CONVERSION OF HARDWOODS TO ETHANOL:

DESIGN AND ECONOMICS OF DELIGNIFICATION AND

ENZYME RECYCLING

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LIST OF ACRONYMS

CS	Carbon Steel
CSL	Corn Steep Liquor
CW	Cooling Water
DB	Declining Balance
FPU	Filter Paper Units
GPM	Gallons Per Minute
IRR	Internal Rate of Return
MACRS	Modified Accelerated Cost Recovery System
MAQ	Methyl Antrhaquinone
MESP	Minimum Ethanol Selling Price
MM	Millions
NREL	National Renewable Energy Laboratory
ODT	Oven Dry Ton
PFD	Process Flow Diagram
SS	Stainless Steel
SSF	Simultaneous Saccharification and Fermentation
WWT	Waste Water Treatment

SUMMARY

The objective of this study was to investigate the possibility of recycling enzymes during saccharification of cellulose for the production of ethanol from woodchips. To make enzyme recycling feasible and economical when woodchips are processed for ethanol, the lignin in the wood is to be removed before the enzymes are added. The enzymes are otherwise irrecoverably adsorbed onto the lignin and are not available for reuse. Since enzymes constitute a major part of the input costs, second only to the feedstock, the ability to reuse the enzymes could lead to a considerable decrease in the production cost of ethanol. Tulip poplar woodchips were selected as the feedstock. The chips were pretreated by exposure to steam at 350 psig (2.51 MPa) for 4 minutes using a continuous digester. Different delignification methods with recovery of byproducts were investigated. Alkali extraction, using dilute NaOH for the removal of lignin after steam pretreatment, was used as the base case against which all other processes were compared. Recovery of furfural and methanol, produced during the pretreatment of the woodchips, for sale as byproducts was one modification to the alkali extraction process that was investigated. The conversion of xylose to furfural and the recovery of the furfural as a byproduct was the third case explored. Solvent extraction using a 50:50 ethanol-water mixture instead of extraction with NaOH was the fourth case examined.

Process flow sheets were then developed to recycle the enzymes during the hydrolysis and fermentation of this prehyrolysed and delignified wood. Two reactor setup schemes were examined for enzyme recycling. One scheme involved a single train of reactors,

with the whole pretreated slurry flowing from one reactor to the next, whereas, in the other scheme, the slurry was split among parallel trains of reactors.

The activity loss of the enzymes was modeled such that a part of the enzymes entering the reactor lost all their activity. The loss of activity in multiple steps, with enzymes losing only some of their activity, was also modeled. Here the enzymes entering the reactor constituted a mixture with different activities instead of all the enzymes having the same activity like in the previous single step model.

The different delignification processes and the enzyme recycling schemes were modeled in Aspen Plus and equipment design specifications were developed. Detailed cash flow analysis was performed to compare the different cases and develop an optimal scheme. Recovering methanol and furfural reduced the minimum ethanol selling price. High temperature ethanol–water pretreatment and lignin extraction reduced the minimum ethanol selling price compared to the base case of steam pretreatment followed by alkali extraction. Enzyme recycling also reduces the minimum ethanol selling price. The magnitude of the price reduction depends on the recycling scheme selected and the rate of enzyme deactivation, which has not been measured.

CHAPTER 1

INTRODUCTION

Global warming and the need for energy security have provided an impetus to search for clean and renewable sources of energy. Ethanol is one such fuel which reduces toxic emissions, reduces the dependence on imported oil, and improves local economy. Ethanol cannot be produced on an increasing scale from the current technology based on corn starch as it competes with food production. Instead, lignocellulosic materials, such as wood chips, crop residues and grasses, provide a practical alternative as they are widely available. The use of hardwoods as the source of feedstock is explored here, as they are widely available in South East USA and easier to process than softwoods. Hardwoods are composed of cellulose surrounded by hemicellulose and lignin. The hemicellulose and cellulose components are polymers of sugar molecules, which on hydrolysis break down into simple sugars. These sugars are fermented to ethanol, which is recovered and concentrated by distillation. Figure 1.1 shows the major process steps involved in the production of ethanol. Cellulose is the primary source of the ethanol and is hydrolyzed in Area 300 of Figure 1.1. Cellulase enzymes are usually used for hydrolysis and yeast cells for the fermentation of the simple sugars formed to ethanol. The woodchips are first pretreated to increase the accessibility of the cellulose to enzymes. This is done by hydrolyzing the hemicellulose and disrupting or dissolving the lignin, both of which form a protective sheath around the cellulose. The crystallinity of the cellulose is also reduced during the pretreatment.

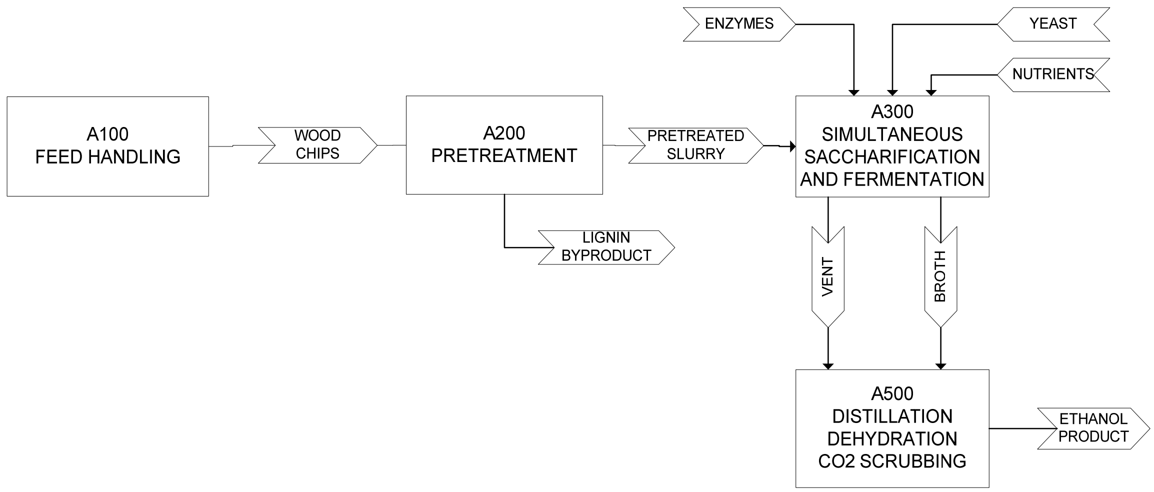

Figure 1.1: Ethanol Production Process Overview

To make cellulosic ethanol cost competitive with corn ethanol and petroleum, the maximum utilization of all the components of the feedstock is to be made. Selective extraction of lignin for use as a source of heat improves the economics of ethanol production. Recovery by delignification of wood before hydrolysis is explored as it makes recycling of enzymes feasible by reducing the nonproductive binding of cellulases during hydrolysis; reduces the flow rate of the stream entering the hydrolysis section and hence reduces equipment capital costs. NaOH pulping is taken as the base case for lignin recovery. Modifications made to this process, to include recovery of byproducts, and the organosolv pulping process are compared to this base case to see the economic viability.

The enzymes constitute the second major cost component after the feedstock and account for 50 percent of the hydrolysis costs. Reusing the enzymes to reduce their cost would make the process more economical. The recycling is modeled with different deactivation schemes and for different setups to see the trends in the economics with changing process variables.

CHAPTER 2

LITERATURE SURVEY

2.1 Delignification Methods

Delignification is the process of removal of lignin by cleavage of inter-unit bonds into shorter fragments and water soluble monomeric products. The lignin fragments undergo condensation reactions to repolymerize and become insoluble when pulping time is high.

The effect of steam hydrolysis on tulip poplar woodchips was studied for a pressure range of 300 psig – 350 psig with residence times between three and seven minutes (Fieber, 1982). Optimal conditions were found to be 350 psig with a residence time of 4 minutes. A pulp with 58.5 percent cellulose was obtained retaining 97 percent of the original cellulose. Lignin extraction of this with 0.1 N NaOH gave a pulp with 8 percent residual lignin. 89 percent of the hemicellulose degraded into soluble sugars and volatiles.

