hungary</b>	<b>\$2 500</b>
REPCEOLAJ KFT	\$2 500
<b>USA</b>	<b>\$18 447</b>
Xylitol USA, Inc.	\$9 981
	\$8 466

A8. Xylitol Discount Cash Flow Sheet

DCFROR Worksheet	2016		2017		2018		2019		2020		2021	
Year	-2	-1	0	1	2	3	4	5				
Fixed Capital Investment	\$ 20 958 084.19	\$ 125 748 505	\$ 62 874 253									
Land	\$ -	\$ -	\$ -									
Working Capital			\$ 10 479 042.09									
Product Price (\$/t)			\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538
Succinic Acid Sales			\$ -	\$ 85 763 299	\$ 128 644 948	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598
Electricity price (\$/ kWh)			\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales			\$ -	\$ 597 391	\$ 896 086	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782
Total Annual Revenue			\$ -	\$ 86 360 690	\$ 129 541 035	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379
Annual Manufacturing Cost												
Feedstock Price (\$/ ton)												
Feedstock cost												
Other Variable Costs												
Fixed Operating Costs												
Total Product Cost			\$ -	\$ 39 626 622	\$ 59 439 933	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244
Annual Depreciation												
Plant Writedown			0.00%	20.00%	20.00%	20.00%	20.00%	20.00%	20.00%	20.00%	20.00%	20.00%
Depreciation Charge			\$ -	\$ 41 916 168	\$ 41 916 168	\$ 41 916 168	\$ 41 916 168	\$ 41 916 168	\$ 41 916 168	\$ 41 916 168	\$ 41 916 168	\$ 41 916 168
Remaining Value			\$ 209 580 842	\$ 167 664 673	\$ 125 748 505	\$ 83 832 337	\$ 41 916 168	\$ -	\$ -	\$ -	\$ -	\$ 0
Net Revenue (R-COM-dk)	(\$20 958 084)	(\$125 748 505)	(\$73 353 295)	\$ 4 817 899	\$ 28 184 933	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967
Losses Forward				\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income				\$ 4 817 899	\$ 28 184 933	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967	\$ 51 551 967
Income Tax				\$ 1 349 012	\$ 7 891 781	\$ 14 434 551	\$ 14 434 551	\$ 14 434 551	\$ 14 434 551	\$ 14 434 551	\$ 14 434 551	\$ 14 434 551
Annual Cash Income			(\$220 059 884)	\$ 45 385 056	\$ 62 209 320	\$ 79 033 585	\$ 79 033 585	\$ 79 033 585	\$ 79 033 585	\$ 79 033 585	\$ 79 033 585	\$ 79 033 585
Discount Factor		1.0970	1.0000	0.9116	0.8310	0.7575	0.6905	0.6295				
Annual Present Value (Discounted Cash Flow)			(\$220 059 884)	\$ 41 371 974	\$ 51 694 246	\$ 59 867 593	\$ 54 573 923	\$ 49 748 334				
Cumulative Discounted Cash Flow			(\$220 059 884)	(\$178 687 910)	(\$126 993 664)	(\$67 126 071)	(\$12 552 148)	\$37 196 186				
Total Capital Investment + Interest		\$ 137 946 110	\$ 73 353 295									
Net Present Worth			\$ 405 346 162									

DCFROR Worksheet	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033
Year	6	7	8	9	10	11	12	13	14	15	16	17
Fixed Capital Investment												
Land												
Working Capital												
Product Price (\$/t)	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538
Succinic Acid Sales	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598
Electricity price (\$/ kWh)	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782
Total Annual Revenue	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379
Annual Manufacturing Cost												
Feedstock Price (\$/ ton)												
Feedstock cost												
Other Variable Costs												
Fixed Operating Costs												
Total Product Cost	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244
Annual Depreciation												
Plant Writedown												
Depreciation Charge												
Remaining Value												
Net Revenue (R-COM-dk)	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136
Losses Forward	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136
Income Tax	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078
Annual Cash Income	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058
Discount Factor	0.5738	0.5231	0.4768	0.4347	0.3962	0.3612	0.3292	0.3001	0.2736	0.2494	0.2273	0.2072
Annual Present Value (Discounted Cash Flow)	\$ 38 615 024	\$ 35 200 569	\$ 32 088 030	\$ 29 250 711	\$ 26 664 276	\$ 24 306 542	\$ 22 198 072	\$ 20 198 072	\$ 18 412 099	\$ 16 784 046	\$ 15 299 951	\$ 13 947 084
Cumulative Discounted Cash Flow	\$75 811 210	\$111 011 779	\$143 099 809	\$172 350 520	\$199 014 797	\$223 321 338	\$245 198 072	\$265 676 696	\$284 088 794	\$300 872 840	\$316 172 791	\$330 119 875
Total Capital Investment + Interest												
Net Present Worth												



DCFROR Worksheet	2034	2035	2036	2037	2038	2039	2040	2041
Year	18	19	20	21	22	23	24	25
Fixed Capital Investment								
Land								
Working Capital								
Product Price (\$/t)	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538	\$ 4 538
Succinic Acid Sales	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598	\$ 171 526 598
Electricity price (\$/ kWh)	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782	\$ 1 194 782
Total Annual Revenue	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379	\$ 172 721 379
Annual Manufacturing Cost								
Feedstock Price (\$/ ton)								
Feedstock cost								
Other Variable Costs								
Fixed Operating Costs								
Total Product Cost	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244	\$ 79 253 244
Annual Depreciation								
Plant Writedown								
Depreciation Charge								
Remaining Value								
Net Revenue (R-COM-dk)	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136
Losses Forward	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136	\$ 93 468 136
Income Tax	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078	\$ 26 171 078
Annual Cash Income	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058	\$ 67 297 058
Discount Factor	0.1889	0.1722	0.1570	0.1431	0.1305	0.1189	0.1084	0.0988
Annual Present Value (Discounted Cash Flow)	\$ 12 713 841	\$ 11 589 646	\$ 10 564 855	\$ 9 630 679	\$ 8 779 106	\$ 8 002 831	\$ 7 295 197	\$ 6 650 134
Cumulative Discounted Cash Flow	\$342 833 716	\$354 423 361	\$364 988 216	\$374 618 895	\$383 398 000	\$391 400 831	\$398 696 028	\$405 346 162
Total Capital Investment + Interest								
Net Present Worth								

## A9. Xylitol Cumulative Cash Flow Sheet

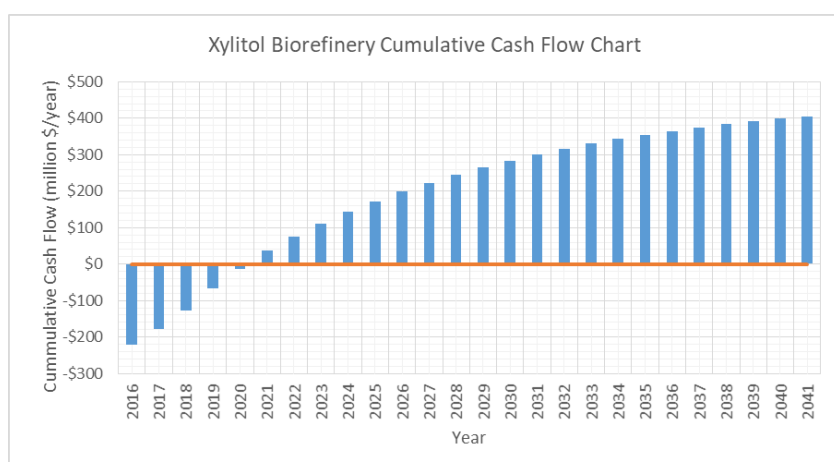


Figure 62 - Xylitol Cumulative Cash Flow

Table 73 - Xylitol 25-Year Cumulative Cash Flow

Year	Financial Year	Cumulative cash flow (\$ million/year)
2016	0	-\$220.1
2017	1	-\$178.7
2018	2	-\$127.0
2019	3	-\$67.1
2020	4	-\$12.5
2021	5	\$37.2
2022	6	\$75.8
2023	7	\$111.0
2024	8	\$143.1
2025	9	\$172.4
2026	10	\$199.0
2027	11	\$223.3
2028	12	\$245.5
2029	13	\$265.7
2030	14	\$284.1
2031	15	\$300.9
2032	16	\$316.2
2033	17	\$330.1
2034	18	\$342.9
2035	19	\$354.5
2036	20	\$365.0
2037	21	\$374.6
2038	22	\$383.4
2039	23	\$391.4
2040	24	\$398.7
2041	25	\$405.4

## Appendix B. Citric Acid (CA) Calculation Summary

Appendix B lists the calculations that were used to conduct the techno-economic analyses for the CA-electricity scenario. Calculations include the specification of components (Table 74) in Aspen Plus<sup>®</sup>, the configuration of process flow sheets (Appendix B2), mass and energy balance flows (Appendix B3), utility consumption (Appendix B4), as well as equipment sizing and costing to determine profitability indicators (Appendices B5-B9).

### B1. CA Component Summary

Table 74 - CA Aspen Component Summary

Component ID	Type	Component name	Alias
GLUCAN	Solid	Dilactic-Acid	C6H10O5
XYLAN	Solid	Glutaric-Acid	C5H8O4
ARABINAN	Solid	Glutaric-Acid	C5H8O4
ACETATE	Solid	Acetic-Acid	C2H4O2-1
ASH	Solid	Calcium-Oxide	CAO
LIG-SOL	Conventional	Vanillin	C8H8O3-D1
LIGNIN	Solid	Vanillin	C8H8O3-D1
CELLULOS	Solid	Dilactic-Acid	C6H10O5
EXTRACT	Conventional	Dextrose	C6H12O6
H2SO4	Conventional	Sulphuric-Acid	H2SO4
H2O	Conventional	Water	H2O
O2	Conventional	Oxygen	O2
N2	Conventional	Nitrogen	N2
CO2	Conventional	Carbon-Dioxide	CO2
H2	Conventional	Hydrogen	H2
GLUCOSE	Conventional	Dextrose	C6H12O6
XYLOSE	Conventional	Xylose	C5H10O5-D2
ARABINOS	Conventional	Arabinose	C5H10O5-D1
CELLOBIO	Conventional	Cellobiose	C12H22O11-D1
GLUCOLIG	Conventional	Dilactic-Acid	C6H10O5
XYLOLIG	Conventional	Glutaric-Acid	C5H8O4
ARABOLIG	Conventional	Glutaric-Acid	C5H8O4
FURFURAL	Conventional	Furfural	C5H4O2
HMF	Conventional	5-Hydroxymethylfurfural	C6H6O3-N5
AACID	Conventional	Acetic-Acid	C2H4O2-1
NH3	Conventional	Ammonia	H3N
AMM-SULF	Solid	Ammonium-Sulphate	(NH4)2SO4
LIME	Solid	Calcium-Oxide	CAO
ETHANOL	Conventional	Ethanol	C2H6O-2
AIR	Conventional	Air	AIR
CASO4	Solid	Calcium-Sulphate	CASO4
CH4	Conventional	Methane	CH4
BIOMASS	Solid	Ethyl-Cyanoacetate	C5H7NO2

CACO3	Conventional	Calcium-Carbonate-Calcite	CACO3
HCO3-	Conventional	HCO3 <sup>-</sup>	HCO3-
H+	Conventional	H <sup>+</sup>	H+
NH4+	Conventional	NH4 <sup>+</sup>	NH4+
NO3-	Conventional	NO3 <sup>-</sup>	NO3-
OH-	Conventional	OH <sup>-</sup>	OH-
CA-ANHYD	Solid	Citric-Acid	C6H8O7
Component ID	Type	Component name	Alias
CITRICA	Solid	Citric-Acid-Monohydrate	C6H10O8
CA(OH)2	Solid	Calcium-Hydroxide	CA(OH)2
CELLULAS	Solid		CHXNXOXSX
PRO-ACID	Conventional	Propionic-Acid	C3H6O2-1
BUT-ACID	Conventional	N-Butyric-Acid	C4H8O2-1
CAYEAST	Solid		CHXOXNXPX
NH4NO3	Solid	Ammonium-Nitrate	NH4NO3
KH2PO4	Solid	Potassium-Dihydrogen-Phosphate	KH2PO4
K+	Conventional	K <sup>+</sup>	K+
OXALI-01	Conventional	Oxalic-Acid	C2H2O4
GLUCO-01	Conventional	Gluconic-Acid	C6H12O7
OCTANOL	Conventional	1-Octanol	C8H18O-1
CAC2O-01	Solid	CaC2O4	CAC2O4
NAOH	Solid	Sodium-Hydroxide	NAOH
NA-CITRA	Solid	C6H5Na3O7	C6H5NA3O7





### B2.3. Area 200 Enzymatic Hydrolysis

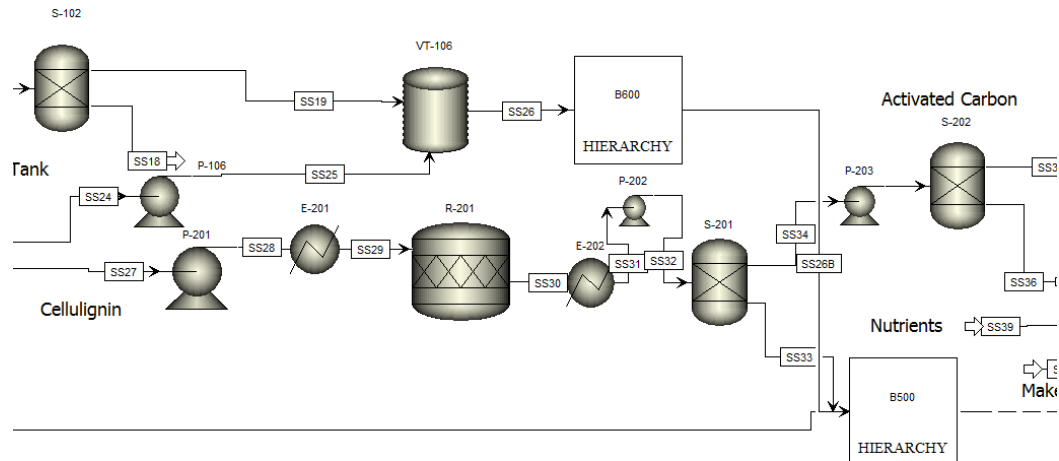


Figure 65 - Aspen CA Enzymatic Hydrolysis Flowsheet (Area 200)

### B2.4. Area 300 Fermentation

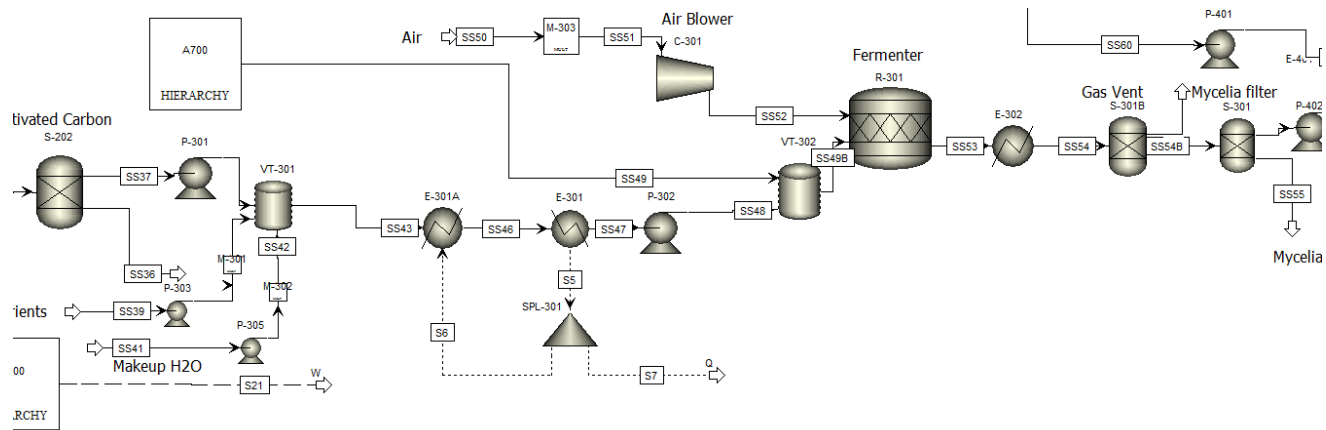


Figure 66 - Aspen CA Fermentation Flowsheet (Area 300)

**B2.5. Area 400 Purification**

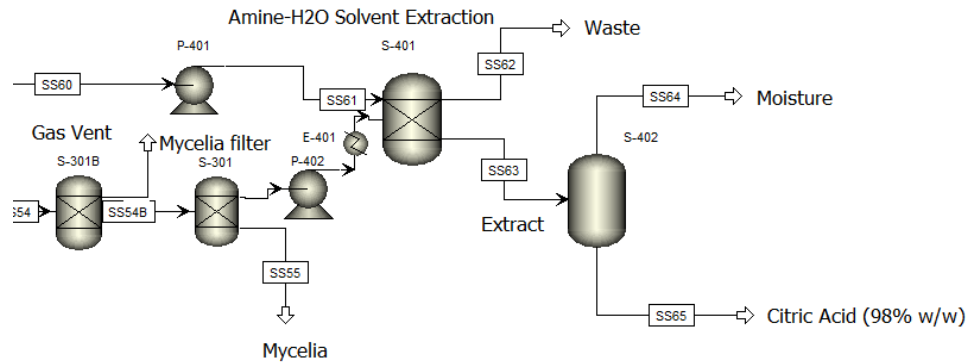


Figure 67 - CA Aspen Purification Flowsheet (Area 400)

**B2.6. Area 500 Combined Heat and Power Plant**

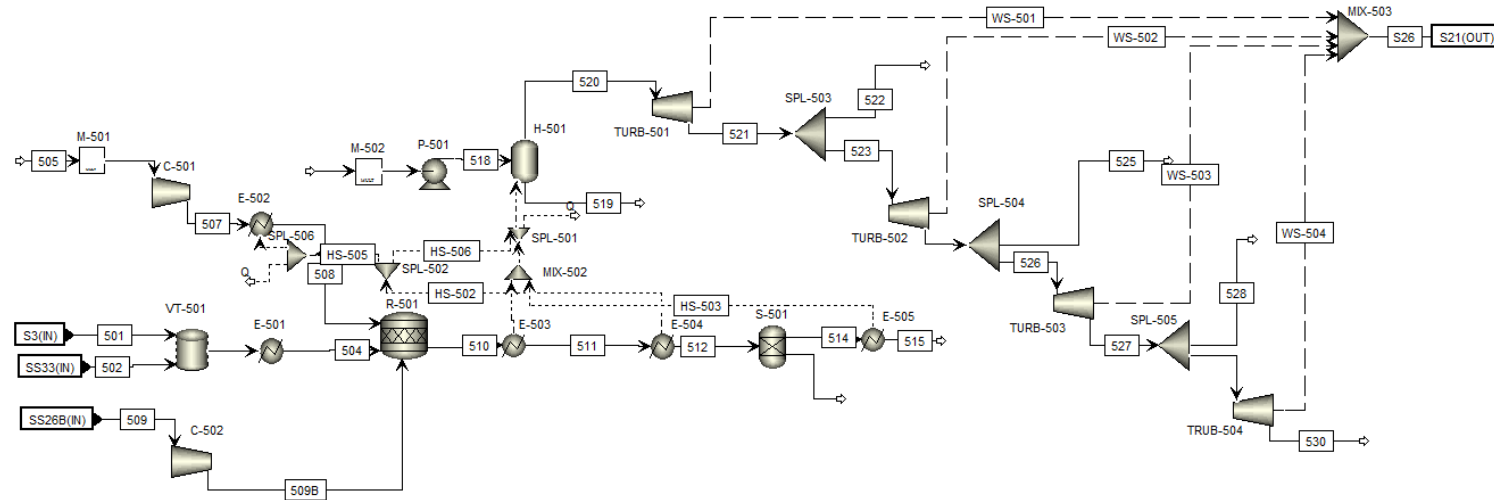


Figure 68 - CA Aspen CHP Plant Flowsheet (Area 500)