Recovery of the volatiles – furfural, methanol and acetic acid - was investigated and it was shown that methanol and furfural recovery improves the economics of the process although acetic acid recovery wasn't practical (Mann, 1983). The xylose produced during hydrolysis of the hemicellulose is not used for the production of ethanol as the current yeasts used for fermentation of six carbon sugars cannot convert the five carbon sugars like xylose. This xylose was instead degraded in a plug flow reactor with a residence time of 10.4 minutes under dilute acidic conditions. 93 percent of the xylose was converted to furfural, which could be recovered to be sold as a byproduct.

The removal of lignin with 95 percent ethanol solution was found to be not as effective as extraction with dilute NaOH (Fieber, 1982). Solvent extraction with a 50:50 ethanol-water mixture on the other hand proved to be more effective as water increased the hydrolytic activity promoting dissolution of the lignin. Sodium bicarbonate was used as a buffer with methylanthraquinone added to stabilize the carbohydrates and improve lignin removal (Faass, 1985). Pulps with a yield of over 60 percent and a residual lignin content of less than 5 percent can be produced using solvent pulping technique when operating temperature is between 220^{o}C and 240^{o}C and the corresponding residence time is between 10 minutes and 40 minutes. Pulping temperature of 220^{o}C was found to be suitable for enzymatic hydrolysis of the cellulose as increasing the severity decreased the susceptibility of the pulp to hydrolysis (Sweeney, 1985).

Organosolv pretreatment is the extraction of lignin by aqueous-organic solvents, such as ethanol, triethylene glycol, tetrahydrofurfuryl alcohol. Ethanol is a good solvent as high liquor penetration into the wood occurs and it is easily recoverable. There isn't significant production of inhibitors to hydrolysis during the pulping with ethanol. This process was first developed by General Electric and University of Pennsylvania and later adapted to the pulp and paper industry as the Alcell process (Aziz, 1989). The process is now being developed as the Lignol process for production of ethanol (Pan, Arato et al.,2005). It uses a 50:50 (w/w) mixture of ethanol and water at around 200^{o}C and 400 psi to extract the lignin. At 600 psig and 230^{o}C, 85 percent of the lignin in the feedstock was found to dissolve in the ethanol-water mixture as long as the ethanol concentration was over 20 percent by volume (Faass, 1985). Catalysts were not required at such high temperatures as acids released from the wood act as the catalysts (Sarkanen,1980).

2.3 Enzyme Recycling

Enzymatic hydrolysis takes place by the adsorption of the cellulase enzymes onto the surface of the cellulose, breakdown of cellulose to fermentable sugars, and the desorption of enzymes into the supernatant (Gregg and Saddler,1996) . 88 percent of the enzymes in the supernatant can be recovered by adsorbing onto fresh substrate due the natural affinity of cellulases for cellulose (Tu, Chandra et al.,2007). But some of the enzymes remain bound to the residual substrate after hydrolysis. It was found that when the residual substrate was brought in contact with fresh substrate the enzymes evenly partitioned between the fresh and the residual substrates and both substrates were hydrolyzed throughout the reaction. This technique is simple and doesn't involve high costs due to chemicals. Experiments with birch containing 4 percent residual lignin and 32 percent residual lignin showed that lower cellulase activities were recovered with the latter as the lignin irrecoverably adsorbs the enzymes (Lee, Yu et al.,1995). Another factor contributing to the lower observed activity is the presence of recycled residual substrate. The substrate was found to become increasingly recalcitrant as the hydrolysis progressed and the observed enzyme activity is lower for a mixture of fresh and recycled substrate when compared with only fresh substrate. Higher enzyme activity was recovered when the non-cellulosic residue was low. This could be because the cellulases are more tightly adsorbed with cellulose-free lignin than with carbohydrate-associated lignin. The recycled enzymes retained almost all their activity when the lignin content was less than 17 percent but only 70 percent could be recovered when the lignin content was as high as 50 percent (Lee, Yu et al.,1995) .

2.3 Process Simulation

Processes for production of fuel grade ethanol from poplar and corn stover were developed by NREL (Wooley, 1999), (Aden, 2002). The feedstock was subjected to steam pretreatment followed by saccharification of the cellulose using cellulase enzymes. Z.mobilis bacterium was used as the ethanologen for converting the simple sugars to ethanol. The yields of the different products were calculated from experimental data and assumed near term research developments. The reports also have detailed design specifications for the equipment to be used at commercial scale along with cost scaling factors. A physical properties database was developed for the biofuel components so that the processes could be modeled in Aspen Plus simulation software (Wooley, 1996).

CHAPTER 3

DELIGNIFICATION PROCESSES

3.1 Introduction

An investigation into the effects of recovering lignin after pretreatment of tulip poplar woodchips by steam hydrolysis is performed. Removal of lignin helps reduce the size of the hydrolysis reactor, decreases the distillation duties and makes enzyme recycling feasible(Lander, Malcolm et al.,1983). Different options for recovery of lignin and recovery of byproducts are presented. The models were simulated in Aspen Plus using the physical property database for biofuel components developed by NREL (Wooley and Putsche,1996). Appendix B has the detailed process flow diagrams and Appendix C gives the material balances. The detailed equipment specifications and costs are given in Appendix D.

3.2 Feedstock Composition

The feedstock composition on a dry basis is shown in Table 3.2.2 (Fieber,1982). The cellulose accounts for 49.7% of the wood while the hemicellulose and lignin account for approximately 23% and 27% respectively. Three quarters of the hemicellulosic portion of the wood is xylan and the remaining portion is composed of mannan, arabinan and galactan.

Table 3.2.1: Feedstock

Feedstock	Tulip Poplar Woodchips
Feed Rate	1000 ODT/day
Moisture Content	50%

Table 3.2.2: Feedstock Composition

Component	% Dry Basis
Glucan	50.0
Hemicellulose	22.7
Xylan	17.3
Mannan	3.7
Arabinan	0.7
Galactan	1.0
Lignin	26.9
Ash	0.4

3.3 Base Case: NaOH Delignification

3.3.1 Process overview

Figure 3.3.1 gives a general overview of the major process steps involved. The wood is first pretreated by steam hydrolysis followed by extraction of the lignin from the wood using dilute NaOH solution. Simultaneous Saccharification and Fermentation (SSF) is then performed on the delignified hydrolyzate slurry using cellulase enzymes for the hydrolysis of cellulose and Z.Mobilis Bacterium for fermentation of the simple sugars to ethanol.

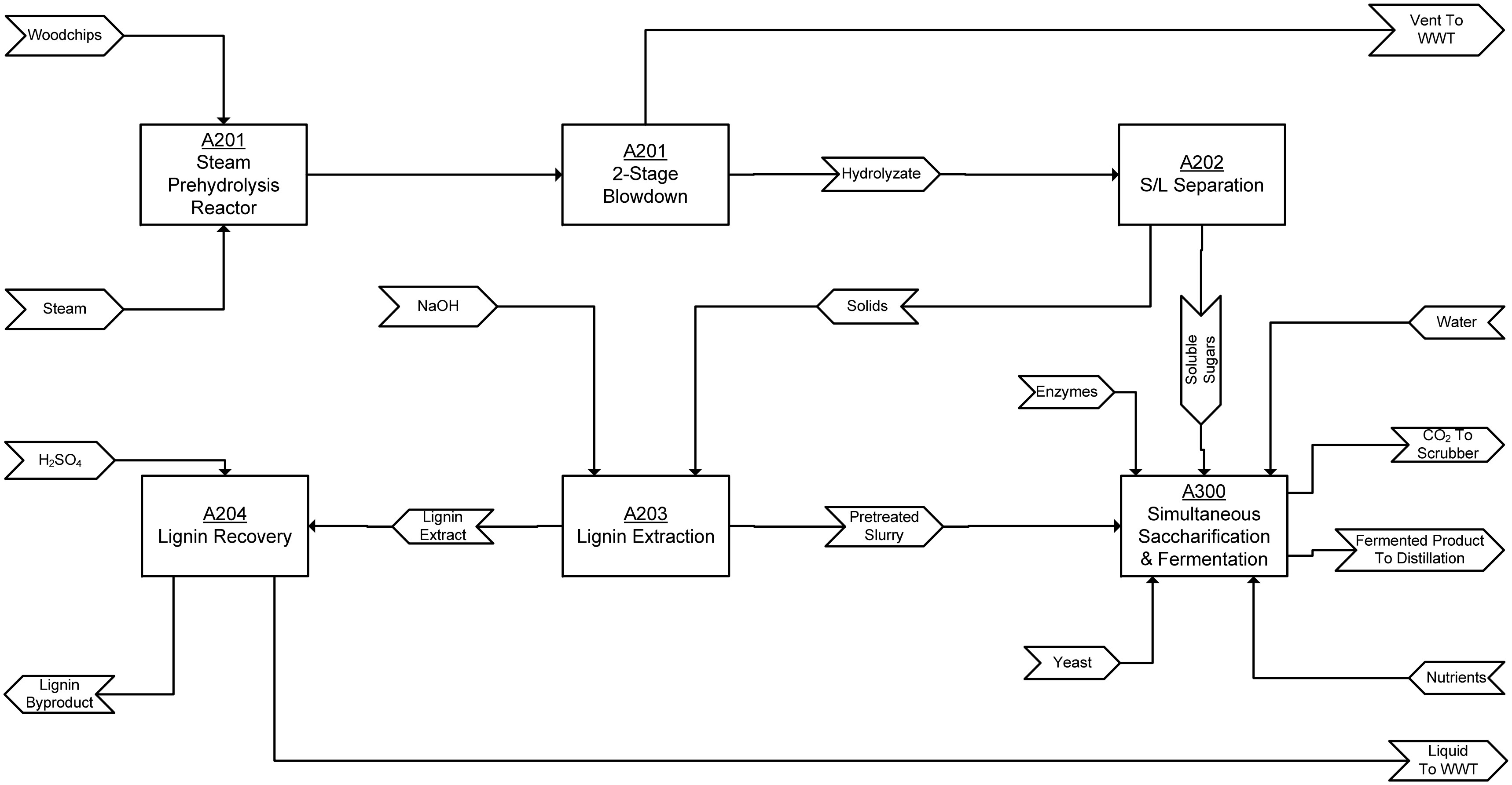

Figure 3.3.1: Overall Process Flow Diagram- Base Case - NaOH Delignification

3.3.2 Steam Hydrolysis (A201- Figure B.1.1)

The feedstock is tulip poplar woodchips (101) and is sent to the prehydrolysis reactor (R201), which is maintained at 350 psig using steam. The biomass is explosively discharged after a residence time of 4 minutes. 89 percent of the hemicellulose portion along with 3 percent of the cellulose is converted to soluble sugars: xylose, mannose, arabinose, galactose and glucose. Table 3.3.2.2 shows the composition of the slurry coming out of the pretreatment digestor (Mann,1983). The hemicellulose hydrolysis makes the cellulose accessible for enzymatic hydrolysis.

Table 3.3.2.1 Pretreatment Reactor Conditions

Residence Time	4 mins
Pressure	350 psig
Solids in the Reactor (Inlet)	38%
Temperature	467^{o}F

The soluble sugars further degrade to form furfural, methanol and acetic acid. These volatile products are toxic to microbes in the SSF section and need to be removed. Too severe operating conditions increase the formation of these degradation products.

Table 3.3.2.2: Composition of Feedstock Before and After Pretreatment

	Before	After
Component	% Dry Basis based on input	% Dry Basis based on output
Glucan	50.0	48.6
Hemicellulose	22.7	2.6
Xylan	17.3	1.5
Mannan	3.7	0.8
Arabinan	0.7	0.1
Galactan	1.0	0.2
Soluble Sugars		8.7
Glucose		1.1
Xylose		5.7
Mannose		1.1
Arabinose		0.1
Galactose		0.7
Lignin	26.9	35.8
Ash	0.4	0.4
Volatiles		4
Methanol		0.3
Furfural		1.6
Acetic Acid		2.1

3.3.3 Two Step Flash (A201 – Figure B.1.1)

The wood product from the reactor (202) is first flashed at 50 psig to recover some of the heat energy using the vapor stream from the flash. A large amount of the volatile organics (methanol, acetic acid and furfural) is vaporized along with water. The bottoms product (203) is flashed again to atmospheric pressure. In the base case, the vapor streams from both the flashes are condensed and sent to waste water treatment.

Table 3.3.3.1: Flash System

Furfural Removal	78.2%
Methanol Removal	64.7%
Acetic Acid Removal	21.9%

3.3.4 Solids Wash (A202 – Figure B.1.2)

The bottoms product from the flash system (205) is sent to a filtration press where the solids are washed and pressed to 50 percent solid content. 90 percent of the soluble sugars in the hydrolyzate are removed in the filtrate (209). A part of the filtrate is recycled to slurry the inlet solids stream to 6 percent solids to ensure uniform filtration. The solid cake (208) is sent to the delignification section to separate the lignin from the wood.

Table 3.3.4.1: Solids Wash

Solids in Cake	50%
Soluble Sugars Removed	90%

3.3.5 Delignification (A203 – Figure B.1.3)

The washed solids (208) are sent to a mixing tank (M-201) in which a pH of 12 is maintained using NaOH (210) to dissolve the lignin in the wood. The final solution has about 15 percent by weight lignin and a cellulose residue containing 9.5 percent lignin is obtained (Mann,1983). The slurry is washed and pressed to remove 90 percent of the dissolved lignin and the delignified wood pulp containing 50 percent moisture is sent to the Simultaneous Saccharification and Fermentation (SSF) tanks (T-301). The dissolved lignin (212) is sent to the lignin recovery mixing tank (M-202).

Table 3.3.5.1: Lignin Extraction

Base used	NaOH
pH	12
Residual Lignin	9.5%
Dissolved Lignin Recovered	90% w/w

3.3.6 Lignin Recovery (A204 – Figure B.1.4)

In the lignin recovery tank (M-202), H_2SO_4 is added to neutralize the NaOH and lower the pH to below 3 so that the lignin precipitates. The slurry is then pressed to get a lignin product (216) containing 50 percent solids. 52 percent of the lignin present in the pretreated slurry is removed.

Table 3.3.6.1: Lignin Recovery

Acid Used	H_2SO_4
Percent Solids in Lignin Product	50%
Overall Lignin Recovery	52%

3.3.7 Simultaneous Saccharification and Fermentation (A300 – Figure B.1.5)

The pretreated solids product after delignification (301) contains 50 percent solids. This is sent to the SSF reactors and is diluted to about 10 percent solids using water (305) and the filtrate containing the soluble sugars from the pretreatment product wash section (209). Cellulase is added at a loading of 15 FPU/g cellulose. Corn Steep Liquor (CSL), Diammonium Phosphate (DAP) nutrients are also fed to the reactor along with 10 percent inoculum for the fermentation. The reactions occurring in the SSF reactors are shown in Table 3.3.7.2 and Table 3.3.7.3 (Wooley, Ruth et al.,1999) . Mannose and Galactose are assumed to have the same conversions as glucose.

Table 3.3.7.1: SSF Conditions

Temperature	30°C
Initial Solids Level	10%
Residence Time	3 days
Cellulase Loading	15 FPU/g cellulose
Inoculum	10%
Corn Steep Liquor Level	0.25%
Ethanol Formed	22.3 MMgal/yr
Ethanol Concentration	29 g/l

Table 3.3.7.2: Saccharification Reactions

Reaction			**Conversion**	
(Cellulose)n	+ m H2O	→m Glucose Olig	Cellulose	0.068
(Cellulose)n	+ ½n H2O	→½n Cellobiose	Cellulose	0.012
(Cellulose)n	+ n H2O	→n Glucose	Cellulose	0.8
Cellobiose	+ 2 H2O	→2 Glucose	Cellobiose	1.0

Table 3.3.7.2: Fermentation Reactions

Reaction					Conversion of Glucose
Glucose		$\rightarrow$2 Ethanol	+ 2 CO_2		0.92
Glucose +	1.2 NH_3	$\rightarrow$6 *Z. mobilis*	+ 2.4 H2O	+ 0.3 O_2	0.027
Glucose +	2 H_2O	$\rightarrow$2 Glycerol			0.002
Glucose +	2 CO_2	$\rightarrow$2 Succinic Acid			0.008
Glucose		$\rightarrow$3 Acetic Acid			0.022
Glucose		$\rightarrow$2 Lactic Acid			0.013

3.4 Modifications to Base Case

3.4.1 Case 2: NaOH Delignification with Furfural and Methanol Recovery

(A206–Figure B.2.5)

The vapor streams (204A, 206A) from the flash vessels shown in Figure B.2.1 are rich in volatile organics. Figure 3.4.1.1 shows the modified process flow diagram with the recovery of methanol and furfural from the overhead streams. These streams are condensed and sent to a stripping column (D-201) in Figure B.2.5 where it is assumed that most of the acetic acid is removed in the bottoms stream (226) while the overhead stream (228) is rich in furfural and methanol. A decanter (T-203) is used to separate this overhead stream (228) into an 89 percent furfural rich stream (230) and the remaining into a methanol rich stream (229). Each of these streams is then sent to their respective recovery columns.