### B2.7. Area 600 Water Treatment Plant

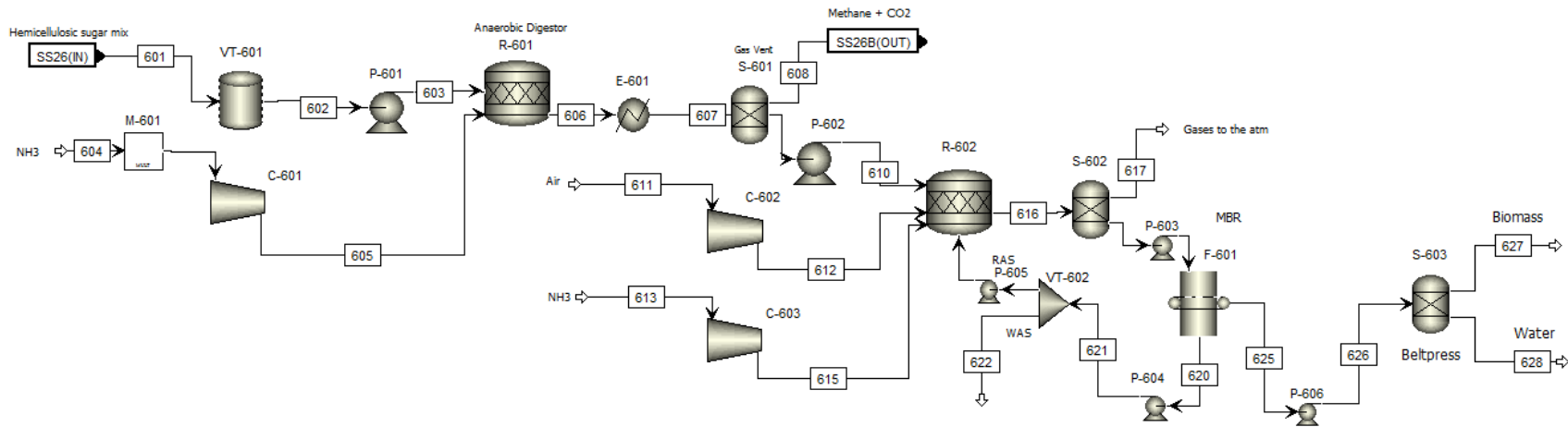
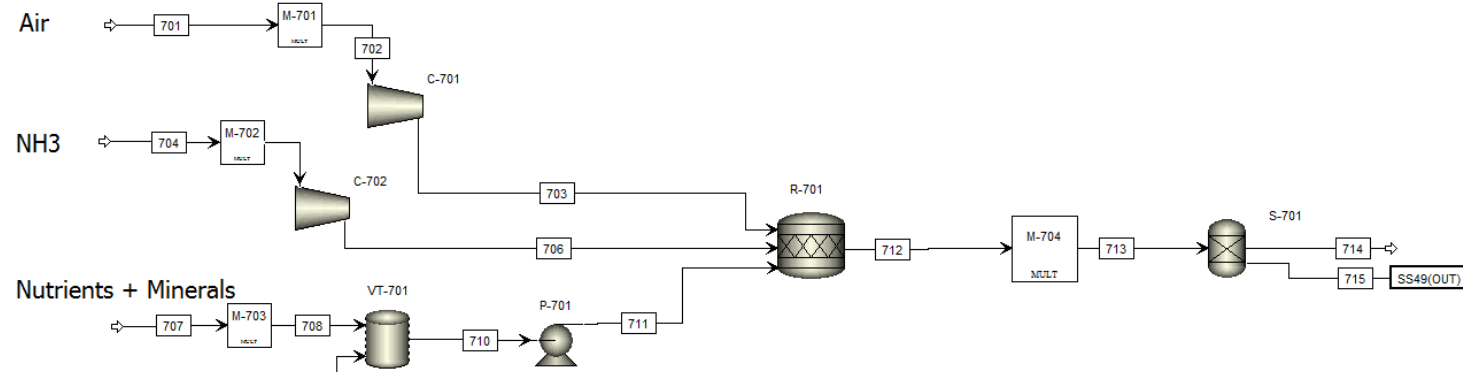


Figure 69 - CA Aspen WWT Flowsheet (Area 600)

### B2.8. Area 700 Candida Oleophila Seed Train



Candida oleophila ATCC 20177

Figure 70 - CA Aspen Candida oleophila Seed Train Flowsheet (Area 700)

## B3. CA Mass and Energy Balance Streams

Streams	S1	S2	S3	S4	S8	S9	SS5	SS6	SS7	SS8	SS9	SS10	SS11	SS12	SS13	SS14
Temperature C	25	25	25	40	25	40.2	40	40	40	124.8	130.6	130.6	60	60	60	25
Pressure atm	1	1	1	1	1	1.2	1	1	2.3	2.3	2.714	3	3	2	1.5	1
Vapor Frac	0	0	0	0	1	1	0	0	0	0	0	0	0	0	0	0
Mole Flow kmol/hr	2862.32	1574.276	1288.044	15992.76	53.268	53.268	14521.42	10372.45	10372.45	10372.45	11897.98	11897.98	11897.98	10694.02	10694.02	83
Mass Flow tons/hr	113.529	62.441	51.088	325.292	1	1	295.365	210.975	210.975	210.975	273.416	273.416	273.416	225.091	225.091	5.131
Volume Flow cum/hr	89.501	49.225	40.275	293.48	1293.722	1132.747	266.48	190.343	190.333	201.173	250.616	250.614	238.675	200.494	200.498	1.411
Enthalpy Gcal/hr	-247.59	-136.174	-111.415	-1098.07	-0.585	-0.578	-997.047	-712.176	-712.168	-696.276	-826.225	-826.223	-842.115	-738.973	-738.975	-12.59
Mass Flow tons/hr																
GLUCAN	2.498	1.374	1.124	0	0	0	0	0	0	0	1.238	1.238	1.238	0	0	0
XYLAN	11.58	6.369	5.211	0	0	0	0	0	0	0	0.637	0.637	0.637	0	0	0
ARABINAN	2.157	1.186	0.971	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	1.362	0.749	0.613	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	2.271	1.249	1.022	0	0	0	0	0	0	0	1.249	1.249	1.249	0	0	0
LIG-SOL	0	0	0	0	0	0	0	0	0	0	0.118	0.118	0.118	0.108	0.108	0
LIGNIN	14.305	7.868	6.437	0	0	0	0	0	0	0	7.75	7.75	7.75	0	0	0
CELLULOS	26.452	14.549	11.904	0	0	0	0	0	0	0	13.069	13.069	13.069	0	0	0
EXTRACT	4.428	2.435	1.992	0	0	0	0	0	0	0	2.435	2.435	2.435	0	0	0
H2SO4	0	0	0	9.433	0	0	8.566	6.118	6.118	6.118	6.118	6.118	6.118	5.576	5.576	0
H2O	48.477	26.662	21.815	315.859	0	0	286.8	204.857	204.857	204.857	230.553	230.553	230.553	210.106	210.106	0
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	0	0	0	0	0	0	0	1.658	1.658	1.658	1.511	1.511	0
XYLOSE	0	0	0	0	0	0	0	0	0	0	6.161	6.161	6.161	5.615	5.615	0
ARABINOS	0	0	0	0	0	0	0	0	0	0	1.348	1.348	1.348	1.229	1.229	0
CELLOBIO	0	0	0	0	0	0	0	0	0	0	0.042	0.042	0.042	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0	0	0	0	0	0	0	0	0	0	0.226	0.226	0.226	0.206	0.206	0
HMF	0	0	0	0	0	0	0	0	0	0	0.065	0.065	0.065	0.06	0.06	0
AACID	0	0	0	0	0	0	0	0	0	0	0.749	0.749	0.749	0.683	0.683	0
NH3	0	0	0	0	1	1	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	5.131
CASO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA-ANHYD	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4NO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
K+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OXALI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Streams	SS14B	SS15	SS16	SS17	SS18	SS19	SS20	SS21	SS22	SS23	SS24	SS25	SS26	SS26B	SS27	SS28
Temperature C	25	78.1	60	60	60	60	60	25	25	25	50.9	50.9	59.7	40	50.9	50.9
Pressure atm	1	1	1	3	3	3	2	1	1	1.5	0.5	1.5	1.5	1.034	0.5	2
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0
Mole Flow kmol/hr	59.286	10753.3	10753.3	10753.3	218.762	10534.54	1203.958	755.346	539.533	539.533	478.85	478.85	11013.39	123.806	1264.642	1264.642
Mass Flow tons/hr	3.665	228.756	228.756	228.756	11.383	217.373	48.325	15	10.714	10.714	10.424	10.424	227.797	2.784	48.615	48.615
Volume Flow cum/hr	1.008	203.211	201.096	201.081	5.425	195.656	38.248	13.646	9.747	9.747	9.06	9.059	204.731	3070.571	38.947	38.946
Enthalpy Gcal/hr	-8.993	-747.968	-751.523	-751.51	-29.575	-721.935	-103.142	-51.566	-36.833	-36.833	-33.593	-33.592	-755.527	-3.658	-106.382	-106.381
Mass Flow tons/hr																
GLUCAN	0	0	0	0	0	0	1.238	0	0	0	0	0	0	0	1.238	1.238
XYLAN	0	0	0	0	0	0	0.637	0	0	0	0	0	0	0	0.637	0.637
ARABINAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	1.249	0	0	0	0	0	0	0	1.249	1.249
LIG-SOL	0	0.108	0.108	0.108	0	0.108	0.01	0	0	0	0.007	0.007	0.115	0	0.003	0.003
LIGNIN	0	0	0	0	0	0	7.75	0	0	0	0	0	0	0	7.75	7.75
CELLULOS	0	0	0	0	0	0	13.069	0	0	0	0	0	0	0	13.069	13.069
EXTRACT	0	0	0	0	0	0	2.435	0	0	0	0	0	0	0	2.435	2.435
H2SO4	0	0	0	0	0	0	0.543	0	0	0	0.38	0.38	0.38	0	0.163	0.163
H2O	0	211.13	211.13	211.13	3.167	207.963	20.448	15	10.714	10.714	9.349	9.349	217.311	0	21.813	21.813
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0.936	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	1.511	1.511	1.511	0	1.511	0.147	0	0	0	0.103	0.103	1.614	0	0.044	0.044
XYLOSE	0	5.615	5.615	5.615	0	5.615	0.546	0	0	0	0.437	0.437	6.052	0	0.109	0.109
ARABINOS	0	1.229	1.229	1.229	0	1.229	0.12	0	0	0	0.084	0.084	1.312	0	0.036	0.036
CELLOBIO	0	0	0	0	0	0	0.042	0	0	0	0	0	0	0	0.042	0.042
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0	0.206	0.206	0.206	0	0.206	0.02	0	0	0	0.014	0.014	0.22	0	0.006	0.006
HMF	0	0.06	0.06	0.06	0	0.06	0.006	0	0	0	0.004	0.004	0.064	0	0.002	0.002
AACID	0	0.683	0.683	0.683	0	0.683	0.066	0	0	0	0.047	0.047	0.729	0	0.02	0.02
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	3.665	0.477	0.477	0.477	0.477	0	0	0	0	0	0	0	0	0	0	0
CASO4	0	7.739	7.739	7.739	7.739	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	1.848	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA-ANHYD	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4NO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
K+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OXALI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Streams	SS29	SS30	SS31	SS32	SS33	SS34	SS35	SS36	SS37	SS38	SS39	SS39B	SS40	SS41	SS41B	SS42	SS43	SS46	SS47	SS48
Temperature C	48	52.9	48	48	48	48	48	48	48	48	20	20	20	25	25	25	34.9	120	30	30
Pressure atm	1	1	1	1.2	1.2	1.2	1.3	1.3	1.3	1.2	1	1.14	1.14	1	1.14	1.14	1.14	1.86	1.86	1.5
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0.014	0	0
Mole Flow kmol/hr	1264.642	1192.512	1192.512	1192.512	400.757	791.755	791.755	143.694	648.062	648.062	25.456	25.456	18.183	1712.118	1712.118	1222.941	1889.186	1889.186	1889.186	1889.186
Mass Flow tons/hr	48.615	48.615	48.615	48.615	19.806	28.809	28.809	2.856	25.952	25.952	5	5	3.571	34	34	24.286	53.81	53.81	53.81	53.81
Volume Flow cum/hr	38.907	28.222	28.169	28.169	14.334	13.839	13.839	2.619	11.221	11.221	0.299	0.299	0.214	30.932	30.932	22.094	33.531	493.277	33.471	33.472
Enthalpy Gcal/hr	-106.463	-106.463	-106.619	-106.619	-35.767	-70.852	-70.852	-9.751	-61.101	-61.101	-7.586	-7.586	-5.419	-116.884	-116.884	-83.488	-150.008	-146.05	-150.216	-150.217
Mass Flow tons/hr																				
GLUCAN	1.238	0.124	0.124	0.124	0.124	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0.637	0.637	0.637	0.637	0.637	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	1.249	1.249	1.249	1.249	1.249	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIG-SOL	0.003	0.003	0.003	0.003	0	0.003	0.003	0.003	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	7.75	7.75	7.75	7.75	7.75	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	13.069	1.291	1.291	1.291	1.291	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	2.435	2.435	2.435	2.435	2.435	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2SO4	0.163	0.163	0.163	0.163	0.163	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	21.813	20.38	20.38	20.38	6.114	14.266	14.266	2.853	11.413	11.413	0	0	0	34	34	24.286	35.698	35.698	35.698	35.698
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0.044	14.395	14.395	14.395	0	14.395	14.395	0	14.395	14.395	4.9	4.9	3.5	0	0	0	17.894	17.894	17.894	17.894
XYLOSE	0.109	0.109	0.109	0.109	0	0.109	0.109	0	0.109	0.109	0	0	0	0	0	0	0.109	0.109	0.109	0.109
ARABINOS	0.036	0.036	0.036	0.036	0	0.036	0.036	0	0.036	0.036	0	0	0	0	0	0	0.036	0.036	0.036	0.036
CELLOBIO	0.042	0.017	0.017	0.017	0.017	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0.006	0.006	0.006	0.006	0.006	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HMF	0.002	0.002	0.002	0.002	0.002	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	0.02	0.02	0.02	0.02	0.02	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CASO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0.085	0.085	0.061	0	0	0	0.061	0.061	0.061	0.061
CA-ANHYD	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4NO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0.015	0.015	0.011	0	0	0	0.011	0.011	0.011	0.011
K+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OXALI-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Streams	SS49B	SS50	SS51	SS52	SS53	SS54	SS54A	SS54B	SS55	SS56	SS57	SS58	SS59	SS60	SS61	SS62	SS63	SS64	SS65	
Temperature C	30	25	25	26.4	25.2	70	70	70	70	70	70	40	40	40	40	40	40	40	71	71
Pressure atm	1.5	1	1	1.014	1	1.1	1.1	1.1	0.9	0.9	1.5	1.5	1	1	1.5	1.5	1.5	0.3	0.3	
Vapor Frac	0	1	1	1	0.228	0.322	1	0	0	0	0	0	0	0	0	0	0	0	1	0
Mole Flow kmol/hr	1889.186	943.333	735.8	735.8	2746.134	2746.134	654.508	2091.626	212.053	1879.574	1879.574	1879.574	1774.56	1267.543	1267.543	317.642	2829.474	2750.544	78.93	
Mass Flow tons/hr	53.81	30	23.4	23.4	78.21	78.21	19.988	58.222	6.546	51.676	51.676	51.676	35.24	25.171	25.171	7.333	69.515	54.622	14.893	
Volume Flow cum/hr	33.472	23067.91	17992.97	17824.06	15396.15	22605.9	16746.81	47.271	5.104	42.168	42.167	41.604	32.217	23.012	23.012	6.466	58.147	258231.4	8.274	
Enthalpy Gcal/hr	-150.217	-0.002	-0.001	0.006	-140.558	-135.768	-2.129	-136.012	-13.453	-122.559	-122.558	-123.663	-120.669	-86.192	-86.192	-22.082	-187.773	-157.859	-0.608	
Mass Flow tons/hr																				
GLUCAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIG-SOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	35.698	0	0	0	39.695	39.695	0	39.695	3.969	35.725	35.725	35.725	35.24	25.171	25.171	6.09	54.807	54.622	0.185	
O2	0	6.988	5.45	5.45	0.195	0.195	0.195	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	23.012	17.95	17.95	17.95	17.95	17.95	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0.933	0.933	0.933	0	0	0	0	0	0	0	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	17.894	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOSE	0.109	0	0	0	0.109	0.109	0	0.109	0	0.109	0.109	0.109	0	0	0	0	0.109	0	0.109	
ARABINOS	0.036	0	0	0	0.036	0.036	0	0.036	0	0.036	0.036	0.036	0	0	0	0.036	0	0	0	
CELLOBIO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HMF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0.91	0.91	0.91	0	0	0	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CASO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0.061	0	0	0	0.66	0.66	0	0.66	0	0.66	0.66	0.66	0	0	0	0.66	0	0	0	
CA-ANHYD	0	0	0	0	17.175	17.175	0	17.175	2.576	14.599	14.599	14.599	0	0	0	0	14.599	0	14.599	
NH4NO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0.011	0	0	0	0.011	0.011	0	0.011	0	0.011	0.011	0.011	0	0	0	0.011	0	0	0	0
K+	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OXALI-01	0	0	0	0	0.537	0.537	0	0.537	0	0.537	0.537	0.537	0	0	0	0.537	0	0	0	0

## B4. CA Utility Consumption Summary

Table 75 - CA Utility Summary

Area Name	Area	Utility	Equipment ID	Equipment Description	Equipment classification	Quantity (tons/h or kW)
Enzymatic Hydrolysis	200	Cooling Water	E-201		Exchanger	9.13
Enzymatic Hydrolysis	200	Cooling Water	E-202		Exchanger	17.28
Purification	400	Cooling Water	E-401		Exchanger	130.49
Purification	400	LPS	S-402		Separator	64.52
Pretreatment	100	LPS	R-101		Reactor	13.14
CA Fermentation	300	LPS	E-302		Exchanger	9.71
Seed Train	700	Electricity	P-701		Pump	0.00
Seed Train	700	Electricity	C-701		Compressor	5.02
Seed Train	700	Electricity	C-702		Compressor	8.40
CHP Plant	500	Electricity	P-501		Pump	463.49
CHP Plant	500	Electricity	C-501		Compressor	319.58
CHP Plant	500	Electricity	C-502		Compressor	16.76
WWT	600	Electricity	P-601		Pump	3.03
WWT	600	Electricity	P-602		Pump	1.69
WWT	600	Electricity	P-603		Pump	1.19
WWT	600	Electricity	P-604		Pump	0.37
WWT	600	Electricity	P-605		Pump	5.19
WWT	600	Electricity	P-606		Pump	2.01
WWT	600	Electricity	C-601		Compressor	0.13
WWT	600	Electricity	C-602		Compressor	501.59
WWT	600	Electricity	C-603		Compressor	8.40
Pretreatment	100	Electricity	P-102		Pump	3.59
Pretreatment	100	Electricity	P-103		Pump	1.59
Pretreatment	100	Electricity	P-104		Pump	0.24
Pretreatment	100	Electricity	P-105		Pump	20.12

Pretreatment	100	Electricity	P-106		Pump	0.45
Enzymatic Hydrolysis	200	Electricity	P-201		Pump	1.73
Enzymatic Hydrolysis	200	Electricity	P-202		Pump	0.28
Enzymatic Hydrolysis	200	Electricity	P-203		Pump	0.07
CA Fermentation	300	Electricity	P-301		Pump	0.02
CA Fermentation	300	Electricity	P-302		Pump	0.19
Purification	400	Electricity	P-401		Pump	0.58
Purification	400	Electricity	P-402		Pump	1.34
CA Fermentation	300	Electricity	C-301		Compressor	8.48
CA Fermentation	300	Electricity	E-301A		Exchanger	4615.66
CA Fermentation	300	Electricity	P-303		Pump	0.00
CA Fermentation	300	Electricity	P-305		Pump	0.22
Pretreatment	100	Electricity	P-101		Pump	12.38
Pretreatment	100	Electricity	A-101	Sulphuric Acid Mixer	Agitator	6.66
Pretreatment	100	Electricity	A-102	Dilute Acid Tank Agitator	Agitator	8.38
Pretreatment	100	Electricity	A-103	Wash water Tank Agitator	Agitator	3.36
Pretreatment	100	Electricity	A-104	Lime Tank Agitator	Agitator	3.53
Pretreatment	100	Electricity	A-105	Recovered Sugar Agitator	Agitator	35.83
Enzymatic Hydrolysis	200	Electricity	A-201	Enzymatic Hydrolysis Agitator	Agitator	185.20
CA Fermentation	300	Electricity	A-301	Nutrient Tank Agitator	Agitator	66.89
CA Fermentation	300	Electricity	A-302	Fermenter Feed Medium Agitator	Agitator	72.76
WWT	600	Electricity	A-602	Aerobic Agitators	Agitator	685.28
Seed Train	700	Electricity	A-701	Seed Nutrient Tank Mixer	Agitator	3.01
Seed Train	700	Electricity	A-702	Seed Fermenter Train Agitator	Agitator	4.52
Pretreatment	100	Electricity	CONV-101	Bagasse and Trash Conveyor	Conveyor	0.37
Pretreatment	100	Electricity	CONV-102	Bypassed Biomass Conveyor	Conveyor	16.85
Pretreatment	100	Electricity	CONV-103	Lime Conveyor	Conveyor	0.34
Pretreatment	100	Electricity	CONV-104	Cellulignin Conveyor	Conveyor	14.45
Pretreatment	100	Electricity	CONV-105	Gypsum Conveyor	Conveyor	7.27

---

Purification	400	Electricity	CONV-401	CA Drying Conveyor	Conveyor	
CHP Plant	500	Electricity	CONV-501	Combined Boiler Feed Conveyor	Conveyor	91.54
WWT	600	Electricity	CONV-601	Solids Belt press	Conveyor	0.32