77 percent of the methanol entering this section is recovered as a 99.5 percent pure distillate stream (234) from the methanol recovery column (D-202) while the furfural

stream is dehydrated in the furfural dehydration column (D-203) to a 98 percent pure stream (236) recovering 60 percent of the furfural in the feed.

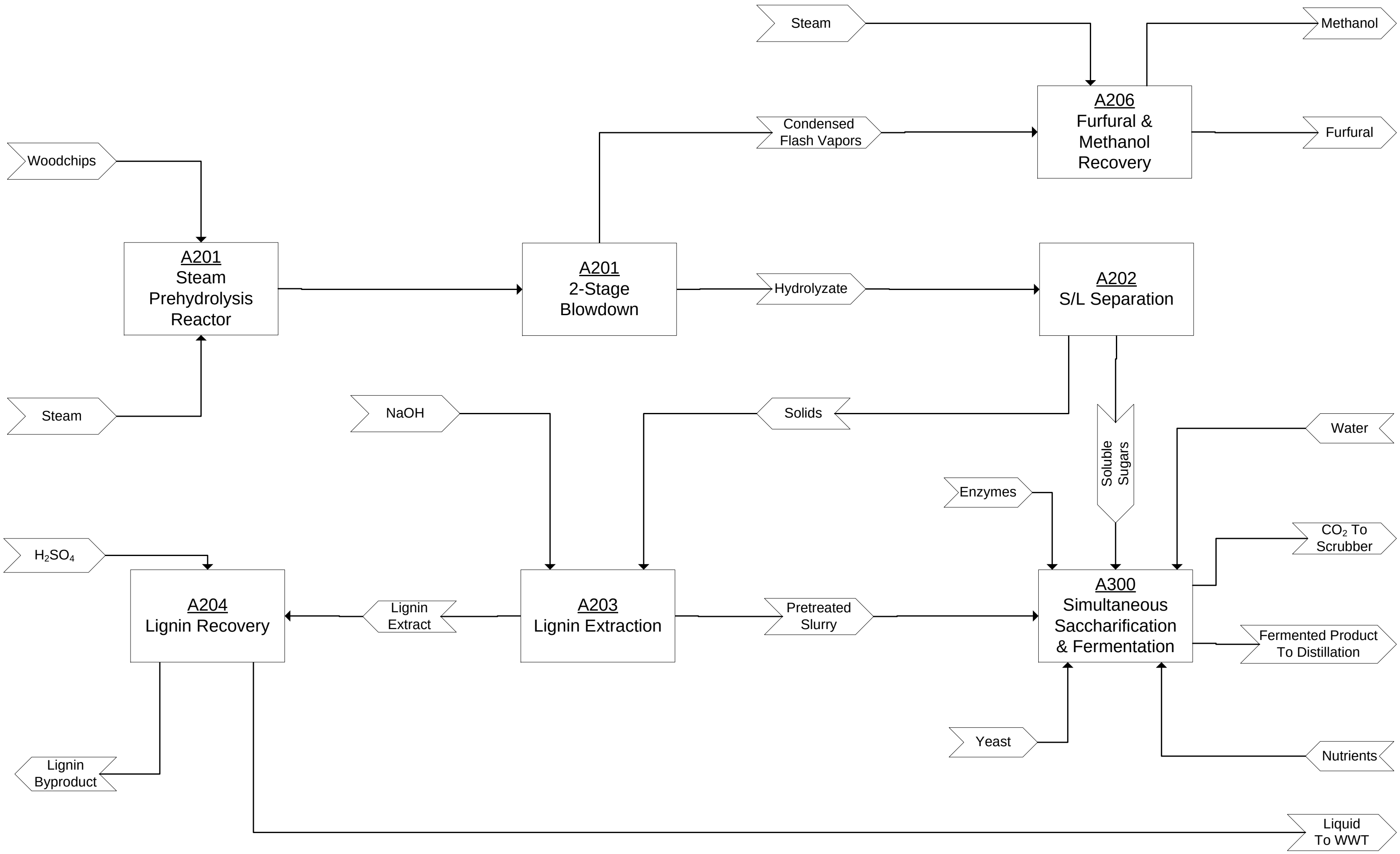

Figure 3.4.1.1: Overall Process Flow Diagram: Case 2 - NaOH Delignification with Furfural and Methanol Recovery

3.4.2 Case 3: NaOH Delignification with Xylose Converted to Furfural (A205 – Figure B.3.5)

The slurry from the flash vessels is washed and filtered to remove the soluble sugars formed due to hydrolysis of the hemicellulose during pretreatment. This soluble sugars stream is rich in xylose and is sent to the SSF section in the Base Case. But the current yeast strains cannot ferment xylose to ethanol. So the option of converting xylose to furfural and selling this as a byproduct is considered to see if it is more economical. The major process steps for this modified case are shown in Figure 3.4.2.1.

The soluble sugars stream (209) from the solids wash section is sent to a plug flow reactor after heating to the reaction temperature of 215^{o}C. The solution is brought to a concentration of 0.02N using H_2SO_4 (218) and 93 percent of the xylose and arabinose degrade to furfural (Mann,1983). NaOH (221) is then added to the product stream to neutralize the H_2SO_4 and the acetic acid so as to prevent corrosion. The heat from this stream is used to raise the temperature of the feed stream (209) and is then sent to the furfural and methanol recovery section to separate the furfural as discussed in Case 2.

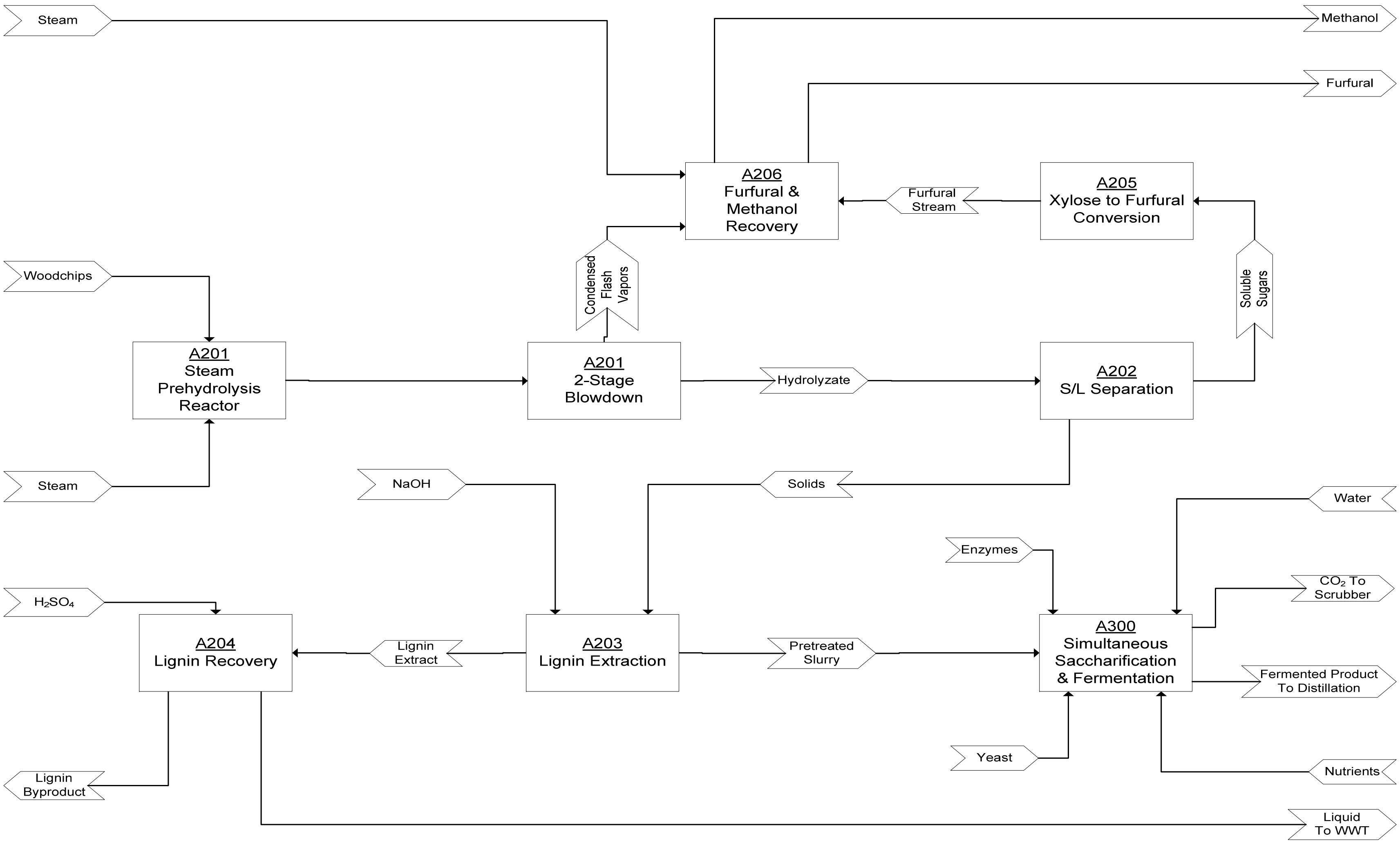

Figure 3.4.2.1: Overall Process Flow Diagram: Case 3 – NaOH Delignification with Xylose Converted to Furfural

3.4.4 Case 4: Organosolv Pretreatment

3.4.4.1 Process overview

The hydrolysis of lignin to a soluble product is carried out with ethanol-water solution in the presence of a buffer agent, sodium bicarbonate, which helps reduce carbohydrate loss and methyl anthraquinone which increases the delignification selectivity. The use of ethanol instead of NaOH reduces the probability of formation of byproducts which have adverse affects during SSF (Lander, Malcolm et al.,1983). 99 percent of the ethanol is recovered during pre-treatment so that the ethanol formation rate in SSF is not product hindered. The process is similar to the base case using NaOH to dissolve the lignin. Figure 3.4.4.1 gives an overview of the process steps involved.

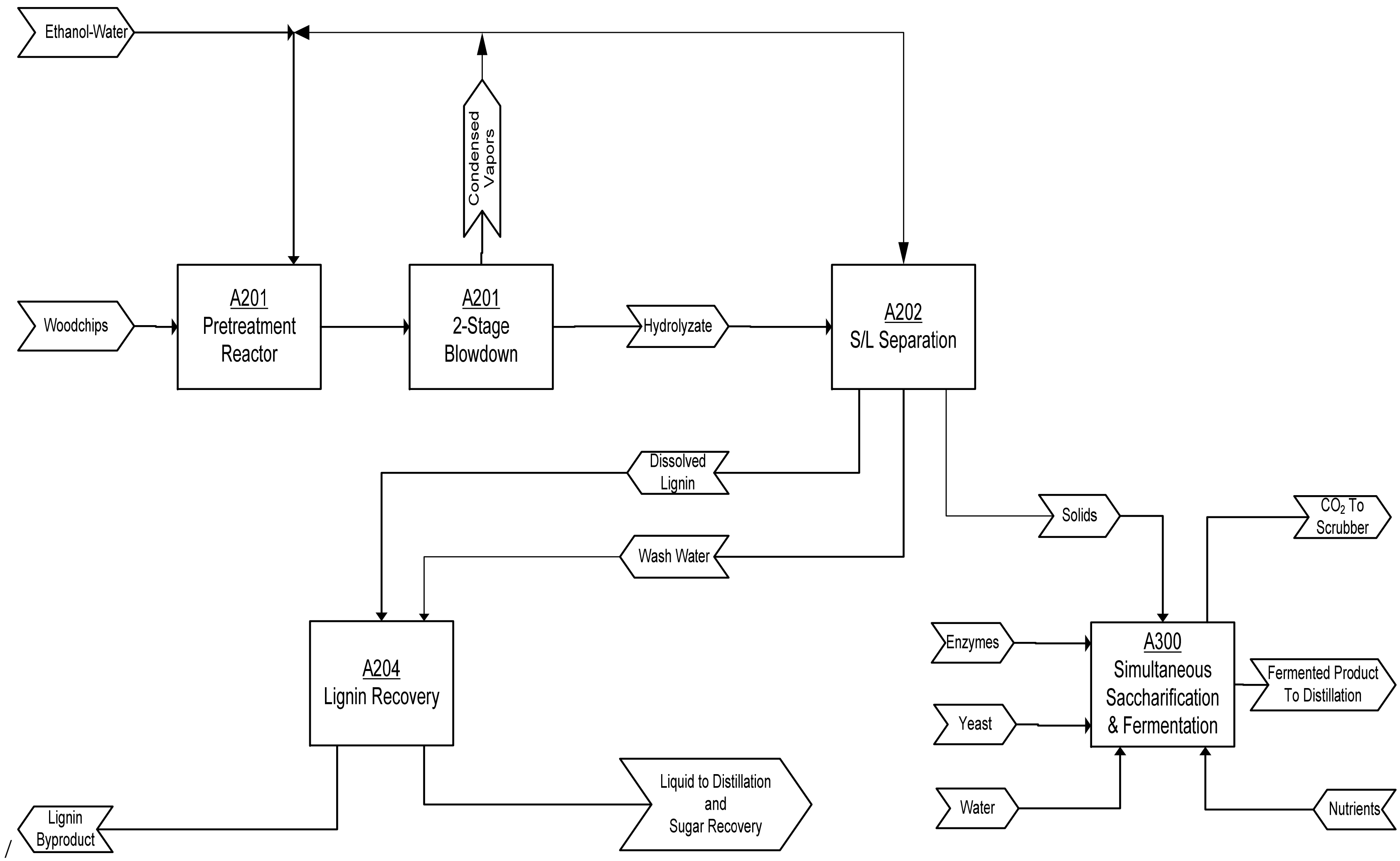

Figure 3.4.4.1: Overall Process Flow Diagram: Case 4 - Organosolv Pretreatment

3.4.4.2 Ethanol and Steam Hydrolysis and Lignin Dissolution (A201 – Figure B.4.1)

The feedstock (101) is sent to a digester (R-201) operating at 600 psig and 446^{o}F. The hydrolysis heat for the feedstock is provided by a partially vaporized stream at 600 psig and 446^{o}F containing 48 percent by mass ethanol and the remaining being water. The heat of vaporization of this stream is used to bring the temperature of the woodchips to the reaction temperature. Excess ethanol is added to ensure that the lignin remains dissolved as long as the concentration of ethanol doesn't fall below 20 percent by volume. Table 3.4.4.2.2 shows the composition of the wood slurry exiting the reactor (Faass,1985).

Table 3.4.4.2.1: Case 4- Pretreatment Reactor Conditions

Pressure	600 psig
Solids in the Reactor (Inlet)	15 percent
Temperature	446^{o}F
Ethanol Concentration	35% w/w
Lignin Dissolved	85%

Table 3.4.4.2.2: Case 4 -Composition of Feedstock Before and After Pretreatment

	% Ethanol-Water Free Basis	
Component	Before	After
Glucan	48.2	40.2
Hemicellulose	21.9	13.1
Xylan	16.7	11.2
Mannan	3.6	1.1
Arabinan	0.7	0.3
Galactan	1.0	0.5
Soluble Sugars		15.5
Glucose		7.5
Xylose		5
Mannose		2.3
Arabinose		0.4
Galactose		0.4
Lignin	25.9	26.2
Soluble Lignin		22.1
Insoluble Lignin		4
MAQ	3.7	3.8
Ash	0.4	0.4
Volatiles		0.8
Methanol		0.1
Furfural		0.2
Acetic Acid		0.5

3.4.4.3 Two Step Flash (A201- Figure B.4.1)

The pretreatment reactor product (202) is first flashed to 81 psig and the bottoms stream from the first flash (203) is flashed again to atmospheric pressure. The vapor streams from the two flashes (204, 206) are used as heating fluids and then recycled. These streams are rich in ethanol and the first flash overhead stream (204A) is used to make up for the ethanol-water stream entering the digester while the second overhead stream (206A) is used to wash the pretreated slurry and separate the dissolved lignin.