---



## B5. CA Equipment Sizing Summaries

### B5.1. Pumps

Equipment ID	Equipment Type	Equipment Description	Quantity	Units of capacity	Duty (kW)	Inlet Flow (m <sup>3</sup> /h)	Inlet Flow (L/s)
P-101	Pump	Dilute Acid Feed Pump	2	L/s	12.38	190.34	52.87
P-102	Pump	Dilute Acid Hydrolysate Pump	2	L/s	3.59	250.62	69.62
P-103	Pump	Gypsum Filter Pump	2	L/s	1.59	200.50	55.69
P-104	Pump	Wash Water Feed Pump	2	L/s	0.24	9.75	2.71
P-105	Pump	Neutralised Hydrolysate Pump	2	L/s	20.12	201.09	55.86
P-106	Pump	Recovered Sugar Pump	2	L/s	0.45	9.06	2.52
P-201	Pump	Enzymatic Tank Feed Pump	2	L/s	1.73	204.73	56.87
P-202	Pump	Enzymatic Hydrolysate Pump	2	L/s	0.28	38.95	10.82
P-203	Pump	Activated Carbon Column Feed Pump	2	L/s	0.07	28.17	7.82
P-301	Pump	Detoxified Glucose Rich Feed Pump	2	L/s	0.02		
P-302	Pump	Sterilizer Feed Pump	2	L/s	0.19	11.22	3.12
P-303	Pump	Nutrient Pump	2	L/s	0.00	33.47	9.30
P-305	Pump	Makeup Water Pump	2	L/s	0.22	0.30	0.08
P-401	Pump	Amine-Water Pump	2	L/s	0.58	5.33	1.48
P-402	Pump	Fermentation Broth Pump	2	L/s	1.34	30.93	8.59
P-501	Pump	Boiler Water Feed Pump	2	L/s	463.49	50.14	13.93
P-601	Pump	Anaerobic Digester Feed Pump	2	L/s	3.03	23.01	6.39
P-602	Pump	Activated Sludge Digester Pump	2	L/s	1.69	46.60	12.94
P-603	Pump	MBR Feed Pump	2	L/s	1.19	195.78	54.38
P-604	Pump	MLSS Pump	2	L/s	0.37	204.73	56.87
P-605	Pump	RAS Pump	2	L/s	5.19	203.95	56.65
P-606	Pump	WAS Pump	2	L/s	2.01	756.15	210.04
P-701	Pump	Seed Train Feed Pump	2	L/s	0.00	555.55	154.32

*B5.2. Conveyors*

Conveyor									
Area	Equipment ID	Name	Vol. flow (m <sup>3</sup> /h)	Mass flow (ton/h)	Number	Material	Residence Time (h)	P (kW) = m x kW/m	Length (m)
100	CONV-101	Bagasse and Trash Conveyor	89.50	113.5	2	304 SS	0.067	0.370	11.93
100	CONV-102	Bypassed Biomass Conveyor	49.23	62.4	2	305 SS	0.167	0.441	16.41
100	CONV-103	Lime Conveyor	1.01	3.7	2	306 SS	0.083	0.004	0.17
100	CONV-104	Cellulignin Conveyor	14.33	19.8	2	307 SS	0.167	0.112	4.78
100	CONV-105	Gypsum Conveyor	5.43	11.4	2	307 SS	0.167	0.041	1.81
500	CONV-501	Combined Boiler Feed Conveyor	54.62	70.9	2	308 SS	0.167	0.502	18.21
600	CONV-601	Solids Belt press	0.16	0.2	2	309 SS	0.167	0.001	0.05

*B5.3. Compressors, Blowers and Turbines*

Equipment ID	Type	Description	Classification	Quantity	Units of capacity	Duty (kW)	Inlet Flow (m <sup>3</sup> /h)
C-301	Compressor	Fermenter Air Blower	Blower	2	m <sup>3</sup> /h	8.48	17909
C-501	Compressor	Boiler Air Blowers	Blower	2	m <sup>3</sup> /h	319.58	608689
C-502	Compressor	Natural Gas Compressor	Compressor	2	kW	16.76	2911
C-601	Compressor	Anaerobic Digester NH <sub>3</sub> Compressor	Compressor	2	kW	0.13	106
C-602	Compressor	Aerobic Digester Air Blowers	Blower	2	m <sup>3</sup> /h	501.59	72498
C-603	Compressor	Aerobic Digester NH <sub>3</sub> Compressor	Compressor	2	kW	8.40	1213
C-701	Compressor	Seed Train Air Blower	Compressor	2	m <sup>3</sup> /h	5.02	725
C-702	Compressor	Seed Train NH <sub>3</sub> Compressor	Compressor	2	kW	8.40	1213
TURB-501	Turbine	Condensing Steam Turbine (Stage 1)	Turbine	1	kW	-7789.61	13429
TURB-502	Turbine	HP Steam Turbine (Stage 2)	Turbine				
TURB-503	Turbine	LP Steam Turbine (Stage 3)	Turbine				
TURB-504	Turbine	Steam Turbine (Stage 4)	Turbine				

*B5.4. Heat Exchangers*

Heat Exchangers										
Equipment	Description	Type	Duty Q (kW)	U (W/m <sup>2</sup> K)	T1 (°C) hot fluid, in	T2 (°C) hot fluid, out	t1 (°C) cold fluid, in	t2 (°C) cold fluid, out	Tlm (°C)	A (m <sup>2</sup> ) = Q/(U x Ft x ΔTlm)
E-101	DA Preheater	Shell and Tube	18482	1150	130.6	60	40	125	11.3	1420.7
E-102	DA Hydrolysate Cooler	Shell and Tube	-18482	1150	130.6	60	40	125	11.3	1420.7
E-103		Jacketed Vessel	-4134		78	60				
E-201	Cellulignin preheater		-96	450	51	48	20	30	24.3	8.7
E-202	Cellulignin slurry cooler		-181	450	52.9	48	20	30	25.4	15.9
E-301A	Fermenter Medium Sterilizer	Vessel	4616				35	120		
E-301	Fermenter Medium Sterilizer	Vessel	-4858		120	30				
E-302	Fermenter Broth Cooler	Shell and Tube	5352	2750	233	170	30	70	151.2	12.9
E-401	Extraction Column Pre-cooler	Shell and Tube	-1369	1150	70	40	20	30	28.9	41.3
E-501										
E-502	Boiler Air Preheater	Shell and Tube	18227	30	270	149	27	116	137.4	4422
E-503	Flue gas heat exchanger 1	Shell and Tube	-156228	30	870	270	31	419	333.9	15598
E-504	Flue gas heat exchanger 2	Shell and Tube	-28931	200	270	149	27	116	137.4	1053

E-505	Flue gas heat exchanger 3	Shell and Tube	-21961	160	149	55	30	35	58.7	2340
E-601	Anaerobic Digester Cooling Jacket	Shell and Tube	-6317	1150	66	40	20	30	27.2	202

### B5.5. Separators and Filters

Flash drums									
Equipment	Description	Flow (m <sup>3</sup> /h)	Configuration	Material	Vessel diameter, D <sub>c</sub> (m)	Vessel length, L <sub>c</sub> (m)	wall thickness, t <sub>w</sub> (m)	Stainless Steel Density, ρ (kg/m <sup>3</sup> )	Shell mass (kg) = $\pi \times D_c \times L_c \times t_w \times \rho$
S-402	Evaporator	60.3	vertical	316 SS	2.40	13.33	0.0187	8000	15038
Columns									
S-202	Activated carbon column	13.8	vertical	316 SS	1.30	5.20	0.0187	8000	3182
S-401	Amine-water extraction column	69.3	vertical	316 SS	2.8	11.22	0.0187	8000	14784

Separators											
Equipment ID	Description	tons/h	m <sup>3</sup> /h	#	Material	Purchased Cost	Cost Year	Purchase Cost in Base Year, C1	Scaling Exponent	kg/h, S1	kg/h, S2
S-101	Cellulignin-Sugars Filter Press	273.4	238.6	15	304 SS	\$149 452	2009	\$20 000	0.8	28630	273416
S-102	Gypsum Rotary Filter Drum	228.7	201	29	Carbon Steel	\$900 727	2016				
S-103	Wash Water Tank Filter	59.04	48.00	3	304 SS	\$30 053	2009	\$20 000	0.8	28630	59039
S-201	Lignin Solids Filter	48.62	38.91	2	304 SS	\$24 363	2009	\$20 000	0.8	28630	48615
S-501	Particulate Bag Filter (Vent Scrubber)	0.00	0.00	N/A	304 SS		1998	\$99 000	0.78	18523	0

F-601	Membrane Bioreactor Filter Modules	830.6	756	N/A					
S-603	Belt press	220	200	8	304 SS	\$50 000	2016	\$50 000	

### B5.6. Vessels and Reactors

Vessels									
Equipment ID	Equipment Type	Description	Classification	Quantity	Units of capacity	Inlet Flow (m <sup>3</sup> /h)	Inlet Flow (L/s)	Residence Time (h)	Volume (m <sup>3</sup> )
VT-101	Vertical Tank	Bagasse and Trash Holding Tank	Vessel	1	m <sup>3</sup>	89.50	24.86	0.5	45
VT-102	Vertical Tank	Sulphuric Acid Storage Tank	Storage Tank	1	m <sup>3</sup>	3.04	0.84	0.5	2
VT-103	Vertical Tank	Dilution Water Storage Tank	Storage Tank	1	m <sup>3</sup>	186.20	51.72	0.5	93
VT-104	Vertical Tank	Wash Water Storage Tank	Storage Tank	1	m <sup>3</sup>	9.75	2.71	1	10
VT-105	Vertical Tank	Wash Water Filtration Tank	Storage Tank	3	m <sup>3</sup>	48.00	13.33	1	48
VT-106	Vertical Tank	Recovered Sugar Holding Tank	Storage Tank	2	m <sup>3</sup>	204.73	56.87	2.5	512
VT-301	Vertical Tank	Fermenter Feed Medium Holding Tank 1	Storage Tank	2	m <sup>3</sup>	33.53	9.31	174.6	5855
VT-302	Vertical Tank	Fermenter Feed Medium Holding Tank 2	Storage Tank	2	m <sup>3</sup>	36.12	10.03	174.6	6307
VT-501	Vertical Tank	Bagasse and Cellulignin Holding Tank	Storage Tank	1	m <sup>3</sup>	54.62	15.17	1	55
VT-502	Vertical Tank	Condensate Storage Tank	Storage Tank	1	m <sup>3</sup>	246.29	68.41	1	246
VT-601	Vertical Tank	Wastewater Feed Holding Tank	Storage Tank	1	m <sup>3</sup>	204.73	56.87	1	205

VT-602	Vertical Tank	Ammonia Pressure Vessel	Storage Tank	1	m3			1	0
VT-603	Vertical Tank	Solids Slurry Holding Tank	Storage Tank	1	m3	556	154.32	8	4444
VT-604	Vertical Tank	WAS Holding Tank	Storage Tank	1	m3	0.16	0.04	1	0
VT-701	Vertical Tank	Seed Nutrient Mixing Tank	Vessel	1	m3	2.69	0.75	16	43

<b>Reactors</b>									
Equipment ID	Type	Description	Quantity	Units of capacity	Inlet Flow (m3/h)	Residence Time (h)	Volume (m3)	Temp (°C)	Pressure (atm)
R-101	Reactor	Dilute Acid Tank	1	m3	239.56	0.5	180	130	2.714
R-102	Reactor	Lime Neutralisation Tank	1	m3	201.51	0.25	50	60	1
R-202	Reactor	Enzymatic Hydrolysis Tank (2-stage)	1	m3	38.91	68	2646	48	1
R-301	Reactor	Citric Acid Fermenter	3	m3	36.29	174.6	93	33	1
R-501	Reactor	CHP Boiler (Combustion Chamber)	1	kW				870	0.966
R-601	Reactor	Anaerobic Digester	4	m3	204.73	90	18425	66	1
R-602	Reactor	Activated Sludge Digester	1	m3	203.95	48	9790		1
R-701	Reactor	Seed Fermenter Train	2	m3	2.69	24	97	30	1

<b>Boiler</b>									
Equipment ID	Type	Description	tons/h	m3/h	L/h	# of equipment	Material	Temp (°C)	Pressure (atm)
H-501	Boiler	CHP Boiler (Water Heater)	1	196	195783	1	Carbon Steel	487	64

### B5.7. Agitators

Type	Streams	Area	Equipment ID	Name	Volumetric flow (m3/h)	Mass flow (ton/h)	#	Material	Residence Time (h)	P (kW) = 0.04-0.1 kW/m3
Vessel	SS7	100	A-101	Sulphuric Acid Mixer	190.34	227.80	1	SS304	0.5	6.66
Reactor	S2 + SS7	100	A-102	Dilute Acid Tank Agitator	239.56	273.42	1	SS304	0.5	8.38

Vessel	SS20 + SS22	100	A-103	Wash water Tank Agitator	48.00	273.416	3	SS304	1	3.36
Reactor	SS14B + SS13	100	A-104	Lime Tank Agitator	201.51		1	SS304	0.25	3.53
Vessel	SS19 + SS25	100	A-105	Recovered Sugar Agitator	204.72	227.797	2	SS304	2.5	35.83
Reactor	SS29	200	A-201	Enzymatic Hydrolysis Agitator	38.91	48.615	1	SS316	68	185.20
Vessel	SS43	300	A-301	Nutrient Tank Agitator	33.53	53.81	2	A285C	174.6	409.82
Reactor	SS48 + 3 m3/h	300	A-302	Fermenter Feed Medium Agitator	36.47	56.81	2	A285C	174.6	445.76
Vessel	SS61	400	A-401	Stripping Water Tank Agitator	23.01	35.24	1	SS304	-	-
Vessel	603	600	A-601	Anaerobic Feed Mixer	204.73	227.80	1		90	
Reactor	610	600	A-602	Aerobic Agitators	203.95		4		48	685
Vessel	710	700	A-701	Seed Nutrient Tank Mixer	2.69	3.00	1	SS304	16	3.01
Reactor	710	700	A-702	Seed Fermenter Train Agitator	2.69	3.00	2	SS304	24	4.52

## B6. CA Equipment Cost Summary

Table 76 - CA Equipment Sizing and Cost Summary

Area	Equipmen	Equipmen	Equipment Description	Specialised Classification	Quantity	Units of ca	Duty (kW)	Inlet Flow (L/s)	Purchase Cost, Ce (\$)	Installed Cost (\$)	Installed Cost in 2016 (\$)
300	C-301	Compress	Fermenter Air Blower	Blower	2	m3/h	8	4975	\$ 255 156	\$ 408 250	\$ 416 804
500	C-501	Compress	Boiler Air Blowers	Blower	2	m3/h	320	169080	\$ 4 164 240	\$ 6 662 784	\$ 6 802 386
500	C-502	Compress	Natural Gas Compressor	Compressor	2	kW	17	809	\$ 1 162 334	\$ 1 859 734	\$ 1 898 700
600	C-601	Compress	Anaerobic Digester NH3 Compressor	Compressor	2	kW	0	29	\$ 989 743	\$ 1 583 588	\$ 1 616 768
600	C-602	Compress	Aerobic Digester Air Blowers	Blower	2	m3/h	502	20138	\$ 765 284	\$ 1 224 455	\$ 1 250 110
600	C-603	Compress	Aerobic Digester NH3 Compressor	Compressor	2	kW	8	337	\$ 1 100 456	\$ 1 760 730	\$ 1 797 622
700	C-701	Compress	Seed Train Air Blower	Compressor	2	m3/h	5	201	\$ 26 632	\$ 42 611	\$ 43 504
700	C-702	Compress	Seed Train NH3 Compressor	Compressor	2	kW	8	337	\$ 1 100 456	\$ 1 760 730	\$ 1 797 622
600	F-601	Filter	Membrane Bioreactor Filter Modules	Filtration Modules	1	m2			\$ 10 745 365	\$ 18 267 121	\$ 18 649 861
500	F-501	Filter	Particulate Bag Filter	Filter Press	1	m2				\$ -	\$ -
700	S-701	Flash Drun	Gas Vent	Separator	1	kg					
400	Dryer after	Flash Drurr	GA Dryer	Heater	1	kW			Included		
100	E-101	Heat Exch:	Dilute Acid Tank Heating Jacket	Heater	1	m2	18482		\$ 303 111	\$ 666 844	\$ 680 816
100	E-102	Heat Exch:	Hydrolysate Cooler	Cooler	1	m2	-18482		\$ 303 111	\$ 666 844	\$ 680 816
100	E-103	Heat Exch:	Lime Tank Cooling Jacket	Cooler	1	m2	-4134		\$ 24 000	\$ 52 800	\$ 53 906
200	E-201	Heat Exch:	Cellulignin Slurry Cooler	Cooler	1	kW	-96		\$ 24 621	\$ 54 166	\$ 55 301
200	E-202	Heat Exch:	Enzymatic Cooling Jacket	Cooler	2	kW	-181		\$ 50 542	\$ 111 192	\$ 113 522
300	E-301	Heat Exch:	Feed Medium Sterilizer	Heater	1	m2	4616		\$ 24 000	\$ 52 800	\$ 53 906
300	E-302	Heat Exch:	Fermenter Cooling Jacket	Cooler	1	m2	-4858		\$ 24 000	\$ 52 800	\$ 53 906
400	E-402	Heat Exch:	Precrystallisation Cooler	Cooler	1	m2	5352		\$ 24 987	\$ 54 971	\$ 56 123
500	E-501	Heat Exch:	Air Preheater	Heater	1	m2	-1369		\$ 27 994	\$ 61 586	\$ 62 876
500	E-502	Heat Exch:	Has no purpose			N/A	0		\$ -	\$ -	\$ -
500	E-503	Heat Exch:	Flue Gas Cooler 1	Cooler	1	m2	18227		\$ 1 114 333	\$ 2 451 532	\$ 2 502 898
500	E-504	Heat Exch:	Flue Gas Cooler 2	Cooler	1	m2	-156228		\$ 4 972 256	\$ 10 938 964	\$ 11 168 161
500	E-505	Heat Exch:	Flue Gas Cooler 3	Cooler	1	m2	-28931		\$ 218 832	\$ 481 430	\$ 491 517
600	E-601	Heat Exch:	Anaerobic Digester Cooling Jacket	Cooler	1	m2	-21961		\$ 531 942	\$ 1 170 272	\$ 1 194 792
500	H-501	Heater	CHP Boiler (Water Heater)	Heater	2	kg/h stean	210000		\$ 25 387 356	\$ 26 097 823	\$ 46 976 081
300	S-301	N/A	Gas Vent	N/A		N/A					
100	P-101	Pump	Dilute Acid Feed Pump	Pump	2	L/s	12	53	\$ 28 449	\$ 65 432	\$ 66 803
100	P-102	Pump	Dilute Acid Hydrolysate Pump	Pump	2	L/s	4	70	\$ 32 564	\$ 74 898	\$ 76 467
100	P-103	Pump	Gypsum Filter Pump	Pump	2	L/s	2	56	\$ 29 150	\$ 67 046	\$ 68 451
100	P-104	Pump	Wash Water Feed Pump	Pump	2	L/s	0	3	\$ 14 810	\$ 34 062	\$ 34 776
100	P-105	Pump	Neutralised Hydrolysate Pump	Pump	2	L/s	20	56	\$ 29 191	\$ 67 140	\$ 68 547
100	P-106	Pump	Recovered Sugar Pump	Pump	2	L/s	0	3	\$ 14 745	\$ 33 914	\$ 34 625
100	P-107	Pump	WWT Transfer Pump	Pump	2	L/s		57	\$ 29 442	\$ 67 716	\$ 69 135
200	P-201	Pump	Enzymatic Tank Feed Pump	Pump	2	L/s	2	11	\$ 17 313	\$ 39 819	\$ 40 654
200	P-202	Pump	Enzymatic Hydrolysate Pump	Pump	2	L/s	0	8	\$ 16 424	\$ 37 776	\$ 38 567
200	P-203	Pump	Activated Carbon Column Feed Pump	Pump	2	L/s	0				
300	P-301	Pump	Detoxified Glucose Rich Feed Pump	Pump	2	L/s	0	3	\$ 14 946	\$ 34 376	\$ 35 096
300	P-302	Pump	Sterilizer Feed Pump	Pump	2	L/s	0	9	\$ 16 865	\$ 38 790	\$ 39 602
300	P-303	Pump	Nutrient Pump	Pump	2	L/s	0	0	\$ 13 844	\$ 31 841	\$ 32 508
300	P-304	Pump	Fermenter Seed Feed Pump	Pump	2	L/s	1	1	\$ -	\$ -	\$ -
300	P-305	Pump	Makeup Water Pump	Pump	2	L/s	0	9	\$ 16 655	\$ 38 306	\$ 39 109
300	P-306	Pump	Fermentation Product Pump	Pump	2	L/s	3	14	\$ 18 209	\$ 41 881	\$ 42 758
400	P-401	Pump	Amine-Water Pump	Pump	2	L/s	1	6	\$ 15 988	\$ 36 772	\$ 37 542
400	P-402	Pump	Fermentation Broth Pump	Pump	2	L/s	1	13	\$ 17 928	\$ 41 235	\$ 42 099
500	P-501	Pump	Boiler Water Feed Pump	Pump	2	L/s	463	54	\$ 28 825	\$ 66 298	\$ 67 687
600	P-601	Pump	Anaerobic Digester Feed Pump	Pump	2	L/s	3	57	\$ 29 442	\$ 67 716	\$ 69 135
600	P-602	Pump	Activated Sludge Digester Pump	Pump	2	L/s	2	57	\$ 29 388	\$ 67 593	\$ 69 009
600	P-603	Pump	MBR Feed Pump	Pump	2	L/s	1	210	\$ 64 496	\$ 148 340	\$ 151 448
600	P-604	Pump	MLSS Pump	Pump	2	L/s	0	154	\$ 52 213	\$ 120 090	\$ 122 606
600	P-605	Pump	RAS Pump	Pump	2	L/s	5	154	\$ 52 213	\$ 120 089	\$ 122 605
600	P-606	Pump	WAS Pump	Pump	2	L/s	2	56	\$ 29 157	\$ 67 062	\$ 68 467
700	P-701	Pump	Seed Train Feed Pump	Pump	2	L/s	0	1	\$ 14 117	\$ 32 468	\$ 33 149