3.4.4.4 Solids Wash (A202 – Figure B.4.2)

The bottoms product from the flash system (205) is washed and pressed to remove the dissolved lignin (208). The ethanol concentration is maintained above 20 percent by volume so that the lignin doesn't precipitate and remains in soluble form. The cake is washed again with the condensed vapors from the flash system (206A) to recover the ethanol in the cake. The lignin remaining in the cake precipitates at this point. The wash streams are recycled to the filter press (S-201) as they have high ethanol concentration. The final wash stream (209) is mostly water and is sent to the lignin recovery section to dilute the concentration of ethanol and precipitate out the lignin. The soluble lignin stream (208) is sent to the lignin recovery section (A204) while the cake (301), which is almost free of ethanol, is sent to the SSF section (A300).

Table 3.4.4.4.1: Case 4 -Solids Wash

Percent Solids after Pressing	50%
Dissolved Lignin Removed in Filtrate	98%
Residual Lignin	7.8% of WIS Solids

3.4.4.5 Lignin Recovery (A204 – Fig B.4.3)

The dissolved lignin stream from the solids wash (208) is sent to the lignin precipitation tank (M-202) where the ethanol concentration is reduced to 15 percent using recycled wash water (209). 87 percent of the dissolved lignin precipitates and the slurry (211) is then pressed to 50 percent lignin solids byproduct (213). 73 percent of the lignin present in the feed is recovered. The ethanol in the filtrate (212) is recovered by sending the stream to the distillation system where the ethanol produced in the SSF section is concentrated. The soluble sugars and any lignin present in the filtrate come out as the bottoms of the distillation column. The bottoms product is concentrated in evaporators and the concentrate is sent to a decanter to separate out the sugars and the lignin. Over 50 percent of these sugars are oligomers and cannot be used for ethanol production (Pan, Arato et al.,2005).

Table 3.4.4.5.1: Case 4 -Lignin Recovery

Percent Solids in Lignin Stream	50%
Overall Lignin Recovery	73%

3.5 Economics

3.5.1 Comparison

A detailed cash flow analysis was performed to compare the four possible pretreatment routes by calculating the installed equipment costs, operating costs and byproduct credits. The equipment costs were found using Aspen Icarus Process Evaluator, 2006 and data from the NREL reports for producing ethanol from yellow poplar (Wooley, Ruth et al.,1999) and corn stover (Aden, Ruth et al.,2002). The plant is assumed to operate for twenty years and the depreciation is calculated using the Modified Accelerated Cost Recovery System (MACRS) with a recovery period of seven years. Double declining balance (DDB) depreciation is used in the first four years while the straight line (SL) depreciation method is used in the next three years. A tax rate of 39 percent is used and an Internal Rate of Return (IRR) of 10 percent. The Minimum Ethanol Selling Price (MESP) required for an Internal Rate of Return (IRR) of 10 percent is used to compare the four scenarios. The comparison tables are shown in Chapter 5. Appendix D gives the detailed equipment design data and the cost.

3.5.2 Sensitivity to Furfural Price

The selling price of furfural is varied to see how this affects the economics of Case 2 and Case 3. The minimum price of furfural below which these two cases aren't economically viable over the Base Case is also found. The assessments are shown in Chapter 5.

CHAPTER 4

ENZYME RECYCLING

4.1 Process Overview (A300 - Fig B.5.1)

Figures 4.1.1 and B.5.1 show the general process steps involved in recycling of enzymes during SSF. The product stream from the SSF reactor (306) is sent to a centrifuge (S-301) where 98 percent of the solids are separated as the cake (308) (Wooley, Ruth, et al., 1999). The cake also contains 28 percent of the liquid entering the centrifuge. A part of the cake is purged (310) to avoid buildup of solids in the system, which could otherwise decrease the enzyme activity, and the purged stream is sent to the ethanol recovery section. The percentage purged is a degree of freedom and is optimized such that the cost is minimized. The supernatant (309) is sent to a tank (T-302) to be mixed with the slurry from the pretreatment section (301) so that the enzymes in the supernatant are adsorbed onto the solid feed (301) and can be reused. The residence time in the tank is one hour and the temperature is maintained at 4^{o}C so that the cellulose in the feed from pretreatment isn't hydrolyzed (Lee, Yu et al.,1995). The slurry from the mixing tank (311) is then sent to a second centrifuge (S-302) where 98 percent of the solids and 28 percent of the liquid entering the centrifuge are separated. The solid stream (312), along with the cake from the first centrifuge (308), is sent as feed to the SSF reactor (T301). The supernatant from the second centrifuge (313) is rich in ethanol and is sent to the ethanol recovery section. About 88 percent of the enzymes in the supernatant sent to the mixing tank are recovered. The activity of the cellulase enzymes decreases with each use and fresh cellulases are added to make up for the activity loss. Simulations were run in

Aspen Plus assuming different values for activity loss and for each of these values the percentage of cake purged (310) is changed such that the cost is minimized. A maximum loss of 40 percent is assumed for the viability of yeast cells (Oliveira, De Castro et al.,2000).

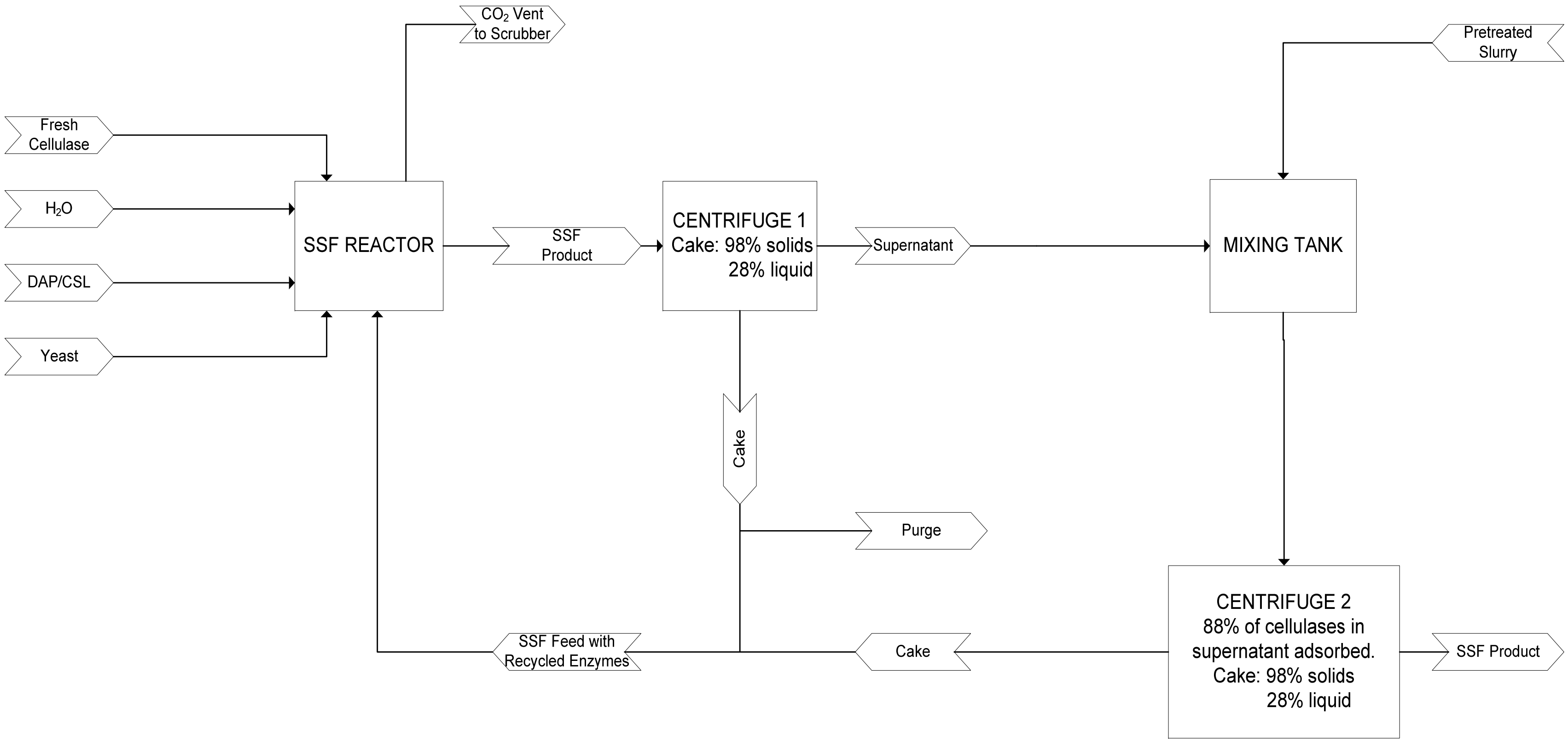

Figure 4.1.1: Recycle Scheme for Single Train of Reactors

4.2 Deactivation Scheme

The deactivation of the cellulase is modeled using a single-step deactivation scheme and a multiple-step deactivation scheme.

4.2.1 Single-Step Deactivation Scheme

A certain percentage, x, of the active cellulases is assumed to be completely deactivated with each pass in the reactor. The process is modeled for different values for x.

$$\begin{array}{lcl} \textit{Active Cellulases} & \xrightarrow{x\%\ \textit{Conversion}} & \textit{Deactivated Cellulases} \\ 50\,FPU/ml & & 0\,FPU/ml \end{array}$$

4.2.2 Multi-Step Deactivation Scheme

The cellulase doesn't become completely inactive but might lose only a part of its activity with each use. In this scheme, the enzymes are modeled as a mixture of cellulases, each with different activity, entering the reactor. With each pass in the reactor a certain percentage of the different cellulases lose their activity to become a less active cellulase.

$$\begin{array}{lclclcl} \textit{Fresh Cellulases} & \xrightarrow{x\%\ \textit{Conversion}} & \textit{Cellu}40 & \xrightarrow{x\%\ \textit{Conversion}} & \textit{Cellu}30 & \xrightarrow{x\%\ \textit{Conversion}} & \textit{Cellu}20 \\ 50\,FPU/ml & & 40\,FPU/ml & & 30\,FPU/ml & & 20\,FPU/ml \end{array}$$

$$\begin{array}{lclcl} \textit{Cellu}20 & \xrightarrow{x\%\ \textit{Conversion}} & \textit{Cellu}10 & \xrightarrow{x\%\ \textit{Conversion}} & \textit{Deactivated Cellulases} \\ 20\,FPU/ml & & 10\,FPU/ml & & 0\,FPU/ml \end{array}$$

It is assumed that the percentage conversion of each type of cellulase,x, to the lower activity cellulase is the same for all types of cellulases entering the reactor. Calculations were performed with several values for x. For each of these values of x, the percentage of overall activity entering the reactor that is lost is then calculated.

4.3 Parallel Reactors Scheme

Here, the hydrolysis and fermentation reactions are carried out in parallel trains of reactors, instead of a single train, as shown in Figures 4.3.1 and B.5.2. The cellulase is recycled using the same procedure as used in the case of a single reactor except that the recovered cellulase is sent to the next train of reactors instead of being sent back to the same reactor. Fresh cellulase is fed only to the first train and then recycled through the remaining trains. The increasing lignin accumulation through the trains affects the enzyme recycling efficiency. The feed from the pretreatment section is split among these trains such that the recycled cellulase activity provides constant enzyme loading for the cellulose entering the reactors. A certain percentage of the active cellulase is assumed to lose its activity with each use through a train of reactors. The distribution ratio of cellulase between the supernatant and the solids in the reactor is unknown. Mass balance of the cellulase was performed for different values of this distribution ratio in each train and the average values were used to find the activity of the cellulase being recycled to the next train of reactors. Appendix A has sample calculations to show how the feed split fractions were calculated.

In this case, the number of parallel trains is a degree of freedom and the optimal number varies for different cellulase activity loss values and also with cellulase cost.

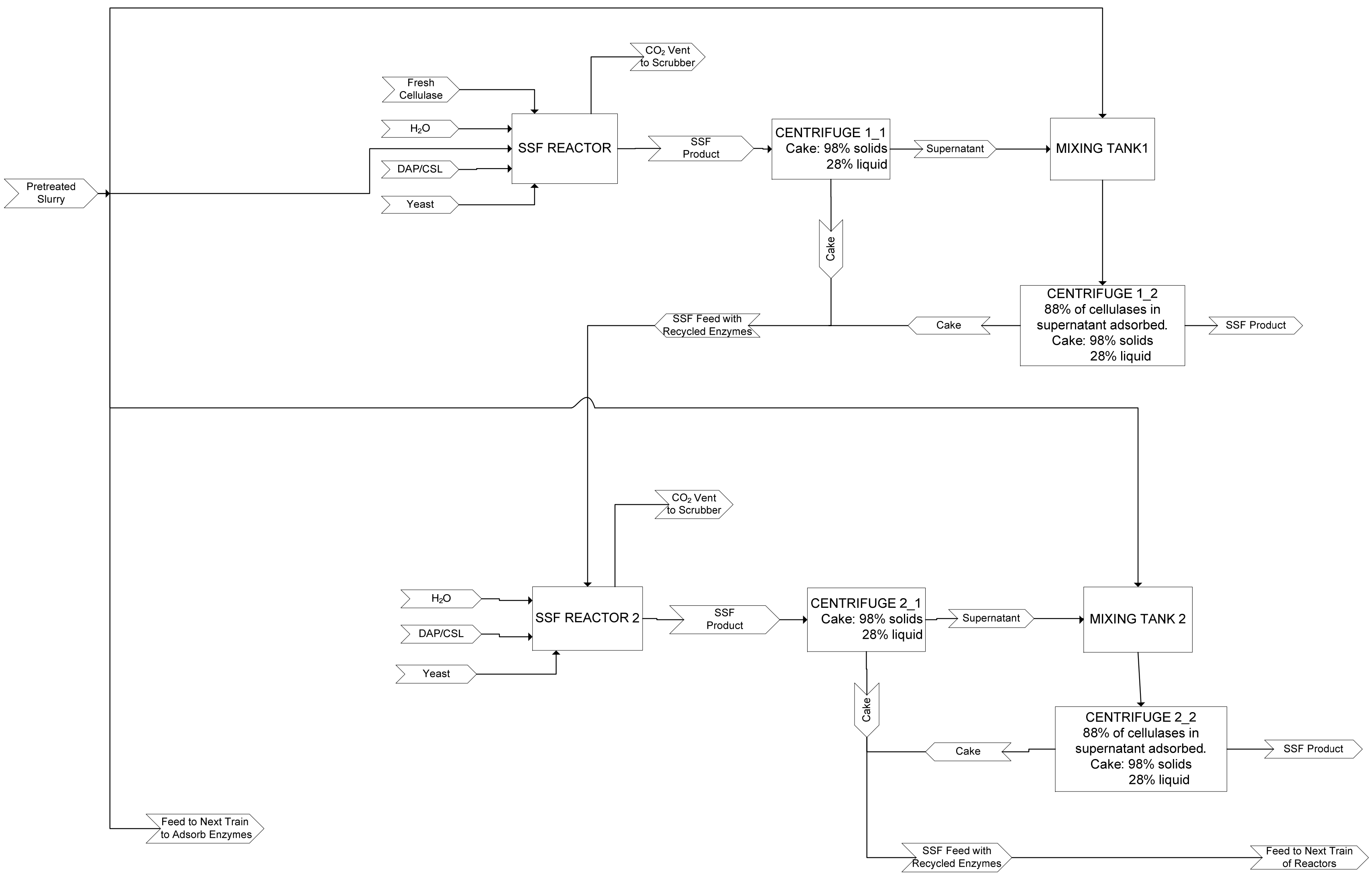

Figure 4.3.1: Recycle Scheme for Parallel Trains of Reactors

4.4 Economics

The change in the value of Minimum Ethanol Selling Price (MESP) over the case with no recycling of enzymes was used to evaluate how cost effective a given recycle scheme is. Only the SSF section with its associated variable operating costs and capital investment changes across the different recycling scenarios.

For each case, the installed equipment costs were calculated using Aspen Icarus Process Evaluator and data from NREL reports (Wooley, Ruth et al.,1999) and (Aden, Ruth et al.,2002). The capital investment was then computed using factors listed in Peters and Timmerhaus, 2002. The base year was chosen to be 2002. The variable operating costs were computed by taking the flow rates of the input raw materials needed for each case from the Aspen Plus simulations. The costs used for the raw materials are listed in Appendix E.