100	R-101	Reactor	Dilute Acid Tank	Reactor	1	m3	67	\$	1 781 318	\$	2 671 977	\$	2 727 962	0.5	179.7	
100	R-102	Reactor	Lime Neutralisation Tank	Reactor	1	m3	56	\$	644 081	\$	966 122	\$	986 365	0.25	50.4	
200	R-202	Reactor	Enzymatic Hydrolysis Tank (2-stage)	Reactor	1	m3	11	\$	15 317 482	\$	22 976 223	\$	23 457 629	68	2645.7	
300	R-301	Reactor	Citric Acid Fermenter	Reactor	3	m3	10	\$	1 600 922	\$	2 401 384	\$	2 451 698	174.6	93.2	
500	R-501	Reactor	CHP Boiler (Combustion Chamber)	Reactor	1	kW										
600	R-601	Reactor	Anaerobic Digester	Reactor	4	m3	57	\$	6 633 846	\$	9 950 769	\$	10 159 262	90	18425.4	
600	R-602	Reactor	Activated Sludge Digester	Reactor	1	m3	57	\$	1 099 503	\$	1 649 255	\$	1 683 811	48	9789.7	
700	R-701	Reactor	Seed Fermenter Train	Reactor	2	m3	1	\$	1 085 652	\$	1 628 478	\$	1 662 599	24	96.8	
100	S-101	Separator	Cellulignin Filter Press	Filter Press	14.94521	kg/h	66	\$	122 327	\$	207 955	\$	212 312			
100	S-102	Separator	Gypsum Filter Press	Filter Press	28.5945	kg/h	56	\$	882 242	\$	1 499 811	\$	1 531 235			
100	S-103	Separator	Wash Water Tank Filter	Washwater Tank	3.005322	kg/h	13	\$	35 890	\$	61 013	\$	62 291			
200	S-201	Separator	Lignin Solids Filter	Filter Press	2.436255	kg/h	11	\$	30 724	\$	52 231	\$	53 325			
200	S-202	Separator	Activated carbon column	Column	2	kg (shell mass)		\$	79 544	\$	159 087	\$	162 420			
400	S-401	Separator	Amine-water extraction column	Column	5	kg	19	\$	253 134	\$	506 269	\$	516 876			
400	S-402	Separator	Evaporator	Evaporator	1	kg (shell mass)	17	\$	256 614	\$	513 227	\$	523 981			
600	S-601	Separator	Biogas Vent	N/A	1	N/A										
600	S-602	Separator	Release to atmosphere	N/A												
600	S-603	Separator	Beltpress	Filter Press	7.835809	kg/h		\$	383 750	\$	652 375	\$	666 044			
500	TURB-501	Turbine	Condensing Steam Turbine (Stage 1)	Turbine	2	kW	-7790	0	\$	4 553 811	\$	8 196 860	\$	8 368 603		
500	TURB-502	Turbine	HP Steam Turbine (Stage 2)	Turbine	0	kW			Included							
500	TURB-503	Turbine	LP Steam Turbine (Stage 3)	Turbine	0	kW			Included							
500	TURB-504	Turbine	Steam Turbine (Stage 4)	Turbine	0	kW			Included							
100	VT-101	Vertical Ta	Bagasse and Trash Holding Tank	Vessel	1	m3	25	\$	20 040	\$	36 072	\$	36 828	0.5	44.8	
100	VT-102	Vertical Ta	Sulphuric Acid Storage Tank	Storage Tank	1	m3	1	\$	1 885	\$	3 392	\$	3 463	0.5	1.5	
100	VT-103	Vertical Ta	Dilution Water Storage Tank	Storage Tank	1	m3	52	\$	33 460	\$	60 228	\$	61 490	0.5	93.1	
100	VT-104	Vertical Ta	Wash Water Storage Tank	Storage Tank	1	m3	3	\$	6 902	\$	12 423	\$	12 684	1	9.7	
100	VT-105	Vertical Ta	Wash Water Filtration Tank	Storage Tank	3	m3	13	\$	21 046	\$	37 883	\$	38 677	1	48.0	
100	VT-106	Vertical Ta	Recovered Sugar Holding Tank	Storage Tank	2	m3	57	\$	110 295	\$	198 530	\$	202 690	2.5	511.8	
300	VT-301	Vertical Ta	Fermenter Feed Medium Holding Tank 1	Storage Tank	2	m3	9	\$	607 266	\$	1 093 078	\$	1 115 981	174.6	5854.5	
300	VT-302	Vertical Ta	Fermenter Feed Medium Holding Tank 2	Storage Tank	2	m3	10	\$	639 744	\$	1 151 539	\$	1 175 666	174.6	6306.9	
500	VT-501	Vertical Ta	Bagasse and Cellulignin Holding Tank	Storage Tank	1	m3	15	\$	23 039	\$	41 471	\$	42 339	1	54.6	
500	VT-502	Vertical Ta	Condensate Storage Tank	Storage Tank	1	m3	68	\$	66 100	\$	118 980	\$	121 473	1	246.3	
600	VT-601	Vertical Ta	Wastewater Feed Holding Tank	Storage Tank	1	m3	57	\$	58 081	\$	104 545	\$	106 736	1	204.7	
600	VT-602	Vertical Ta	Ammonia Pressure Vessel	Storage Tank	1	m3		\$	10	\$	18	\$	18	1	0.0	
600	VT-603	Vertical Ta	Solids Slurry Holding Tank	Storage Tank	1	m3	154	\$	500 734	\$	901 320	\$	920 205	8	4444.4	
600	VT-604	Vertical Ta	WAS Holding Tank	Storage Tank	1	m3	0	\$	401	\$	722	\$	737	1	0.2	
700	VT-701	Vertical Ta	Seed Nutrient Mixing Tank	Vessel	1	m3	1	\$	19 489	\$	35 080	\$	35 815	16	43.0	
100	CONV-601	Conveyer	Bagasse and Trash Conveyer	Conveyer	2	m	25	\$	107 686	\$	183 066	\$	186 901	0.07	6.0	
100	CONV-602	Conveyer	Bypassed Biomass Conveyer	Conveyer	2	m	14	\$	118 067	\$	200 714	\$	204 920	0.17	8.2	
100	CONV-603	Conveyer	Lime Conveyer	Conveyer	2	m	0	\$	80 390	\$	136 663	\$	139 526	0.08	0.1	
100	CONV-604	Conveyer	Cellulignin Conveyer	Conveyer	2	m	4	\$	91 085	\$	154 844	\$	158 089	0.17	2.4	
100	CONV-105	Conveyer	Gypsum Conveyer	Conveyer	3	m	2	\$	126 293	\$	214 698	\$	219 197	0.17	0.9	
500	CONV-501	Conveyer	Combined Boiler Feed Conveyer	Conveyer	2	m	15	\$	122 240	\$	207 808	\$	212 162	0.17	9.1	
600	CONV-601	Conveyer	Solids Beltpress	Conveyer	2	m	0	\$	80 125	\$	136 212	\$	139 066	0.17	0.0	
100	HOP-101	Hopper	Bagasse and Trash Hopper	Hopper	1	m3	14	\$	26 412	\$	44 900	\$	45 841	1.00	49.2	
100	HOP-102	Hopper	Lime Hopper	Hopper	1	m3	0	\$	6 408	\$	10 893	\$	11 122	1.00	1.0	
100	A-101	Agitator	Sulphuric Acid Mixer	Agitator	1	kg/h	57	\$	4 590	\$	7 802	\$	7 966			
100	A-102	Agitator	Dilute Acid Tank Agitator	Agitator		kg/h		\$	-	\$	-	\$	-			
100	A-103	Agitator	Washwater Tank Agitator	Agitator	3	kg/h	66	\$	81 262	\$	138 146	\$	141 040			
100	A-104	Agitator	Lime Tank Agitator	Agitator		kg/h		\$	-	\$	-	\$	-			
100	A-105	Agitator	Recovered Sugar Agitator	Agitator	2	kg/h	57	\$	74 174	\$	126 095	\$	128 737			
200	A-201	Agitator	Cellulase Fermenter Agitator	Agitator	1	kg/h	11	\$	-	\$	-	\$	-			
300	A-301	Agitator	Nutrient Tank Agitator	Agitator	2	kg/h	9	\$	15 806	\$	26 870	\$	27 433			
300	A-302	Agitator	Fermenter Feed Medium Agitator	Agitator	2	kg/h	10	\$	16 240	\$	27 608	\$	28 187			
400	A-401	Agitator	Stripping Water Tank Agitator	Agitator	1	kg/h	9	\$	29 174	\$	49 596	\$	50 635			
400	A-402	Agitator	Extractant Tank Mixer	Agitator		kg/h		\$	-	\$	-	\$	-			
600	A-601	Agitator	Anaerobic Feed Mixer	Agitator	1	kg/h	57	\$	-	\$	-	\$	-			
600	A-602	Agitator	Aerobic Agitators	Agitator		kg/h		\$	-	\$	-	\$	-			
700	A-701	Agitator	Seed Nutrient Tank Mixer	Agitator	1	kg/h	1	\$	37 310	\$	63 427	\$	64 756			
700	A-702	Agitator	Seed Fermenter Train Agitator	Agitator	2	kg/h	1	\$	7 689	\$	13 071	\$	13 345			

## B7. CA Pricing from Literature and Industry

Table 77 - CA Pricing According to Literature

Year	CA Price (\$/ton)	World Demand (tons/year)	Reference
1990	-	598000	
1995	1750	-	(SRI Consulting, 2000)
1996	1330	-	(SRI Consulting, 2000)
1997	1215	-	(SRI Consulting, 2000)
1998	1220	-	(SRI Consulting, 2000)
1999	1115	879000	(SRI Consulting, 2000)
2000	1400	-	(SRI Consulting, 2000)
2002	-	-	
2003	-	-	
2004	-	-	
2006	1100	1400000	(ICIS Chemical Business, 2006)
2007	-	1600000	(Berovic & Legisa, 2007)
2009	-	-	
2013	-	-	
2014	-	1800000	
2015	-	2000000	
2016	-	1480567	
2017	-	2000000	

Table 78 - CA Pricing by Industry and Country

<b>Manufacturers by Country</b>	<b>Price (\$/ton)</b>
<b>China</b>	
CBH Qingdao Co., Ltd.	1150
Dalian Richon Chem Co., Ltd.	1000
Guangzhou ZIO Chemical Co., Ltd.	925
Oriental Biotech (Wuxi) Co., Ltd.	950
Qingdao Yuyin Chemical Co., Ltd.	750
Shanghai Xinglu Chemical Technology Co., Ltd.	1750
Tianjin Haitong Chemical Industrial Co., Ltd.	800
Tianjin Yieldstone Chem Co., Ltd.	920
<b>India</b>	
Acuro Organics Ltd.	920
AMA Specialty India	1040
Benzer Multitech India Private Ltd.	980
Deshmukh Chemicals	1070
Doshi and Sons.	1530
Heda Chemicals Pvt. Ltd.	1230
Hemadri Chemicals	950
Manibhadra Chemicals	1460
Mohini Auxichem Private Limited	1230
Radhe Krishna Enterprise	1300
Sarojanand Foods & Trading	1040
Shreenathji Chemicals	920
Wang Pharmaceuticals and Chemicals	1430
<b>Taiwan</b>	
Chung Hwa Chemical Industrial Works, Ltd.	1000
<b>Thailand</b>	
Thai Poly Chemicals Co., Ltd.	1000
<b>Average Market Price</b>	<b>1102</b>

### B8. CA Discount Cash Flow Sheet

DCFROR Worksheet			2016	2017	2018	2019	2020	2021
Year	-2	-1	0	1	2	3	4	5
Fixed Capital Investment	\$ 27 730 570.11	\$ 166 383 421	\$ 83 191 710					
Land	\$ -	\$ -	\$ -					
Working Capital			\$ 13 865 285.05					
Product Price (\$/t)			\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102
Succinic Acid Sales			\$ -	\$ 53 200 152	\$ 79 800 228	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304
Electricity price (\$/ kWh)			\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales			\$ -	\$ 2 065 800	\$ 3 098 699	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599
Total Annual Revenue			\$ -	\$ 55 265 952	\$ 82 898 927	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903
Annual Manufacturing Cost								
Feedstock Price (\$/ ton)								
Feedstock cost								
Other Variable Costs								
Fixed Operating Costs								
Total Product Cost			\$ -	\$ 30 001 274	\$ 45 001 911	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548
Annual Depreciation								
Plant Writedown			0.00%	20.00%	20.00%	20.00%	20.00%	20.00%
Depreciation Charge			\$ -	\$ 55 461 140	\$ 55 461 140	\$ 55 461 140	\$ 55 461 140	\$ 55 461 140
Remaining Value			\$ 277 305 701	\$ 221 844 561	\$ 166 383 421	\$ 110 922 280	\$ 55 461 140	\$ 0
Net Revenue (R-COM-dk)	(\$27 730 570)	(\$166 383 421)	(\$97 056 995)	\$ -30 196 463	\$ -17 564 124	\$ -4 931 785	\$ -4 931 785	\$ -4 931 785
Losses Forward				\$ -	\$ -30 196 463	\$ -47 760 586	\$ -52 692 371	\$ -57 624 156
Taxable Income				\$ -30 196 463	\$ -47 760 586	\$ -52 692 371	\$ -57 624 156	\$ -62 555 941
Income Tax				\$ -	\$ -	\$ -	\$ -	\$ -
Annual Cash Income			(\$291 170 986)	\$ 25 264 678	\$ 37 897 017	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355
Discount Factor	1.0970	1.0000		0.9116	0.8310	0.7575	0.6905	0.6295
Annual Present Value (Discounted Cash Flow)			(\$291 170 986)	\$ 23 030 700	\$ 31 491 385	\$ 38 275 765	\$ 34 891 308	\$ 31 806 115
Cumulative Discounted Cash Flow			(\$291 170 986)	(\$268 140 286)	(\$236 648 901)	(\$198 373 136)	(\$163 481 829)	(\$131 675 714)
Total Capital Investment + Interest	\$ 182 522 612	\$ 97 056 995						
Net Present Worth			\$ 77 227 685					

DCFROR Worksheet	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033
Year	6	7	8	9	10	11	12	13	14	15	16	17
Fixed Capital Investment												
Land												
Working Capital												
Product Price (\$/t)	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102
Succinic Acid Sales	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304
Electricity price (\$/ kWh)	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599
Total Annual Revenue	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903
Annual Manufacturing Cost												
Feedstock Price (\$/ ton)												
Feedstock cost												
Other Variable Costs												
Fixed Operating Costs												
Total Product Cost	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548
Annual Depreciation												
Plant Writedown												
Depreciation Charge												
Remaining Value												
Net Revenue (R-COM-dk)	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355
Losses Forward	\$ -62 555 941	\$ -12 026 586	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income	\$ -12 026 586	\$ 38 502 770	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355
Income Tax	\$ -	\$ 10 780 776	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219
Annual Cash Income	\$ 50 529 355	\$ 39 748 580	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136
Discount Factor	0.5738	0.5231	0.4768	0.4347	0.3962	0.3612	0.3292	0.3001	0.2736	0.2494	0.2273	0.2072
Annual Present Value (Discounted Cash Flow)	\$ 28 993 723	\$ 20 790 993	\$ 17 346 954	\$ 15 813 085	\$ 14 414 845	\$ 13 140 242	\$ 11 978 342	\$ 10 919 182	\$ 9 953 675	\$ 9 073 542	\$ 8 271 232	\$ 7 539 865
Cumulative Discounted Cash Flow	(\$102 681 991)	(\$81 890 998)	(\$64 544 044)	(\$48 730 959)	(\$34 316 114)	(\$21 175 872)	(\$9 197 530)	\$1 721 652	\$11 675 327	\$20 748 869	\$29 020 101	\$36 559 966
Total Capital Investment + Interest												
Net Present Worth												

<b>DCFROR Worksheet</b>	2033	2034	2035	2036	2037	2038	2039	2040	2041
Year	17	18	19	20	21	22	23	24	25
Fixed Capital Investment									
Land									
Working Capital									
Product Price (\$/t)	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102	\$ 1 102
Succinic Acid Sales	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304	\$ 106 400 304
Electricity price (\$/ kWh)	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599	\$ 4 131 599
Total Annual Revenue	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903	\$ 110 531 903
Annual Manufacturing Cost									
Feedstock Price (\$/ ton)									
Feedstock cost									
Other Variable Costs									
Fixed Operating Costs									
Total Product Cost	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548	\$ 60 002 548
Annual Depreciation									
Plant Writedown									
Depreciation Charge									
Remaining Value									
Net Revenue (R-COM-dk)	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355
Losses Forward	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355	\$ 50 529 355
Income Tax	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219	\$ 14 148 219
Annual Cash Income	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136	\$ 36 381 136
Discount Factor	0.2072	0.1889	0.1722	0.1570	0.1431	0.1305	0.1189	0.1084	0.0988
Annual Present Value (Discounted Cash Flow)	\$ 7 539 865	\$ 6 873 168	\$ 6 265 422	\$ 5 711 415	\$ 5 206 395	\$ 4 746 030	\$ 4 326 372	\$ 3 943 821	\$ 3 595 097
Cumulative Discounted Cash Flow	\$36 559 966	\$43 433 134	\$49 698 556	\$55 409 971	\$60 616 366	\$65 362 395	\$69 688 767	\$73 632 588	\$77 227 685
Total Capital Investment + Interest									
Net Present Worth									

### B9. CA Cumulative Cash Flow Sheet

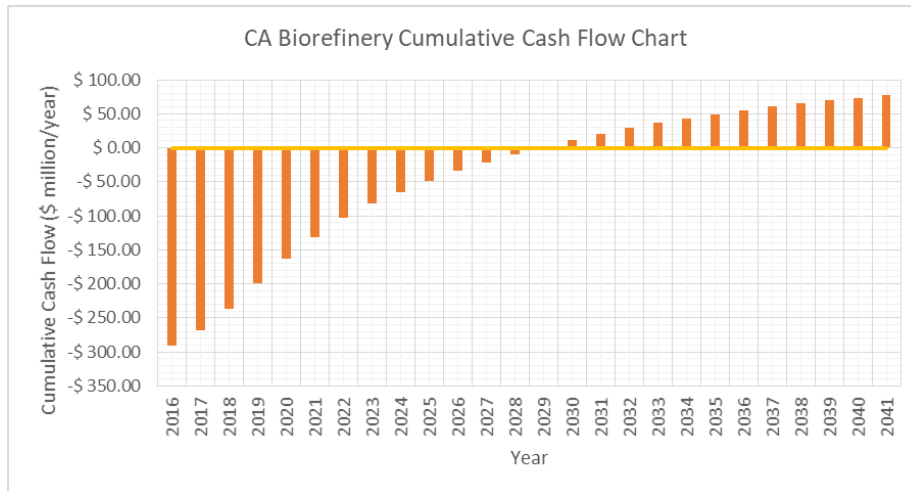


Figure 71 - CA Cumulative Cash Flow

Table 79 - CA 25-Year Cumulative Cash Flow

Year	Financial Year	Cumulative cash flow (\$ million/year)
2016	0	-\$291.17
2017	1	-\$268.14
2018	2	-\$236.65
2019	3	-\$198.38
2020	4	-\$163.49
2021	5	-\$131.69
2022	6	-\$102.70
2023	7	-\$81.91
2024	8	-\$64.56
2025	9	-\$48.75
2026	10	-\$34.33
2027	11	-\$21.19
2028	12	-\$9.22
2029	13	\$1.70
2030	14	\$11.65
2031	15	\$20.73
2032	16	\$29.00
2033	17	\$36.54
2034	18	\$43.41
2035	19	\$49.67
2036	20	\$55.38
2037	21	\$60.59
2038	22	\$65.34
2039	23	\$69.66
2040	24	\$73.61
2041	25	\$77.20

## Appendix C. Glutamic Acid (GA) Calculation Summary

The following appendix specifies the calculations that were used to conduct the techno-economic analyses for the glutamic acid-electricity scenario. Calculations include the specification of components (Table 80) in Aspen Plus<sup>®</sup>, the configuration of process flow sheets (Appendix C2), mass and energy balance flows (Appendix c3), utility consumption (Appendix c4), as well as equipment sizing and costing to determine profitability indicators (Appendices A5-A9).

### C1. GA Component Summary

Table 80 - GA Component Summary

Component ID	Type	Component name	Alias
GLUCAN	Solid	Dilactic-Acid	C6H10O5
XYLAN	Solid	Glutaric-Acid	C5H8O4
ARABINAN	Solid	Glutaric-Acid	C5H8O4
ACETATE	Solid	Acetic-Acid	C2H4O2-1
ASH	Solid	Calcium-Oxide	CAO
LIG-SOL	Conventional	Vanillin	C8H8O3-D1
LIGNIN	Solid	Vanillin	C8H8O3-D1
CELLULOS	Solid	Dilactic-Acid	C6H10O5
EXTRACT	Conventional	Dextrose	C6H12O6
H2SO4	Conventional	Sulphuric-Acid	H2SO4
H2O	Conventional	Water	H2O
O2	Conventional	Oxygen	O2
N2	Conventional	Nitrogen	N2
CO2	Conventional	Carbon-Dioxide	CO2
H2	Conventional	Hydrogen	H2
GLUCOSE	Conventional	Dextrose	C6H12O6
XYLOSE	Conventional	Xylose	C5H10O5-D2
ARABINOS	Conventional	Arabinose	C5H10O5-D1
CELLOBIO	Conventional	Cellobiose	C12H22O11-D1
GLUCOLIG	Conventional	Dilactic-Acid	C6H10O5
XYLOLIG	Conventional	Glutaric-Acid	C5H8O4
ARABOLIG	Conventional	Glutaric-Acid	C5H8O4
FURFURAL	Conventional	Furfural	C5H4O2
HMF	Conventional	5-Hydroxymethylfurfural	C6H6O3-N5
AACID	Conventional	Acetic-Acid	C2H4O2-1
NH3	Conventional	Ammonia	H3N
AMM-SULF	Solid	Ammonium-Sulfate	(NH4)2SO4
LIME	Solid	Calcium-Oxide	CAO
ETHANOL	Conventional	Ethanol	C2H6O-2
AIR	Conventional	Air	AIR
XYLITOL	Conventional	Xylitol	C5H12O5
ARABITOL	Conventional	D-Arabinitol	C5H12O5-N2
SORBITOL	Conventional	Sorbitol	C6H14O6



CASO4	Solid	Calcium-Sulphate	CASO4
CH4	Conventional	Methane	CH4
BIOMASS	Solid	Ethyl-Cyanoacetate	C5H7NO2
CACO3	Conventional	Calcium-Carbonate-Calcite	CACO3
HCO3-	Conventional	Hco3-	HCO3-
H+	Conventional	H+	H+
NH4+	Conventional	Nh4+	NH4+
Component ID	Type	Component name	Alias
NO3-	Conventional	No3-	NO3-
OH-	Conventional	Oh-	OH-
CA-ANHYD	Solid	Citric-Acid	C6H8O7
CITRICA	Solid	Citric-Acid-Monohydrate	C6H10O8
CA(OH)2	Solid	Calcium-Hydroxide	CA(OH)2
CELLULAS	Solid		CHXNXOXSX
PRO-ACID	Conventional	Propionic-Acid	C3H6O2-1
BUT-ACID	Conventional	N-Butyric-Acid	C4H8O2-1
CAYEAST	Solid		CHXOXNXPX
NH4NO3	Solid	Ammonium-Nitrate	NH4NO3
KH2PO4	Solid	Potassium-Dihydrogen-Phosphate	KH2PO4
K+	Conventional	K <sup>+</sup>	K+
OXALI-01	Conventional	Oxalic-Acid	C2H2O4
GLUCO-01	Conventional	Gluconic-Acid	C6H12O7
01-Oct-01	Conventional	1-Octanol	C8H18O-1
CAC2O-01	Solid	CaC <sub>2</sub> O <sub>4</sub>	CAC2O4
NAOH	Solid	Sodium-Hydroxide	NAOH
NA-CITRA	Solid	C <sub>6</sub> H <sub>5</sub> Na <sub>3</sub> O <sub>7</sub>	C6H5NA3O7
GLUTACID	Conventional	L-Glutamic-Acid	C5H9NO4
CA-GLUT	Solid	Calcium-L-Glutamate	C5H9NO4-N3
K2HPO4	Solid	Dipotassium-Phosphate	K2HPO4
MGSO4	Solid	Magnesium-Sulphate	MGSO4
UREA	Conventional	Urea	CH4N2O
POTAS-01	Solid	Potassium-Dihydrogen-Phosphate	KH2PO4
K2SO4	Solid	Potassium-Sulphate	K2SO4
FESO4	Solid	Ferrous-Sulphate	FESO4
MNSO4	Solid	Manganese-Sulphate	MNSO4
OLEIC-01	Conventional	Oleic-Acid	C18H34O2
BREVIBAC	Solid		CXHXNXOX

## C2. Glutamic Acid Aspen Simulation Process Flow Diagrams

### C2.1. Overall GA Biorefinery - Hybrid Membrane Reactor System

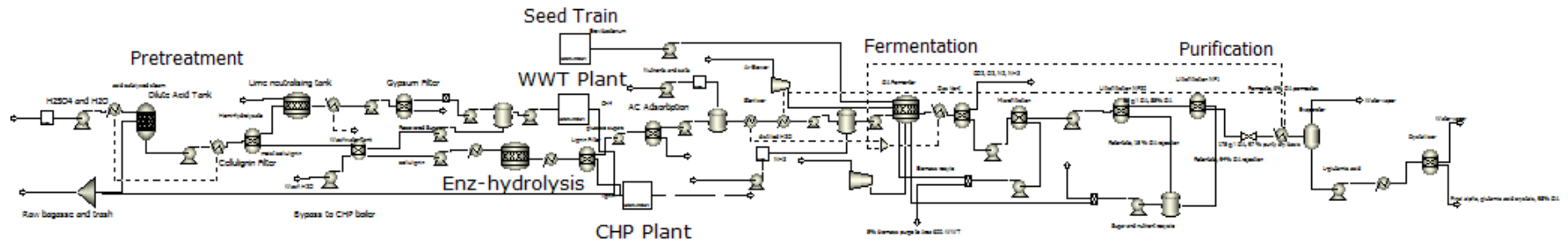


Figure 72 - Aspen GA Biorefinery Flowsheet

### C2.2. Area 100 Pretreatment

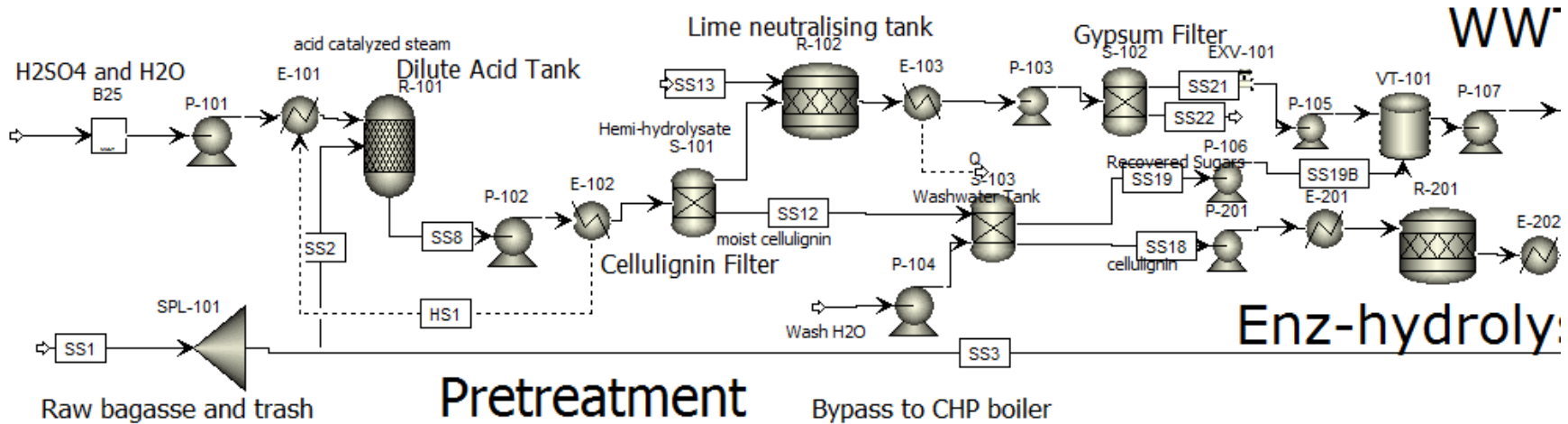


Figure 73 – Aspen GA Pretreatment Flowsheet

### C2.3. Area 200 Enzymatic Hydrolysis

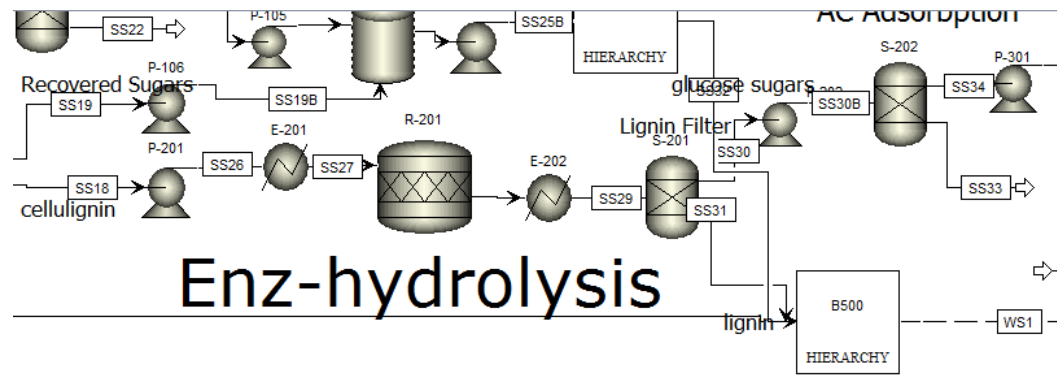


Figure 74 – Aspen GA Enzymatic Hydrolysis Flowsheet (Area 200)

### C2.4. Area 300 Fermentation

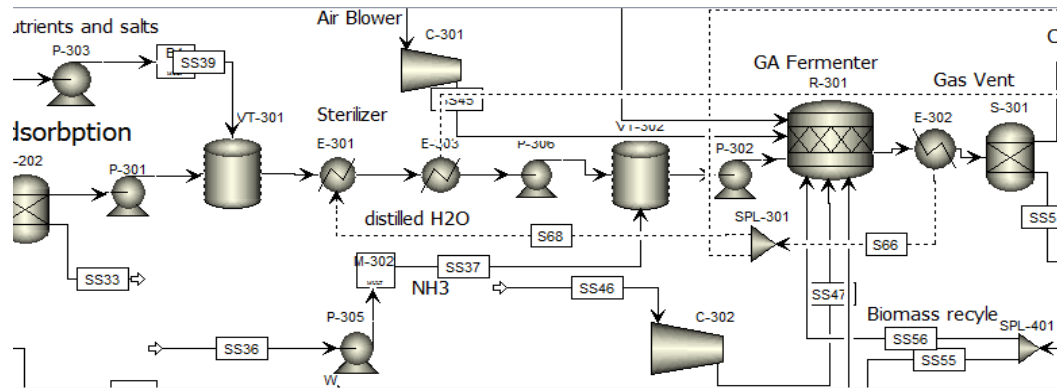


Figure 75 - Aspen GA Fermentation Flowsheet (Area 300)

### C2.5. Area 400 Purification

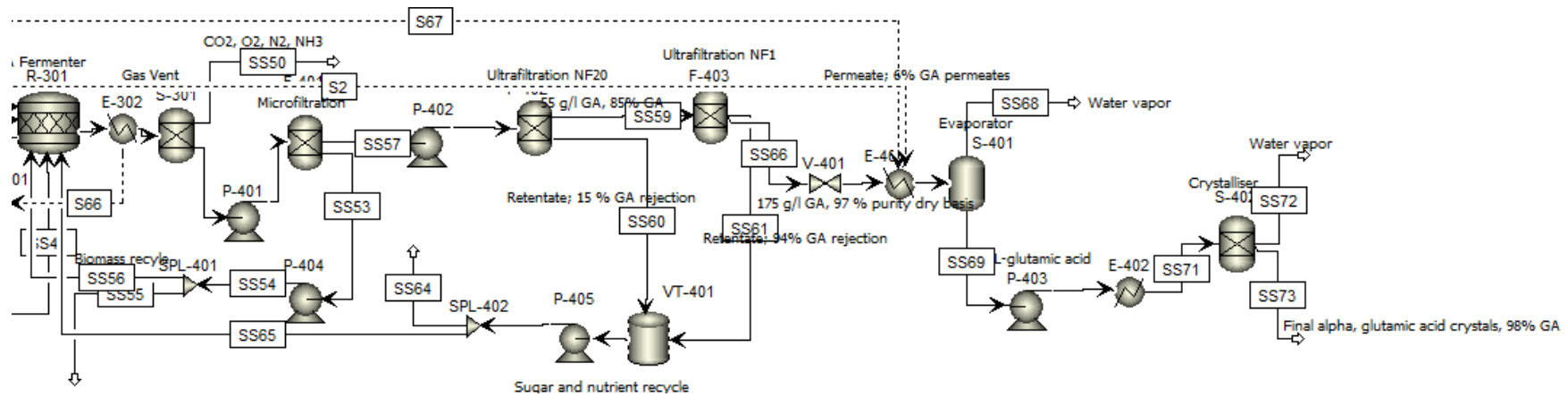


Figure 76 – Aspen GA Purification Flowsheet (Area 400)

### C2.6. Area 500 CHP Plant

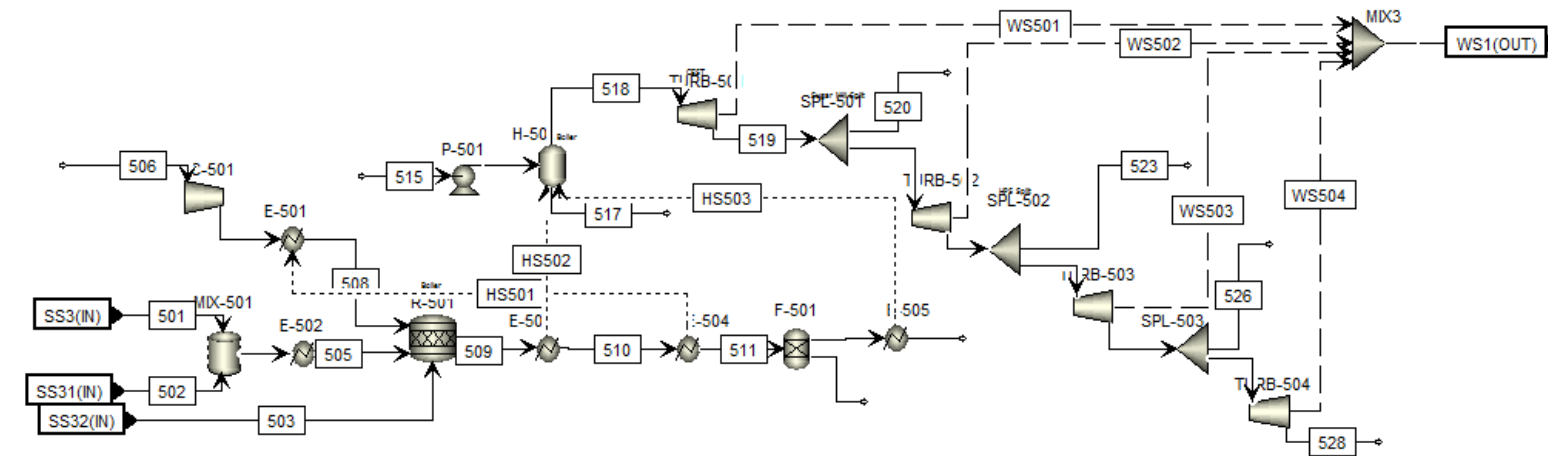


Figure 77 – Aspen GA CHP Plant Flowsheet (Area 500)

C2.7. Area 600 WWTW

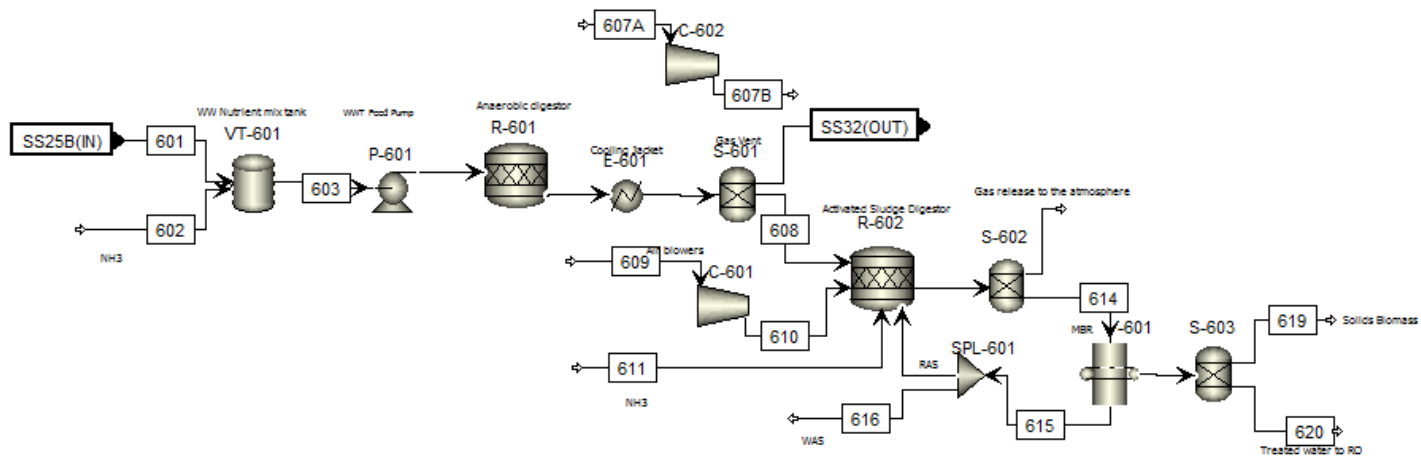


Figure 78 – Aspen GA WWT Plant Flowsheet (Area 600)

C2.8. Area 700 Brevibacterium Seed Train

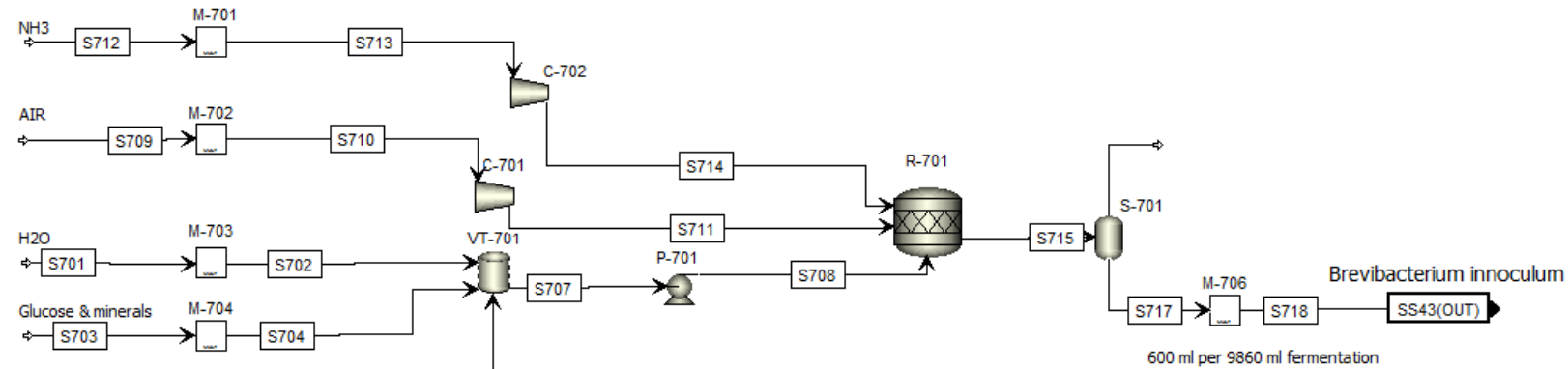


Figure 79 – Aspen GA Brevibacterium Seed Train Flowsheet (Area 700)

### C3. GA Mass and Energy Balance Streams

Streams	S1	SS1	SS2	SS3	SS4	SS5	SS6	SS7	SS8	SS9	SS10	SS11	SS12	SS13	SS14	SS15	SS16	SS17	SS18	
Temperature C	70	25	25	25	40	40	40	117.6	130.6	130.6	60	60	60	25	78.4	60	25	25	50.5	
Pressure atm	0.3	1	1	1	1	1	1.8	1.8	2.714	2.714	3	3	2	2	1	1	1	1	1.5	0.5
Vapor Frac	0.295	0	0	0	0	0	0	0.026	0	0	0	0	0	0	0	0	0	0	0	0
Mole Flow kmol/hr	4342.518	2862.32	1860.508	1001.812	15992.76	14521.42	14521.42	14521.42	16323.87	16323.87	16323.87	14699.46	1624.409	83	14782.46	14782.46	755.346	755.346	1717.944	
Mass Flow tons/hr	97.088	113.529	73.794	39.735	325.292	295.365	295.365	295.365	369.159	369.159	369.159	307.837	61.322	5.131	312.968	312.968	15	15	62.037	
Volume Flow cum/hr	119958.7	89.501	58.176	31.325	293.48	266.48	266.471	6985.975	340.31	340.306	323.99	275.067	49.001	1.411	278.878	275.911	13.646	13.646	50.175	
Enthalpy Gcal/hr	-292.166	-247.59	-160.933	-86.656	-1098.07	-997.047	-997.04	-973.051	-1128.17	-1128.17	-1149.84	-1014.04	-135.807	-12.59	-1026.63	-1031.6	-51.566	-51.566	-141.077	
Mass Flow tons/hr																				
GLUCAN	0	2.498	1.623	0.874	0	0	0	0	1.463	1.463	1.463	0	1.463	0	0	0	0	0	1.463	
XYLAN	0	11.58	7.527	4.053	0	0	0	0	0.753	0.753	0.753	0	0.753	0	0	0	0	0	0.753	
ARABINAN	0	2.157	1.402	0.755	0	0	0	0	0.14	0.14	0.14	0	0.14	0	0	0	0	0	0.14	
ACETATE	0	1.362	0.886	0.477	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
ASH	0	2.271	1.476	0.795	0	0	0	0	1.476	1.476	1.476	0	1.476	0	0	0	0	0	1.476	
LIG-SOL	0	0	0	0	0	0	0	0	0.139	0.139	0.139	0.127	0.012	0	0.127	0.127	0	0	0.004	
LIGNIN	0	14.305	9.298	5.007	0	0	0	0	9.159	9.159	9.159	0	9.159	0	0	0	0	0	9.159	
CELLULOS	0	26.452	17.194	9.258	0	0	0	0	15.445	15.445	15.445	0	15.445	0	0	0	0	0	15.445	
EXTRACT	0	4.428	2.878	1.55	0	0	0	0	2.878	2.878	2.878	0	2.878	0	0	0	0	0	2.878	
H2SO4	0	0	0	0	9.433	8.566	8.566	8.566	8.566	8.566	8.566	7.806	0.76	0	0	0	0	0	0.228	
H2O	84.721	48.477	31.51	16.967	315.859	286.8	286.8	286.8	317.159	317.159	317.159	289.03	28.129	0	290.464	290.464	15	15	30.19	
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
GLUCOSE	0	0	0	0	0	0	0	0	1.977	1.977	1.977	1.801	0.175	0	1.801	1.801	0	0	0.053	
XYLOSE	0	0	0	0	0	0	0	0	7.346	7.346	7.346	6.694	0.651	0	6.694	6.694	0	0	0.13	
ARABINOS	0	0	0	0	0	0	0	0	1.434	1.434	1.434	1.307	0.127	0	1.307	1.307	0	0	0.038	
CELLOBIO	0	0	0	0	0	0	0	0	0.049	0.049	0.049	0	0.049	0	0	0	0	0	0.049	
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
FURFURAL	0	0	0	0	0	0	0	0	0.226	0.226	0.226	0.206	0.02	0	0.206	0.206	0	0	0.006	
HMF	0	0	0	0	0	0	0	0	0.065	0.065	0.065	0.06	0.006	0	0.06	0.06	0	0	0.002	
AACID	0	0	0	0	0	0	0	0	0.886	0.886	0.886	0.807	0.079	0	0.807	0.807	0	0	0.024	
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
LIME	0	0	0	0	0	0	0	0	0	0	0	0	0	5.131	0.668	0.668	0	0	0	
CASO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	10.835	10.835	0	0	0	
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
NH4NO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
GLUTACID	12.367	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
CA-GLUT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
K2HPO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	
MGSO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	

Streams	SS19	SS19B	SS20	SS21	SS22	SS23	SS24	SS25	SS25B	SS26	SS27	SS28	SS29	SS30	SS30B	SS31	SS32	SS33	SS34	SS35	SS36	SS36B	SS37
Temperature C	50.5	50.5	60	60	60	60.1	60.1	59.7	59.7	50.5	48	52.5	48	48	48	48	40	48	48	48	25	25	25
Pressure atm	0.5	2	3	3	3	1.5	2.5	2	2.5	2	1	1	1	1	1.5	1	2	1	1	1.5	1	1.2	1.2
Vapor Frac	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1	0	0	0	0	0	0
Mole Flow kmol/hr	661.811	661.811	14782.46	14480.06	302.401	14480.06	14480.06	15141.87	15141.87	1717.944	1717.944	1632.7	1632.7	1234.663	1234.663	398.037	143.955	229.614	1005.049	1005.049	4854.357	4854.357	4854.357
Mass Flow tons/hr	14.285	14.285	312.968	297.108	15.86	297.108	297.108	311.393	311.393	62.037	62.037	62.037	62.037	39.981	39.981	22.056	3.238	4.563	35.418	35.418	96.4	96.4	96.4
Volume Flow cum/hr	12.478	12.477	275.891	268.367	7.524	268.389	268.379	280.862	280.857	50.174	50.131	37.503	37.439	21.801	21.8	15.644	1842.418	4.184	17.617	17.617	87.701	87.7	87.7
Enthalpy Gcal/hr	-46.297	-46.296	-1031.59	-990.443	-41.143	-990.443	-990.434	-1036.73	-1036.73	-141.075	-141.168	-141.168	-141.352	-104.017	-104.017	-37.335	-4.255	-15.581	-88.435	-88.435	-331.4	-331.399	-331.399
Mass Flow tons/hr																							
GLUCAN	0	0	0	0	0	0	0	0	0	1.463	1.463	0.146	0.146	0	0	0.146	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0.753	0.753	0.753	0.753	0	0	0.753	0	0	0	0	0	0	0
ARABINAN	0	0	0	0	0	0	0	0	0	0.14	0.14	0.14	0.14	0	0	0.14	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	1.476	1.476	1.476	1.476	0	0	1.476	0	0	0	0	0	0	0
LIG-SOL	0.009	0.009	0.127	0.127	0	0.127	0.127	0.136	0.136	0.004	0.004	0.004	0.004	0.004	0.004	0	0	0.004	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	9.159	9.159	9.159	9.159	0	0	9.159	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	15.445	15.445	1.526	1.526	0	0	1.526	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	2.878	2.878	2.878	2.878	0	0	2.878	0	0	0	0	0	0	0
H2SO4	0.532	0.532	0	0	0	0	0	0.532	0.532	0.228	0.228	0.228	0.228	0	0	0.228	0	0	0	0	0	0	0
H2O	12.939	12.939	290.464	286.107	4.357	286.107	286.107	299.046	299.046	30.19	30.19	28.496	28.496	22.797	22.797	5.699	0	4.559	18.237	18.237	96.4	96.4	96.4
O2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
N2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	1.089	0	0	0	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0.123	0.123	1.801	1.801	0	1.801	1.801	1.924	1.924	0.053	0.053	17.012	17.012	17.012	17.012	0	0	0	17.012	17.012	0	0	0
XYLOSE	0.521	0.521	6.694	6.694	0	6.694	6.694	7.215	7.215	0.13	0.13	0.13	0.13	0.13	0.13	0	0	0	0.13	0.13	0	0	0
ARABINOS	0.089	0.089	1.307	1.307	0	1.307	1.307	1.396	1.396	0.038	0.038	0.038	0.038	0.038	0.038	0	0	0	0.038	0.038	0	0	0
CELLOBIO	0	0	0	0	0	0	0	0	0	0.049	0.049	0.02	0.02	0	0	0.02	0	0	0	0	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0.014	0.014	0.206	0.206	0	0.206	0.206	0.22	0.22	0.006	0.006	0.006	0.006	0	0	0.006	0	0	0	0	0	0	0
HMF	0.004	0.004	0.06	0.06	0	0.06	0.06	0.064	0.064	0.002	0.002	0.002	0.002	0	0	0.002	0	0	0	0	0	0	0
AACID	0.055	0.055	0.807	0.807	0	0.807	0.807	0.862	0.862	0.024	0.024	0.024	0.024	0	0	0.024	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0.668	0	0.668	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CASO4	0	0	10.835	0	10.835	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2.149	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH4NO3	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
KH2PO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUTACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CA-GLUT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
K2HPO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MGSO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
POTAS-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
K2SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FESO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MNSO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
OLEIC-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0

Streams	SS38	SS38B	SS39	SS40	SS40A	SS40B	SS41	SS41A	SS42	SS43	SS43B	SS44	SS45	SS46	SS47	SS48	SS49	SS50	SS51	SS52	SS53
Temperature C	20	20	20	47.6	50	121	30.9	50	30.9	35	35	25	26.4	25	26.2	64.3	32	32	32	32	32
Pressure atm	1	1.135	1.135	1.135	1.135	1.135	1.2	1.2	1.5	1	1.133	1	1.014	1	1.014	1	1	1	1	2.467	2.467
Vapor Frac	0	0	0	0	0	0.79	0	0	0	0	0	1	1	1	1	0.29	0.229	1	0	0	0
Mole Flow kmol/hr	7.955	7.955	14.884	1019.933	1019.933	1019.933	5874.291	1019.933	5874.291	259.139	259.139	3144.445	3144.445	89.224	89.224	14141.74	14141.74	3165.479	10976.26	10976.26	15.861
Mass Flow tons/hr	1	1	1.871	37.289	37.289	37.289	133.689	37.289	133.689	5.242	5.242	100	100	1.675	1.675	341.821	341.821	100.846	240.975	240.975	1.978
Volume Flow cum/hr	0.347	0.347	0.649	18.252	18.274	22791.8	105.974	18.274	105.973	5.328	5.328	76893.03	76171.18	2166.985	2144.752	113554.8	81179.27	79218.46	205.397	205.385	1.501
Enthalpy Gcal/hr	-1.326	-1.326	-2.482	-90.917	-90.85	-81.2	-422.25	-90.85	-422.249	-15.155	-15.155	-0.005	0.026	-0.979	-0.978	-762.197	-778.24	-8.513	-770.682	-770.675	-0.364
Mass Flow tons/hr																					
GLUCAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABINAN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ACETATE	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ASH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIG-SOL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIGNIN	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CELLULOS	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
EXTRACT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
H2O	0	0	0	18.237	18.237	18.237	114.637	18.237	114.637	4.025	4.025	0	0	0	0	214.485	214.485	0	214.485	214.485	0
O2	0	0	0	0	0	0	0	0	0	0.003	0.003	23.292	23.292	0	0	18.852	18.852	18.852	0	0	0
N2	0	0	0	0	0	0	0	0	0	0.003	0.003	76.708	76.708	0	0	76.712	76.712	76.712	0	0	0
CO2	0	0	0	0	0	0	0	0	0	0.001	0.001	0	0	0	0	4.122	4.122	4.122	0	0	0
H2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOSE	0	0	0	17.012	17.012	17.012	17.012	17.012	17.012	0.157	0.157	0	0	0	0	2.215	2.215	0	2.215	2.215	0
XYLOSE	0	0	0	0.13	0.13	0.13	0.13	0.13	0.13	0	0	0	0	0	0	0.434	0.434	0	0.434	0.434	0
ARABINOS	0	0	0	0.038	0.038	0.038	0.038	0.038	0.038	0	0	0	0	0	0	0.127	0.127	0	0.127	0.127	0
CELLOBIO	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
GLUCOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
XYLOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
ARABOLIG	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
FURFURAL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
HMF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
AACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NH3	0	0	0	0	0	0	0	0	0	1.039	1.039	0	0	1.675	1.675	1.161	1.161	1.161	0	0	0
AMM-SULF	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
LIME	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CASO4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CH4	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BIOMASS	0	0	0	0	0	0	0	0	0	0.007	0.007	0	0	0	0	1.978	1.978	0	1.978	1.978	1.978
NH4NO3	0.361	0.361	0.675	0.675	0.675	0.675	0.675	0.675	0.675	0	0	0	0	0	0	2.25	2.25	0	2.25	2.25	0
KH2PO4	0.087	0.087	0.162	0.162	0.162	0.162	0.162	0.162	0.162	0	0	0	0	0	0	0.54	0.54	0	0.54	0.54	0
GLUTACID	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	15.478	15.478	0	15.478	15.478	0
CA-GLUT	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
K2HPO4	0	0	0	0	0	0	0	0	0	0.004	0.004	0	0	0	0	0.015	0.015	0	0.015	0.015	0
MGSO4	0.043	0.043	0.081	0.081	0.081	0.081	0.081	0.081	0.081	0.002	0.002	0	0	0	0	0.277	0.277	0	0.277	0.277	0
UREA	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
POTAS-01	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
K2SO4	0.087	0.087	0.162	0.162	0.162	0.162	0.162	0.162	0.162	0	0	0	0	0	0	0.54	0.54	0	0.54	0.54	0
FESO4	0.208	0.208	0.389	0.389	0.389	0.389	0.389	0.389	0.389	0	0	0	0	0	0	1.296	1.296	0	1.296	1.296	0
MNSO4	0.208	0.208	0.389	0.389	0.389	0.389	0.389	0.389	0.389	0	0	0	0	0	0	1.296	1.296	0	1.296	1.296	0
OLEIC-01	0.007	0.007	0.013	0.013	0.013	0.013	0.013	0.013	0.013	0	0	0	0	0	0	0.045	0.045	0	0.045	0.045	0





#### C4. GA Utility Consumption Summary

Table 81 - GA Utility Summary

Area Name	Area	Utility	Equipment ID	Description	Classification	Quantity (tons/h or kW)
Purification	400	LPS	S-401	Evaporation Column	Separator	61.36
Pretreatment	100	LPS	R-101	Dilute Acid Tank	Reactor	12.27
Enzymatic Hydrolysis	200	Cooling Water	E-201	Cellulignin Slurry Cooler	Exchanger	10.31
Enzymatic Hydrolysis	200	Cooling Water	E-202	Enzymatic Cooling Jacket	Exchanger	20.43
Purification	400	Cooling Water	E-402	Pre-crystallisation Cooler	Exchanger	45.51
Brevibacterium Seed Train	700	Chilled Water	R-701	Seed Fermenter Train	Reactor	98.92
CHP Plant	500	Electricity	P-501	Boiler Water Feed Pump	Pump	452.33
CHP Plant	500	Electricity	C-501	Boiler Air Blowers	Compressor	329.63
WWT	600	Electricity	P-601	Wastewater Feed Pump	Pump	8.33
WWT	600	Electricity	C-601	Aerobic Digester Air Blowers	Compressor	501.59
WWT	600	Electricity	C-602	Natural Gas Compressor	Compressor	22.81
Brevibacterium Seed Train	700	Electricity	P-701	Seed Train Feed Pump	Pump	0.14
Brevibacterium Seed Train	700	Electricity	C-701	Seed Train Air Blower	Compressor	28.90
Brevibacterium Seed Train	700	Electricity	C-702	Seed Train NH3 Compressor	Compressor	48.38
Pretreatment	100	Electricity	P-101	Dilute Acid Feed Pump	Pump	10.67
Pretreatment	100	Electricity	P-102	Dilute Acid Hydrolysate Pump	Pump	4.87
Pretreatment	100	Electricity	P-103	Gypsum Filter Pump	Pump	27.61
Pretreatment	100	Electricity	P-104	Wash Water Feed Pump	Pump	0.34
Pretreatment	100	Electricity	P-105	Neutralised Hydrolysate Pump	Pump	13.43
Pretreatment	100	Electricity	P-106	Recovered Sugar Pump	Pump	0.58
Pretreatment	100	Electricity	P-107	WWT Transfer Pump	Pump	7.03
Enzymatic Hydrolysis	200	Electricity	P-201	Enzymatic Tank Feed Pump	Pump	2.23
Enzymatic Hydrolysis	200	Electricity	P-202	Activated Carbon Column Feed Pump	Pump	0.55
GA Fermentation	300	Electricity	P-301	Detoxified Glucose Rich Feed Pump	Pump	0.44
GA Fermentation	300	Electricity	P-302	Sterilizer Feed Pump	Pump	0.94
GA Fermentation	300	Electricity	P-303	Nutrient Pump	Pump	0.00

GA Fermentation	300	Electricity	P-305	Makeup Water Pump	Pump	0.88
GA Fermentation	300	Electricity	P-306	Fermenter Feed Medium Holding Tank 2	Pump	0.06
Purification	400	Electricity	P-401	Micro filter Feed Pump	Pump	8.93
Purification	400	Electricity	P-402	Nanofiltration Feed Pump 1	Pump	104.33
Purification	400	Electricity	P-403	GA Crystalliser Feed Pump	Pump	0.16
Purification	400	Electricity	P-404	Biomass Recycle Pump	Pump	0.05
Purification	400	Electricity	P-405	Sugar Recycle Pump	Pump	57.33
GA Fermentation	300	Electricity	C-301	Fermenter Air Blower	Compressor	36.23
GA Fermentation	300	Electricity	C-302	Fermenter NH3 Compressor	Compressor	1.02
Pretreatment	100	Electricity	A-101	Sulphuric Acid Mixer	Agitator	9.33
Pretreatment	100	Electricity	A-102	Dilute Acid Tank Agitator	Agitator	11.36
Pretreatment	100	Electricity	A-103	Wash water Tank Agitator	Agitator	4.39
Pretreatment	100	Electricity	A-104	Lime Tank Agitator	Agitator	5.02
Pretreatment	100	Electricity	A-105	Recovered Sugar Agitator	Agitator	78.64
Enzymatic Hydrolysis	200	Electricity	A-201	Enzymatic Hydrolysis Agitator	Agitator	238.62
CA Fermentation	300	Electricity	A-301	Nutrient Tank Agitator	Agitator	36.41
CA Fermentation	300	Electricity	A-302	Fermenter Feed Medium Agitator	Agitator	36.41
WWT	600	Electricity	A-602	Aerobic Agitators	Agitator	939.47
Seed Train	700	Electricity	A-701	Seed Nutrient Tank Mixer	Agitator	15.88
Seed Train	700	Electricity	A-702	Seed Fermenter Train Agitator	Agitator	27.78
Pretreatment	100	Electricity	CONV-101	Bagasse and Trash Conveyor	Conveyor	0.37
Pretreatment	100	Electricity	CONV-102	Bypassed Biomass Conveyor	Conveyor	0.26
Pretreatment	100	Electricity	CONV-103	Lime Conveyor	Conveyor	0.01
Pretreatment	100	Electricity	CONV-104	Cellulignin Conveyor	Conveyor	0.12
Pretreatment	100	Electricity	CONV-105	Gypsum Conveyor	Conveyor	0.06
Purification	400	Electricity	CONV-401	GA Drying Conveyor	Conveyor	
CHP Plant	500	Electricity	CONV-501	Combined Boiler Feed Conveyor	Conveyor	0.42
WWT	600	Electricity	CONV-601	Solids Belt press	Conveyor	4.24

## C5. GA Equipment Sizing Summaries

## C5.1. Pumps

Area	Equipment ID	Type	Description	Quantity	Units of capacity	Duty (kW)	Usage (kW or tons/h)	Inlet Flow (m3/h)	Inlet Flow (L/s)
100	P-101	Pump	Dilute Acid Feed Pump	2	L/s	10.67	10.67	266.48	74.02
100	P-102	Pump	Dilute Acid Hydrolysate Pump	2	L/s	4.87	4.87	340.31	94.53
100	P-103	Pump	Gypsum Filter Pump	2	L/s	27.62	27.62	13.65	3.79
100	P-104	Pump	Wash Water Feed Pump	2	L/s	0.34	0.34	13.65	3.79
100	P-105	Pump	Neutralised Hydrolysate Pump	2	L/s	0.00	0.00	268.38	74.55
100	P-106	Pump	Recovered Sugar Pump	2	L/s	0.58	0.58	12.48	3.47
100	P-107	Pump	WWT Transfer Pump	2	L/s	7.03	7.03	280.86	78.02
200	P-201	Pump	Enzymatic Tank Feed Pump	2	L/s	0.74	0.74	50.17	13.94
200	P-202	Pump	Activated Carbon Column Feed Pump	2	L/s	0.55	0.55	21.80	6.06
300	P-301	Pump	Detoxified Glucose Rich Feed Pump	2	L/s	0.44	0.44	17.62	4.89
300	P-302	Pump	Sterilizer Feed Pump	2	L/s	0.00	0	18.27399	5.08
300	P-303	Pump	Nutrient Pump	2	L/s	0.94	0.94	0.35	0.10
300	P-304	Pump	Fermenter Seed Feed Pump	2	L/s	0.00	0.00	5.33	1.48
300	P-305	Pump	Makeup Water Pump	2	L/s	0.00	0.00	87.70	24.36
300	P-306	Pump	Fermenter Feed Medium Holding Tank 2	2	L/s	0.88	0.88	18.27	5.08
300	P-307	Pump	Fermentation Medium Pump	2	L/s	0.06	0.06	18.25	5.07
400	P-401	Pump	Micro filter Feed Pump	2	L/s	8.93	8.93	223.19	62.00

Area	Equipment ID	Type	Description	Quantity	Units of capacity	Duty (kW)	Usage (kW or tons/h)	Inlet Flow (m <sup>3</sup> /h)	Inlet Flow (L/s)
400	P-402	Pump	Nanofiltration Feed Pump 1	2	L/s	104.33	104.33	203.81	56.61
400	P-403	Pump	GA Crystalliser Feed Pump	2	L/s	0.16	0.16	7.66	2.13
400	P-404	Pump	Biomass Recycle Pump	2	L/s	0.05	0.05	1.50	0.42
400	P-405	Pump	Sugar Recycle Pump	2	L/s	57.33	57.33	121.78	33.83
400	Pump before F-403	Pump	Nanofiltration Feed Pump 2	2	L/s	0.05	0.05	202.90	56.36
500	P-501	Pump	Boiler Water Feed Pump	2	L/s	452.33	452.33	191.07	53.08
600	P-601	Pump	Wastewater Feed Pump	2	L/s	8.33	8.33	281.07	78.07
700	P-701	Pump	Seed Train Feed Pump	2	L/s	0.14	0.14	14.18	3.94

### C5.2. Conveyors

Conveyors										
Streams	Area	Equipment ID	Name	Vol. flow (m <sup>3</sup> /h)	Mass flow (ton/h)	Number	Material	Residence Time (h)	Length (m)	P (kW) = m x kW/m
SS1	100	CONV-101	Bagasse and Trash Conveyor	89.50	113.5	2	304 SS	0.067	11.93	0.37
SS3	100	CONV-102	Bypassed Biomass Conveyor	31.33	39.7	2	304 SS	0.167	10.44	0.26
SS13	100	CONV-103	Lime Conveyor	1.41	5.1	2	304 SS	0.083	0.24	0.01
SS31	100	CONV-104	Lignin Conveyor	15.64	22.1	2	304 SS	0.167	5.21	0.12
SS22	100	CONV-105	Gypsum Conveyor	7.52	15.9	2	304 SS	0.167	2.51	0.06
504	500	CONV-501	Combined Boiler Feed Conveyor	46.98	61.8	2	304 SS	0.167	15.66	0.42
618	600	CONV-601	Solids Belt press	275.65	302.6	2	304 SS	0.167	91.88	4.24

*C5.3. Compressors, Blowers and Turbines*

Area	Equipment ID	Type	Description	Classification	Quantity	Units of capacity	Duty (kW)	Usage (kW or tons/h)	Inlet Flow (m3/h)
300	C-301	Compressor	Fermenter Air Blower	Blower	2	m3/h	36.23	36.23	76893
300	C-302	Compressor	Fermenter NH3 Compressor	Compressor	2	kW	1.02	1.02	-
500	C-501	Compressor	Boiler Air Blowers	Blower	2	m3/h	329.63	329.63	625114
600	C-601	Compressor	Aerobic Digester Air Blowers	Blower	2	m3/h	501.59	501.59	68103
600	C-602	Compressor	Natural Gas Compressor	Compressor	2	kW	22.81	22.81	
700	C-701	Compressor	Seed Train Air Blower	Compressor	2	m3/h	28.90	28.90	3924
700	C-702	Compressor	Seed Train NH3 Compressor	Compressor	2	kW	48.38	48.38	
300	TURB-501	Turbine	Condensing Steam Turbine (Stage 1)	Turbine	2	kW	-17285	-	10617
300	TURB-502	Turbine	HP Steam Turbine (Stage 2)	Turbine	0	kW			
500	TURB-503	Turbine	LP Steam Turbine (Stage 3)	Turbine	0	kW			
600	TURB-504	Turbine	Steam Turbine (Stage 4)	Turbine	0	kW			

*C5.4. Heat Exchangers*

Equipment	Description	Duty Q (kW)	U (W/m2K)	T1 (°C) hot fluid, in	T2 (°C) hot fluid, out	t1 (°C) cold fluid, in	t2 (°C) cold fluid, out	Tlm (°C)	A (m2) = Q/(U x Ft x ΔTlm)
E-101		27899	680	131	60	40	118	16.25	2524.88
E-102	Heated hydrolysate is being cooled with water	-25209	680	131	60	20	30	65.86	562.92
E-103									
E-201	Cellulignin slurry is cooled with water	-108.00	450	50	48	20	30	23.78	10.09
E-202	Cellulignin slurry is	214.36	450	52	48	20	30	24.88	19.15

cooled with water									
Equipment	Description	Duty Q (kW)	U (W/m <sup>2</sup> K)	T1 (°C) hot fluid, in	T2 (°C) hot fluid, out	t1 (°C) cold fluid, in	t2 (°C) cold fluid, out	Tlm (°C)	A (m <sup>2</sup> ) = Q/(U x Ft x ΔTlm)
E-301	Fermenter nutrient sterilizer	11301	2750	233	170	48	121	116.93	35.14
E-302		-18658							
E-402	GA viscous solution cooled with water	-478	450	80	30	20	30	24.85	42.7
E-501	Combustion air preheater	29255	7.9	270	149	27	166	112.76	32841.4
E-503	Heat travels from combustion chamber to water boiler	-157935	680	871	270	20	30	487.17	476.7
E-504	Heat travels from combustion chamber back to inlet feed air	-29255	680	270	149	20	30	178.79	240.6
E-505	Heat travels from combustion chamber to water boiler	-22217	680	149	55	20	30	68.64	476.0

## C5.5. Separators and Filters

Separation Equipment						
Equipment	Equipment Name	tons/h	m <sup>3</sup> /h	L/h	# of presses	Material
S-101	Cellulignin Filter Press	369.16	324	323991	20	304 Stainless Steel
S-102	Gypsum Filter Press	312.97	276	275891	17	Carbon Steel
S-201	Lignin Solids Filter	62.04	37	37439	2	304 Stainless Steel
S-402	Vacuum Crystalliser	15.08	7.4	7369	N/A	304 Stainless Steel
F-501	Particulate Bag Filter (Vent Scrubber)	1.16	N/A	N/A	N/A	304 Stainless Steel
F-601	Membrane Bioreactor Filter Modules	1017.08	926	926133	N/A	
S-603	Belt press	311.48	281	281075	11	304 Stainless Steel

Membranes											
Equipment	Description	Membrane Material	Membrane Module Type	Surface Area per membrane module (m <sup>2</sup> )	Membrane Flux (L/m <sup>2</sup> .h)	Outlet Conc. (g GA/l)	Transmembrane Pressure (bar)	Operating Pressure (bar)	Cross Flow Velocity (m/s)	Output of GA per module per day (kg GA/day/module)	No. modules (n) = (required output /day)/(module output/day)
F-401	Microfiltration Module Train	PVDF	flat-sheet cross-flow	13	49	64.5	-	2.5	1.25	986.076	301
F-402	Nanofilter Module Train NF20	Polyamide	flat-sheet cross-flow	13	75	55	15	40		1287	231
F-403	Nanofilter Module Train NF1	Polyamide	flat-sheet cross-flow	13	75	175	15	40		4095	72
F-501	Particulate Bag Filter						-				



Cost Parameter	Value
$C2 (13m^2) = C1*(S2/S1)^n$	14202
C1 (cost per 0.15 m <sup>2</sup> modules)	400
S2 (m <sup>2</sup> )	13
S1 (m <sup>2</sup> )	0.15
n	0.8
cost per 0.15 module	400
Membrane material	Polyamide
Membrane Type	Flat sheet

Flash drums								
Equipment	Flow (m <sup>3</sup> /h)	Configuration	Material	Vessel diameter, D <sub>c</sub> (m)	Vessel length, L <sub>c</sub> (m)	wall thickness, t <sub>w</sub> (m)	Stainless Steel Density, ρ (kg/m <sup>3</sup> )	Shell mass (kg) = π x D <sub>c</sub> x L <sub>c</sub> x t <sub>w</sub> x ρ
S-401	7.66	horizontal	304 SS	3	1.08	0.0187	8000	1528.1

### C5.6. Vessels and Reactors

Vessels									
Area	Equipment ID	Type	Description	Classification	Quantity	Units of capacity	Inlet Flow (m <sup>3</sup> /h)	Residence Time (h)	Volume (m <sup>3</sup> )
100	VT-101	Vertical Tank	Recovered Sugar Holding Tank	Storage Tank	2	m <sup>3</sup>	280.86	4	1123
300	VT-301	Vertical Tank	Fermenter Feed Medium Holding Tank 1	Storage Tank	2	m <sup>3</sup>	18.25	28.5	520
300	VT-302	Vertical Tank	Fermenter Feed Medium Holding Tank 2	Storage Tank	2	m <sup>3</sup>	105.97	28.5	3020
400	VT-401	Vertical Tank	Sugar Recycle Holding Tank	Storage Tank	1	m <sup>3</sup>	121.74	24	2921.66
600	VT-601	Vertical Tank	Wastewater Feed Holding Tank	Storage Tank	2	m <sup>3</sup>	280.86	24	6740.57

700	VT-701	Vertical Tank	Seed Nutrient Mixing Tank	Vessel	1	m3	14.18	16	226.81
-----	--------	---------------	---------------------------	--------	---	----	-------	----	--------

<b>Reactors</b>									
<b>Area</b>	<b>Equipment ID</b>	<b>Type</b>	<b>Description</b>	<b>Classification</b>	<b>Quantity</b>	<b>Units of capacity</b>	<b>Inlet Flow (m3/h)</b>	<b>Residence Time (h)</b>	<b>Volume (m3)</b>
100	R-101	Reactor	Dilute Acid Tank	Reactor	2	m3	351.66	0.5	264
100	R-102	Reactor	Lime Neutralisation Tank	Reactor	1	m3	276.48	0.25	69
200	R-202	Reactor	Enzymatic Hydrolysis Tank (2-stage)	Reactor	1	m3	50.13	68	3409
300	R-301	Reactor	Glutamic Acid Fermenter	Reactor	8	m3	198	34.5	6831
500	R-501	Reactor	CHP Boiler (Combustion Chamber)	Reactor	1	kW			
600	R-601	Reactor	Anaerobic Digester	Reactor	1	m3	280.86	90	25277.14
600	R-602	Reactor	Activated Sludge Digester	Reactor	1	m3	930.09	48	44644.09
700	R-701	Reactor	Seed Fermenter Train	Reactor	2	m3	14.18	28	396.92

## C5.7. Agitators

Agitators											
Type	Streams	Area	Equipment ID	Name	Vol. flow (m <sup>3</sup> /h)	Mass flow (ton/h)	#	Material	Residence Time (h)	Volume (m <sup>3</sup> )	P (kW) = 0.04-0.1 kW/m <sup>3</sup>
Vessel	SS6	100	A-101	Sulphuric Acid Mixer	266.47	295.37	1	316SS	0.5	133	9.3
Reactor	SS6+SS2	100	A-102	Dilute Acid Tank Agitator	324.65	369.16	1	316SS	0.5	162	11.4
Vessel	SS12+SS16	100	A-103	Washwater Tank Agitator	62.65	76.32	3	316SS	1	63	4.4
Reactor	SS11+SS13	100	A-104	Lime Tank Agitator	286.61	329.68	1	316SS	0.25	72	5.0
Vessel	SS19B+SS24	100	A-105	Recovered Sugar Agitator	280.86	311.39	1	316SS	4	1123	78.6
Reactor	SS27	200	A-201	Enzymatic Hydrolysis Agitator	50.13	62.04	1	316SS	68	3409	238.6
Vessel	SS40	300	A-301	Nutrient Tank Agitators	18.25	37.29	2	316SS	28.5	520	36.4
Reactor	SS40	300	A-302	Fermenter Agitator	18.25	37.29	8	316SS	28.5	520	36.4
Reactor	608	600	A-602	Aerobic Agitators	279.61	308.24	4	316SS	48	13421	939.5
Vessel	S707	700	A-701	Seed Nutrient Tank Mixer	14.18	6.16	1	316SS	16	227	15.9
Reactor	S707	700	A-702	Seed Fermenter Train Agitator	14.18	6.16	2	316SS	28	397	27.8

### C5. GA Equipment Cost Summary

Table 82 - GA Equipment Sizing and Cost Summary

Plant Section	Area	Equipment ID	Equipment Type	Equipment Description	Quantity	Duty (kW)	Inlet Flow (L/s) w/	Estimated Purchased	Installed Cost in 2007	Installed Cost in 2016	Residence Volume (m <sup>3</sup> )
Fermentation	300	C-301	Compressor	Fermenter Air Blower	2	36.2		\$ 1 269 565	\$ 2 031 304	\$ 2 073 865	
Fermentation	300	C-302	Compressor	Fermenter NH3 Compressor	2	1.0		\$ 1 014 012	\$ 1 622 419	\$ 1 656 412	
CHP	500	C-501	Compressor	Boiler Air Blowers	2	329.6		\$ 2 069 312	\$ 3 310 899	\$ 3 380 270	
WWT	600	C-601	Compressor	Aerobic Digester Air Blowers	2	501.6		\$ 2 381 354	\$ 3 810 166	\$ 3 889 998	
WWT	600	C-602	Compressor	Natural Gas Compressor	2	22.8		\$ 1 199 365	\$ 1 918 985	\$ 1 959 192	
Seed Trains	700	C-701	Compressor	Seed Train Air Blower	2	28.9		\$ 81 087	\$ 129 739	\$ 132 457	
Seed Trains	700	C-702	Compressor	Seed Train NH3 Compressor	2	48.4		\$ 1 324 457	\$ 2 119 132	\$ 2 163 533	
Enzymatic Hydrolysis	200	CONV-201	Conveyer	Lignin Solid Sludge Conveyer	2						
Purification	400	F-401	Filter	Microfiltration Module Train	1			\$ 4 185 346	\$ 7 115 089	\$ 7 264 167	
Purification	400	F-402	Filter	Nanofilter Module Train NF20	1			\$ 3 206 736	\$ 5 451 452	\$ 5 565 672	
Purification	400	F-403	Filter	Nanofilter Module Train NF1	1			\$ 1 007 831	\$ 1 713 313	\$ 1 749 211	
CHP	500	F-501	Filter	Particulate Bag Filter	1			\$ 11 400	\$ 26 089	\$ 26 635	
WWT	600	F-601	Filter	Membrane Bioreactor Filter Modules	1			\$ 23 067 854	\$ 23 067 854	\$ 23 551 180	
Purification	400	Dryer after S-402	Flash Drum	GA Dryer	1			Included			
Seed Trains	700	S-701	Flash Drum	Gas Vent	1			Included			
Pretreatment	100	E-101	Heat Exchanger	Dilute Acid Tank Heating Jacket	1	25509.2		\$ 580 475	\$ 1 277 046	\$ 1 303 803	
Pretreatment	100	E-102	Heat Exchanger	Hydrolysate Cooler	1	-25209.3		\$ 115 895	\$ 254 970	\$ 260 312	
Pretreatment	100	E-103	Heat Exchanger	Lime Tank Cooling Jacket	1						
Enzymatic Hydrolysis	200	E-201	Heat Exchanger	Cellulignin Slurry Cooler	1	-108.0		\$ 2 969	\$ 5 343	\$ 5 455	
Enzymatic Hydrolysis	200	E-202	Heat Exchanger	Enzymatic Cooling Jacket	2	214.4		\$ 8 647	\$ 15 564	\$ 15 890	
Fermentation	300	E-301	Heat Exchanger	Feed Medium Sterilizer	1	11301.0		\$ 27 295	\$ 60 048	\$ 61 306	
Fermentation	300	E-302	Heat Exchanger	Fermenter Cooling Jacket	1	-18658.0		Included			
Purification	400	E-402	Heat Exchanger	Precrystallisation Cooler	1	-477.6		\$ 28 162	\$ 50 692	\$ 51 754	
CHP	500	E-501	Heat Exchanger	Air Preheater	1			\$ 12 115 034	\$ 21 807 062	\$ 22 263 972	
CHP	500	E-502	Heat Exchanger	Has no purpose				Has no purpose (was to be used as preheater)			
CHP	500	E-503	Heat Exchanger	Flue Gas Cooler 1	1	0.0		\$ 112 486	\$ 247 470	\$ 252 655	
CHP	500	E-504	Heat Exchanger	Flue Gas Cooler 2	1	0.0		\$ 72 281	\$ 159 019	\$ 162 351	
CHP	500	E-505	Heat Exchanger	Flue Gas Cooler 3	1			\$ 112 361	\$ 247 195	\$ 252 374	
WWT	600	E-601	Heat Exchanger	Anaerobic Digester Cooling Jacket	1			Included			
CHP	500	H-501	Heater	CHP Boiler (Water Heater)	2	210000.0		\$ 20 610 999	\$ 45 344 199	\$ 46 294 268	
Fermentation	300	S-301	N/A	Gas Vent							
Pretreatment	100	P-101	Pump	Dilute Acid Feed Pump	2	10.7	74.0	\$ 33 630	\$ 77 348	\$ 78 969	
Pretreatment	100	P-102	Pump	Dilute Acid Hydrolysate Pump	2	4.9	94.5	\$ 38 512	\$ 88 578	\$ 90 434	
Pretreatment	100	P-103	Pump	Gypsum Filter Pump	2	27.6	3.8	\$ 15 167	\$ 34 884	\$ 35 615	
Pretreatment	100	P-104	Pump	Wash Water Feed Pump	2	0.3	3.8	\$ 15 167	\$ 34 884	\$ 35 615	
Pretreatment	100	P-105	Pump	Neutralised Hydrolysate Pump	2	0.0	74.5	\$ 33 757	\$ 77 642	\$ 79 268	
Pretreatment	100	P-106	Pump	Recovered Sugar Pump	2	0.6	3.5	\$ 15 061	\$ 34 640	\$ 35 366	
Pretreatment	100	P-107	Pump	WWT Transfer Pump	2	7.0	78.0	\$ 34 590	\$ 79 558	\$ 81 225	
Enzymatic Hydrolysis	200	P-201	Pump	Enzymatic Tank Feed Pump	2	0.7	13.9	\$ 18 212	\$ 41 888	\$ 42 766	
Enzymatic Hydrolysis	200	P-202	Pump	Activated Carbon Column Feed Pump	2	0.5	6.1	\$ 15 884	\$ 36 533	\$ 37 298	
Fermentation	300	P-301	Pump	Detoxified Glucose Rich Feed Pump	2	0.4	4.9	\$ 15 520	\$ 35 696	\$ 36 444	
Fermentation	300	P-302	Pump	Sterilizer Feed Pump	2	0.0	5.1	\$ 15 578	\$ 35 829	\$ 36 580	
Fermentation	300	P-303	Pump	Nutrient Pump	2	0.9	0.1	\$ 13 850	\$ 31 855	\$ 32 523	
Fermentation	300	P-304	Pump	Fermenter Seed Feed Pump	2	0.0	1.5	\$ 14 386	\$ 33 089	\$ 33 782	

Plant Section	Area	Equipmen	Equipment Type	Equipment Description	Quantity	Duty (kW)	Inlet Flow (L/s) wher	Equipment Cost, Ce (\$)	Installed Cost in 2007 (\$)	Installed Cost in 2016 (\$)	Residence Time (h)	Volume (m3)
Fermentation	300	P-305	Pump	Makeup Water Pump	2	0.0	24.4	\$	21 093 \$	48 515 \$	49 531	
Fermentation	300	P-306	Pump	Fermenter Feed Medium Holding Tank 2	2	0.9	5.1	\$	15 578 \$	35 829 \$	36 580	
Fermentation	300	P-307	Pump	Fermentation Medium Pump	2	0.1	5.1	\$	15 576 \$	35 824 \$	36 575	16
Purification	400	P-401	Pump	Microfilter Feed Pump	2	8.9	62.0	\$	30 705 \$	70 622 \$	72 102	0
Purification	400	P-402	Pump	Nanofiltration Feed Pump 1	2	104.3	56.6	\$	29 379 \$	67 571 \$	68 987	
Purification	400	P-403	Pump	GA Crystalliser Feed Pump	2	0.2	2.1	\$	14 613 \$	33 610 \$	34 314	
Purification	400	P-404	Pump	Biomass Recycle Pump	2	0.1	0.4	\$	13 988 \$	32 171 \$	32 845	
Purification	400	P-405	Pump	Sugar Recycle Pump	2	57.3	33.8	\$	23 601 \$	54 282 \$	55 419	
Purification	400	Pump befo	Pump	Nanofiltration Feed Pump 2	2	0.1	56.4	\$	29 316	\$	-	
CHP	500	P-501	Pump	Boiler Water Feed Pump	2	452.3	53.1	\$	28 499 \$	65 549 \$	66 922	
WWT	600	P-601	Pump	Wastewater Feed Pump	2	8.3	78.1	\$	34 604 \$	79 590 \$	81 258	
Seed Trains	700	P-701	Pump	Seed Train Feed Pump	2	0.1	3.9	\$	15 215 \$	34 993 \$	35 727	
Pretreatment	100	R-101	Reactor	Dilute Acid Tank	2			\$	4 843 183 \$	7 264 775 \$	7 416 989	0.5
Pretreatment	100	R-102	Reactor	Lime Neutralisation Tank	1			\$	829 544 \$	1 244 315 \$	1 270 387	0.25
Enzymatic Hydrolysis	200	R-202	Reactor	Enzymatic Hydrolysis Tank (2-stage)	1			\$	18 760 642 \$	28 140 963 \$	28 730 584	68
Fermentation	300	R-301	Reactor	Glutamic Acid Fermenter	8			\$	19 650 061 \$	29 475 092 \$	30 092 665	34.5
CHP	500	R-501	Reactor	CHP Boiler (Combustion Chamber)	1			\$				
WWT	600	R-601	Reactor	Anaerobic Digestor	1			\$	2 036 819 \$	3 055 229 \$	3 119 243	90
WWT	600	R-602	Reactor	Activated Sludge Digestor	1			\$	2 947 945 \$	4 421 917 \$	4 514 567	48
Seed Trains	700	R-701	Reactor	Seed Fermenter Train	2			\$	3 358 336 \$	5 037 504 \$	5 143 051	28
Pretreatment	100	S-101	Separator	Cellulignin Filter Press	1			\$	155 537 \$	264 412 \$	269 952	
Pretreatment	100	S-102	Separator	Gypsum Filter Press	1			\$	169 210 \$	287 658 \$	293 685	
Pretreatment	100	S-103	Separator	2 Stage Washwater Tank	3			\$	76 076 \$	136 938 \$	139 807	1
Enzymatic Hydrolysis	200	S-201	Separator	Lignin Solids Filter	1			\$	37 340 \$	63 479 \$	64 809	
Purification	400	S-401	Separator	Evaporation Column	1			\$	42 170 \$	84 341 \$	86 108	
Purification	400	S-402	Separator	Vacuum Crystalliser	1			\$	195 896 \$	333 022 \$	340 000	
WWT	600	S-601	Separator	Biogas Vent	1			\$				
WWT	600	S-602	Separator	Release to atmosphere				\$				
WWT	600	S-603	Separator	Beltpress	50			\$	50 000 \$	83 256 \$	85 000	
Purification	400	SPL-401	Splitter	Biomass Purge				\$				
Purification	400	SPL-402	Splitter	Sugar Water Purge (Sent to seed train)				\$				
CHP	500	SPL-501	Splitter	Sugar Mill Steam Spiltter	1			\$				
CHP	500	SPL-502	Splitter	HP Steam Splitter	1			\$				
CHP	500	SPL-502	Splitter	LP Steam Splitter	1			\$				
WWT	600	SPL-601	Splitter	Sludge Recycle Holding Vessel	1			\$	211 901 \$	381 422 \$	389 414	2
CHP	500	TURB-501	Turbine	Condensing Steam Turbine (Stage 1)	2	-17284.7		\$	4 890 330 \$	8 802 593 \$	8 987 029	
CHP	500	TURB-502	Turbine	HP Steam Turbine (Stage 2)	0			Included				
CHP	500	TURB-503	Turbine	LP Steam Turbine (Stage 3)	0			Included				
CHP	500	TURB-504	Turbine	Steam Turbine (Stage 4)	0			Included				
Purification	400	V-401	Valve	Expansion Valve for Pressure Loss								

Plant Section	Area	Equipmen	Equipment Type	Equipment Description	Quantity	Duty (kW)	Inlet Flow (L/s) wher	Equipment Cost, Ce (\$)	Installed Cost in 2007 (\$)	Installed Cost in 2016 (\$)	Residence Time (h)	Volume (m3)
Pretreatment	100	VT-101	Vertical Tank	Recovered Sugar Holding Tank	2		78.0	\$	191 222 \$	344 200 \$	351 412 4	1123
Fermentation	300	VT-301	Vertical Tank	Fermenter Feed Medium Holding Tank 1	2		5.1	\$	111 553 \$	200 795 \$	205 002 28.5	520
Fermentation	300	VT-302	Vertical Tank	Fermenter Feed Medium Holding Tank 2	2		29.4	\$	382 095 \$	687 772 \$	702 182 28.5	3020
Purification	400	VT-401	Vertical Tank	Sugar Recycle Holding Tank	1		33.8	\$	373 319 \$	671 975 \$	686 054 24	2922
WWT	600	VT-601	Vertical Tank	Wastewater Feed Holding Tank	2		78.0	\$	670 227 \$	1 206 409 \$	1 231 686 24	6741
Seed Trains	700	VT-701	Vertical Tank	Seed Nutrient Mixing Tank	1		3.9	\$	62 397 \$	112 315 \$	114 669 16	227
Pretreatment	100	CONV-101	Conveyor	Bagasse and Trash Conveyor	2	0.4	24.9	\$	107 686 \$	183 065 \$	186 901 0.07	6
Pretreatment	100	CONV-102	Conveyor	Bypassed Biomass Conveyor	2	0.3	8.7	\$	104 225 \$	323 097 \$	329 867 0.17	5
Pretreatment	100	CONV-103	Conveyor	Lime Conveyor	2	0.0	0.4	\$	80 546 \$	161 092 \$	164 467 0.08	0
Pretreatment	100	CONV-104	Conveyor	Lignin Conveyor	2	0.1	4.3	\$	92 098 \$	138 147 \$	141 041 0.17	3
Pretreatment	100	CONV-105	Conveyor	Gypsum Conveyor	2	0.1	2.1	\$	85 818 \$	197 382 \$	201 518 0.17	1
Purification	400	CONV-401	Conveyor	GA Drying Conveyor	2			\$	\$	- \$	-	
CHP Plant	500	CONV-501	Conveyor	Combined Boiler Feed Conveyor	2	0.4	13.1	\$	116 333 \$	197 765 \$	201 909 0.17	8
WWT	600	CONV-601	Conveyor	Solids Beltpress	2	4.2	76.6	\$	293 172 \$	439 758 \$	448 972 0.17	46
Pretreatment	100	A-101	Agitator	Sulphuric Acid Mixer	1	9.3	74.0	\$	25 324 \$	37 986 \$	38 782	
Pretreatment	100	A-102	Agitator	Dilute Acid Tank Agitator	1	11.4	90.2					
Pretreatment	100	A-103	Agitator	Washwater Tank Agitator	3	4.4	17.4	\$	19 674 \$	29 512 \$	30 130	
Pretreatment	100	A-104	Agitator	Lime Tank Agitator	1	5.0	79.6					
Pretreatment	100	A-105	Agitator	Recovered Sugar Agitator	1	78.6	78.0	\$	111 841 \$	167 761 \$	171 276	
Enzymatic Hydrolysis	200	A-201	Agitator	Enzymatic Hydrolysis Agitator	1	238.6	13.9					
Fermentation	300	A-301	Agitator	Nutrient Tank Agitators	2	36.4	5.1	\$	58 148 \$	87 222 \$	89 049	
Fermentation	300	A-302	Agitator	Fermenter Agitator	8	36.4	5.1					
WWT	600	A-602	Agitator	Aerobic Agitators	4	939.5	77.7					
Seed Trains	700	A-701	Agitator	Seed Nutrient Tank Mixer	1	15.9	3.9	\$	33 048 \$	49 573 \$	50 611	
Seed Trains	700	A-702	Agitator	Seed Fermenter Train Agitator	2	27.8	3.9	\$	47 481 \$	71 222 \$	72 714	

## C6. GA Pricing from Literature and Industry

Table 83 - GA Pricing According to Literature

Year	L-Glutamic Acid Price (\$/ton)	World Demand (tons/year)	Reference
1995	\$1 000	800000	(Eyer, et al., 1995) (Heinzle, et al., 2006)
2002	-	900000	(Delaunay, et al., 2002)
2003	\$1 200	765000	(Hermann, 2003)
2004	-	1000000	(Stephan, et al., 2004)
2006	-	1500000	(Lakshmi & Mangala, 2011)
2007	-	1500000	(Amin & Al-Talhi, 2007)
2009	-	2000000	(Sano, 2009)
2013	\$2 500	2300000	(Binder, 2013)
2014	-	2900000	(Global Market Insights, Inc., 2016)
2016	\$4 300	-	(Pal, et al., 2016)
2017	\$6 500	-	(Qingdao Greaty International Trade Co., 2017)
2023	-	4000000	(Global Market Insights, Inc., 2016)

C7. GA Discount Cash Flow Sheet

DCFROR Worksheet			2016	2017	2018	2019	2020	2021
Year	-2	-1	0	1	2	3	4	5
Fixed Capital Investment	\$ 40 183 424.85	\$ 241 100 549	\$ 120 550 275					
Land	\$ -	\$ -	\$ -					
Working Capital			\$ 20 091 712.42					
Product Price (\$/t)			\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625
Succinic Acid Sales			\$ -	\$ 148 386 330	\$ 222 579 495	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660
Electricity price (\$/ kWh)			\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales			\$ -	\$ 2 958 847	\$ 4 438 271	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695
Total Annual Revenue			\$ -	\$ 151 345 177	\$ 227 017 766	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355
Annual Manufacturing Cost								
Feedstock Price (\$/ ton)								
Feedstock cost								
Other Variable Costs								
Fixed Operating Costs								
Total Product Cost			\$ -	\$ 54 552 955	\$ 81 829 432	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910
Annual Depreciation								
Plant Writedown			0.00%	20.00%	20.00%	20.00%	20.00%	20.00%
Depreciation Charge			\$ -	\$ 80 366 850	\$ 80 366 850	\$ 80 366 850	\$ 80 366 850	\$ 80 366 850
Remaining Value			\$ 401 834 248	\$ 321 467 399	\$ 241 100 549	\$ 160 733 699	\$ 80 366 850	\$ 0
								\$ 0
Net Revenue (R-COM-dk)	(\$40 183 425)	(\$241 100 549)	(\$140 641 987)	\$ 16 425 373	\$ 64 821 484	\$ 113 217 595	\$ 113 217 595	\$ 113 217 595
Losses Forward				\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income				\$ 16 425 373	\$ 64 821 484	\$ 113 217 595	\$ 113 217 595	\$ 113 217 595
Income Tax				\$ 4 599 104	\$ 18 150 016	\$ 31 700 927	\$ 31 700 927	\$ 31 700 927
Annual Cash Income			(\$421 925 961)	\$ 92 193 118	\$ 127 038 318	\$ 161 883 518	\$ 161 883 518	\$ 161 883 518
Discount Factor		1.0970	1.0000	0.9116	0.8310	0.7575	0.6905	0.6295
Annual Present Value (Discounted Cash Flow)			(\$421 925 961)	\$ 84 041 129	\$ 105 565 372	\$ 122 626 054	\$ 111 783 094	\$ 101 898 900
Cumulative Discounted Cash Flow			(\$421 925 961)	(\$337 884 832)	(\$232 319 461)	(\$109 693 407)	\$ 2 089 687	\$ 103 988 587
Total Capital Investment + Interest	\$ 264 487 302		\$ 140 641 987					
Net Present Worth			\$ 866 474 202					



DCFROR Worksheet	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033
Year	6	7	8	9	10	11	12	13	14	15	16	17
Fixed Capital Investment												
Land												
Working Capital												
Product Price (\$/t)	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625
Succinic Acid Sales	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660
Electricity price (\$/ kWh)	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695
Total Annual Revenue	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355
Annual Manufacturing Cost												
Feedstock Price (\$/ ton)												
Feedstock cost												
Other Variable Costs												
Fixed Operating Costs												
Total Product Cost	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910
Annual Depreciation												
Plant Writedown												
Depreciation Charge												
Remaining Value												
Net Revenue (R-COM-dk)	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445
Losses Forward	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445
Income Tax	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645
Annual Cash Income	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801
Discount Factor	0.5738	0.5231	0.4768	0.4347	0.3962	0.3612	0.3292	0.3001	0.2736	0.2494	0.2273	0.2072
Annual Present Value (Discounted Cash Flow)	\$ 79 976 646	\$ 72 904 874	\$ 66 458 408	\$ 60 581 958	\$ 55 225 121	\$ 50 341 952	\$ 45 890 567	\$ 41 832 787	\$ 38 133 807	\$ 34 761 903	\$ 31 688 152	\$ 28 886 191
Cumulative Discounted Cash Flow	\$183 965 233	\$256 870 107	\$323 328 515	\$383 910 473	\$439 135 595	\$489 477 547	\$535 368 114	\$577 200 900	\$615 334 708	\$650 096 610	\$681 784 763	\$710 670 954
Total Capital Investment + Interest												
Net Present Worth												

DCFROR Worksheet	2034	2035	2036	2037	2038	2039	2040	2041
Year	18	19	20	21	22	23	24	25
Fixed Capital Investment								
Land								
Working Capital								
Product Price (\$/t)	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625	\$ 3 625
Succinic Acid Sales	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660	\$ 296 772 660
Electricity price (\$/ kWh)	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080	\$ 0.080
Electricity Sales	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695	\$ 5 917 695
Total Annual Revenue	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355	\$ 302 690 355
Annual Manufacturing Cost								
Feedstock Price (\$/ ton)								
Feedstock cost								
Other Variable Costs								
Fixed Operating Costs								
Total Product Cost	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910	\$ 109 105 910
Annual Depreciation								
Plant Writedown								
Depreciation Charge								
Remaining Value								
Net Revenue (R-COM-dk)	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445
Losses Forward	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -	\$ -
Taxable Income	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445	\$ 193 584 445
Income Tax	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645	\$ 54 203 645
Annual Cash Income	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801	\$ 139 380 801
Discount Factor	0.1889	0.1722	0.1570	0.1431	0.1305	0.1189	0.1084	0.0988
Annual Present Value (Discounted Cash Flow)	\$ 26 331 989	\$ 24 003 636	\$ 21 881 163	\$ 19 946 366	\$ 18 182 649	\$ 16 574 885	\$ 15 109 284	\$ 13 773 276
Cumulative Discounted Cash Flow	\$737 002 943	\$761 006 579	\$782 887 742	\$802 834 107	\$821 016 756	\$837 591 641	\$852 700 925	\$866 474 202
Total Capital Investment + Interest								
Net Present Worth								

## C8. GA Cumulative Cash Flow Sheet

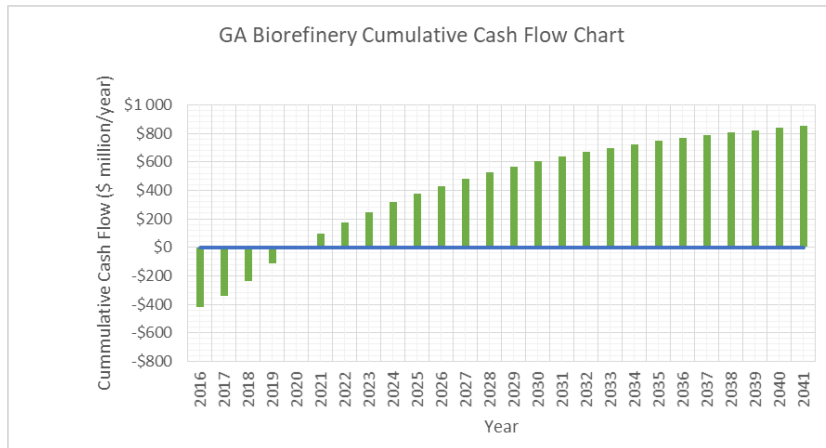


Figure 80 - GA Cumulative Cash Flow

Table 84 - GA 25-Year Cumulative Cash Flow

Year	Financial Year	Cumulative cash flow (\$ million/year)
2016	0	-\$421.93
2017	1	-\$338.66
2018	2	-\$234.15
2019	3	-\$112.80
2020	4	-\$2.19
2021	5	\$98.64
2022	6	\$177.65
2023	7	\$249.67
2024	8	\$315.32
2025	9	\$375.16
2026	10	\$429.72
2027	11	\$479.45
2028	12	\$524.78
2029	13	\$566.10
2030	14	\$603.77
2031	15	\$638.11
2032	16	\$669.42
2033	17	\$697.95
2034	18	\$723.96
2035	19	\$747.68
2036	20	\$769.29
2037	21	\$789.00
2038	22	\$806.96
2039	23	\$823.33
2040	24	\$838.26
2041	25	\$851.86

## Appendix D. CHP Baseline Aspen Simulation Process Flow Diagram

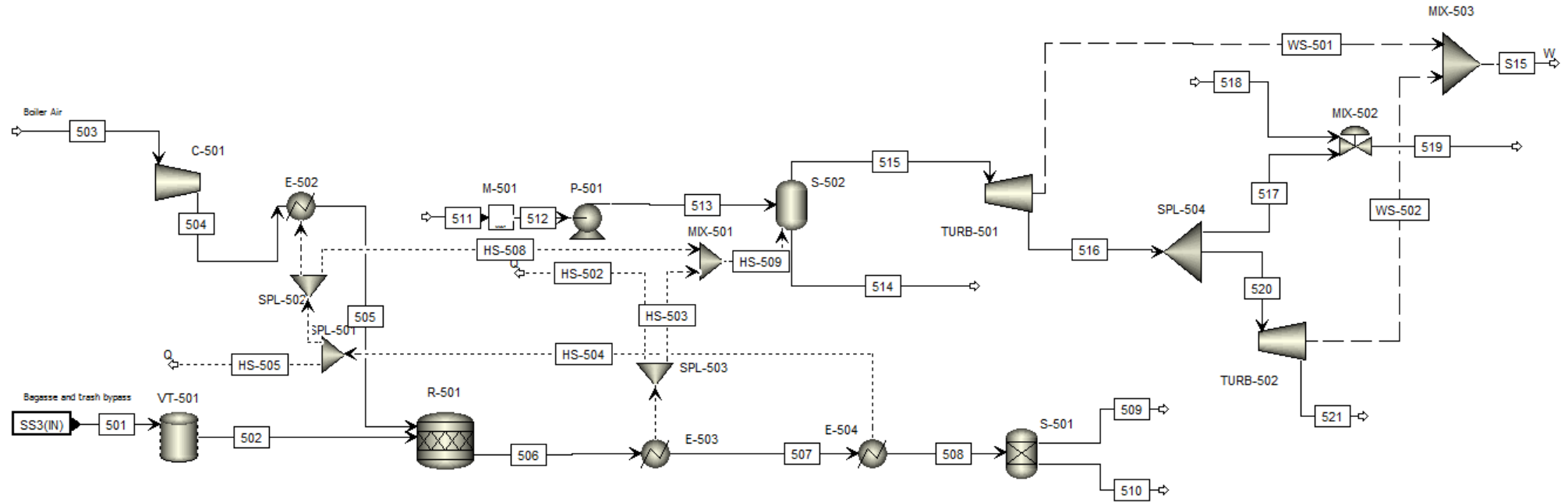


Figure 81 - CHP Baseline Aspen PFD

## Appendix E. Equipment Purchase Cost Reference Sheet

Table 85 - Equipment Purchase Costs for Plant Equipment (adapted from (Towler & Sinnott, 2012))

Purchased Equipment Cost ( $C_e$ ) Constants, $C_e = a + bS^n$ (US Gulf Coast basis, Jan 2007, CEPCI = 509.7)				Reference	Sinnott & Towler, 2012, Pg 313	
Equipment	Units for Size, S	$S_{lower}$	$S_{upper}$	a	b	n
<b>Agitators and Mixers</b>						
Propellor	driver power, kW	5	75	15000	990	1.05
Spiral ribbon mixer	driver power, kW	5	35	27000	110	2
Static Mixer	Litres/s	1	50	500	1030	0.4
<b>Boilers</b>						
Packaged, 15 to 40 bar	kg/h steam	5000	200000	106000	8.7	1
Field erected, 10 to 70 bar	kg/h steam	20000	800000	110000	45	0.9
<b>Centrifuges</b>						
High-speed disk	diameter, m	0.26	0.49	50000	423000	0.7
Atmospheric suspended basket	power, kW	2	20	57000	660	1.5
<b>Compressors</b>						
Blower	m <sup>3</sup> /h	200	5000	3800	49	0.8
Centrifugal	driver power, kW	75	30000	490000	16800	0.6
Reciprocating	driver power, kW	93	16800	220000	2300	0.75
<b>Conveyers</b>						
Belt, 0.5 m wide	length, m	10	500	36000	640	1
Belt, 1.0 m wide	length, m	10	500	40000	1160	1
Bucket elevator, 0.5 m bucket	length, m	10	30	15000	2300	1
<b>Crushers</b>						
Reversible hammer mill	tonne/h	30	400	60000	640	1
Pulverizers	kg/h	200	4000	14000	590	0.5
<b>Crystallizers</b>						
Scraped surface crystallizer	length, m	7	280	8400	11300	0.8
<b>Dryers</b>						
Direct contact rotary	area, m <sup>2</sup>	11	180	13000	9100	0.9
Pan	area, m <sup>2</sup>	3	20	8700	6800	0.5
Spray dryer	evap rate kg/h	400	4000	350000	1900	0.7

Equipment	Units for Size, S	S <sub>lower</sub>	S <sub>upper</sub>	a	b	n
<b>Evaporators</b>						
Vertical tube	area, m <sup>2</sup>	11	640	280	30500	0.55
Agitated falling film	area, m <sup>2</sup>	0.5	12	75000	56000	0.75
<b>Exchangers</b>						
U-tube shell and tube	area, m <sup>2</sup>	10	1000	24000	46	1.2
Double pipe	area, m <sup>2</sup>	1	80	1600	2100	1
Thermosyphon reboiler	area, m <sup>2</sup>	10	500	26000	104	1.1
U-tube kettle reboiler	area, m <sup>2</sup>	10	500	25000	340	0.9
Plate and frame	area, m <sup>2</sup>	1	500	1350	180	0.95
<b>Filters</b>						
Plate and frame	capacity, m <sup>3</sup>	0.4	1.4	110000	77000	0.5
Vacuum drum	area, m <sup>2</sup>	10	180	-63000	80000	0.3
<b>Furnaces</b>						
Cylindrical	duty, MW	0.2	60	68500	93000	0.8
Box	duty, MW	30	120	37000	95000	0.8
<b>Pressure vessels</b>						
Vertical, 304 ss	shell mass, kg	120	250000	15000	68	0.85
Horizontal, 304 ss	shell mass, kg	120	50000	11000	63	0.85
<b>Pumps and drivers</b>						
Single-stage centrifugal	flow, L/s	0.2	126	6900	206	0.9
Explosion proof motor	power, kW	1	2500	-950	1770	0.6
Condensing steam turbine	power, kW	100	20000	-12000	1630	0.75
<b>Reactors</b>						
Jacketed, agitated	volume, m <sup>3</sup>	0.5	100	53000	28000	0.8
Jacketed, agitated, glass lined	volume, m <sup>3</sup>	0.5	25	11000	76000	0.4
<b>Tanks</b>						
Floating roof	capacity, m <sup>3</sup>	100	10000	97000	2800	0.65
Cone roof	capacity, m <sup>3</sup>	10	4000	5000	1400	0.7

## Appendix F. Feed and Cleaning Schedules

Figure 82, Figure 83 and Figure 84 show the feed and cleaning schedules for the catalytic reactor and fermenters. The 3 xylitol reactors have a residence time of 2.5 hours and it is assumed that 1 hour is required for cleaning and refilling. A similar half day schedule is used for 3 fermenters used in CA production, whereas the 8 GA fermenters operate individually for 28.5h and use 6 h of cleaning and refilling owing to the larger GA production capacity.

### F1. Xylitol Reactor Feed and Cleaning Schedule

1	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
2	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
3	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5

Figure 82 - Xylitol Catalytic Reactor (x3) feed and Cleaning Schedule (half-hour schedule)

### F2. CA Fermenter Feed and Cleaning Schedule

1	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
2	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
3	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5

Figure 83 - CA Fermenter (x3) Feed and Cleaning Schedule (half-day schedule)

### F3. GA Fermenter Feed and Cleaning Schedule

1	1	2	3	4	5	6	0.5	1	2	3	4	5	6	7
2	28	1	2	3	4	5	6	0.5	1	2	3	4	5	6
3	27	28	1	2	3	4	5	6	0.5	1	2	3	4	5
4	26	27	28	1	2	3	4	5	6	0.5	1	2	3	4
5	25	26	27	28	1	2	3	4	5	6	0.5	1	2	3
6	24	25	26	27	28	1	2	3	4	5	6	0.5	1	2
7	23	24	25	26	27	28	1	2	3	4	5	6	0.5	1
8	22	23	24	25	26	27	28	1	2	3	4	5	6	0.5

Figure 84 - GA Fermenter (x8) Feed and Cleaning Schedule (hourly schedule)