4.4.1 Cash Flow Analysis

The whole ethanol plant is assumed to have a lifetime of 20 years and a construction period of three years. The Modified Accelerated Cost Recovery System (MACRS) is used to calculate the depreciation with a recovery period of seven years. Double declining balance (DDB) depreciation is used in the first four years while the straight line (SL) depreciation method is used in the next three years. The tax rate is assumed to be 39 percent and the MESP is calculated such that the Net Present Value (NPV) of the plant is zero with an Internal Rate of Return (IRR) of 10 percent.

It is assumed that all areas of the plant other than the SSF section add the same fixed value to the MESP across the different scenarios as discussed in Section 5.1. Only the contribution of the SSF section to the MESP changes with the scenarios. This value is calculated by performing discounted cash flow analysis only for the SSF section taking into account the equipment costs and the operating costs. The additional revenue to be generated by ethanol sales to give an NPV of zero is found. As the annual ethanol production is known, the selling price of ethanol required to generate this revenue is calculated. This is the contribution of the SSF section to the MESP. The difference between this value for a given scenario and that for the case with no recycling of enzymes is then used as an approximate indicator to assess the viability of the scenario. The cash flow summaries for the different scenarios are presented in Appendix F.

For each activity loss model, the percentage of the cake exiting the first centrifuge to be purged is a degree of freedom. The cash flow analysis is performed for several values of the purge percentage and the value that results in the maximum reduction in MESP is used for that model.

4.4.2 Sensitivity to Cellulase Cost

The cash flow analysis was repeated for all the scenarios for 10 percent and 20 percent increase and decrease over the current cellulase cost. The variable operating costs change with changes in cellulase cost and as a result the MESP also changes.

CHAPTER 5

RESULTS AND DISCUSSION

5.1 Reconciling MESP

The minimum ethanol selling price (MESP) values are used to compare the different delignification processes and enzyme recycling scenarios. For the delignification processes discussed in Chapter 3, rigorous cash flow analysis was performed for the pretreatment section (A200) and the SSF section (A300) to calculate the value these sections add to the overall plant MESP. This value added by the pretreatment and SSF sections discussed in Chapter 3 is referred to as MESP3. Cash flow analysis was performed for only the SSF section (A300) to compare the enzyme recycling scenarios discussed in Chapter 4. The value added by this section to the overall plant MESP is referred to as MESP4. Benchmark studies performed by NREL for producing ethanol from corn stover (Aden, Ruth et al.,2002) and yellow poplar (Wooley, Ruth et al.,1999) were used to calculate the operating costs and the capital investment required for the feed preparation section (A100), distillation and product recovery section (A500), waste water treatment section (A600), storage and utilities. Table 5.1.1 and Table 5.1.2 list the value to be added to MESP3 and MESP4 respectively for the four delignification processes.

Table 5.1.1: Costs for Sections other than Pretreatment and SSF

	Value Added to MESP3	Operating Costs		Total Project Investment
	$/gal	MM$/yr	cents/gal	MM$/Yr
Base Case	0.53	7.35	32.9	26
Case 2	0.54	7.38	33.1	27.4
Case 3	0.58	7.41	35.2	27.9
Case 4	0.70	7.75	42.2	29.5

Table 5.1.2: Costs for Sections other than SSF

	Value added to MESP4 ($/gal)		Operating Costs		Total Project Investment
	Without Enzyme Recycling	With Enzyme Recycling	MM$/yr	cents/gal	MM$/Yr
Base Case	1.39	1.23	24.7	110	34.3
Case 2	1.32	1.17	22.8	102	37.6
Case 3	1.34	1.19	20.9	99.2	40.8
Case 4	1.41	1.26	18.9	103	39.4

The values listed in Table 5.1.1, when added to the MESP3 values obtained for the pretreatment and SSF sections of the delignification processes discussed in Chapter 3, give the MESP value required for the whole ethanol production plant. For the enzyme recycling scenarios discussed in Chapter 4, the values shown in Table 5.1.2 are added to the MESP4 values to give the overall plant's MESP value. The ethanol production is higher with enzyme recycling as the overall cellulose mass entering the SSF reactor is greater than in the case without recycle. This reduces the overall MESP as shown in Table 5.1.2.

5.2 Delignification Processes

5.2.1 Comparison

The base case involved delignification using dilute NaOH with no recovery of furfural or methanol, produced during steam hydrolysis of the wood, as byproducts. Further processing required for recovering the furfural and methanol as byproducts in case 2 and case 3 increased the equipment and operating costs. In case 4, the lignin is removed using ethanol-water instead of NaOH. Table 5.2.1.1 gives the total installed equipment costs of the four cases. The detailed cost and design data for the equipment used is listed in Appendix D. These costs calculated here are for the pretreatment with delignification section and the saccharification and fermentation section.

Table 5.2.1.1: Installed Equipment Costs for Pretreatment, SSF

Pretreatment Method	**Cost**
NaOH delignification	
Base Case	$13.3 MM
Case 2	$ 15.2 MM
Case 3	$ 17.2 MM
Organosolv Pretreatment	
Case 4	$14.8 MM

The input raw material flow rates, the utilities consumed and the byproducts produced in each of the four cases are listed in Table 5.2.1.2. The associated operating costs are given in Table 5.2.1.3 in million dollars per year and in cents per gallon of ethanol produced. Lignin is recovered as a byproduct in all four cases but only in case 2 and case 3 are furfural and methanol recovered. Since xylose and arabinose are made to undergo further degradation in presence of an acid catalyst to produce furfural in case 3, the recovery of furfural to be sold as byproduct is higher here. But more NaOH and H_2SO_4 is consumed in this case for this production of furfural. The recovery of methanol and furfural in case 2 and case 3 requires higher utilities usage for the stripping and recovery columns.

Table 5.2.1.2: Raw Material Consumption and Byproduct Production

	Units	**Base Case**	**Case 2**	**Case 3**	**Case 4**
Raw Materials					
Hardwood chips	ODT/day	1000	1000	1000	1000
NaOH	tons/day	30	30	44	
H_2SO_4	tons/day	37	37	39	
Corn Steep Liquor	lb/hr	1680	1680	1650	1675
Purchased Cellulase Enzyme	lb/hr	12100	12100	12100	12100
Diammonium Phosphate	lb/hr	211	211	207	211
Zymo Yeast	lb/hr	330	315	315	341
Ethanol Make Up	tons/day				0.05
Utilities					
25 psig Steam	1000 lbs/day		10	39	
150 psig Steam	1000 lbs/day		198	939	
300 psig Steam	1000 lbs/day	1320	1320	1320	
600 psig Steam	1000 lbs/day			149	785
Cooling Water	1000 gal/day	3470	4070	6320	2460
Process Water	1000 gal/day	596	596	597	749
Electricity	KW	3730	3810	3950	6250
Byproducts					
Lignin	ODT/day	277	277	277	393
Furfural	tons/day		7	22	
Methanol	tons/day		2	2	

Table 5.2.1.3: Variable Operating Costs

	MM$/yr (2002)			
Raw Material	**Base Case**	**Case 2**	**Case 3**	**Case 4**
Hardwood Chips	19.3	19.3	19.3	19.3
NaOH	2.80	2.80	4.12	0
H_2SO_4	0.84	0.84	0.90	0
25 psig Steam	0	0.01	0.02	0
150 psig Steam	0	0.12	0.58	0
300 psig Steam	0.89	0.89	0.89	0
600 psig Steam	0	0.00	0.12	0.62
Cooling Water	0.62	0.73	1.13	0.44
Process Water	0.24	0.24	0.24	0.30
Electricity	1.72	1.76	1.82	2.89
Lignin	-8.06	-8.06	-8.06	-11.40
Furfural	0.00	-2.08	-6.36	0
Methanol	0.00	-0.23	-0.23	0
Corn Steep Liquor	1.12	1.12	1.10	1.12
Purchased Cellulase Enzyme	5.52	5.52	5.52	5.52
Diammonium Phosphate	0.12	0.12	0.12	0.12
Zymo	0.46	0.44	0.44	0.47
Ethanol Make Up	0	0	0	0.002
Total Variable Operating Costs	**25.5**	**23.5**	**21.6**	**19.3**
	Cents/Gallon Ethanol (2002)			
Raw Material	**Base Case**	**Case 2**	**Case 3**	**Case 4**
Hardwood chips	86.3	86.3	91.6	105
NaOH	12.5	12.5	19.6	0
H2SO4	3.78	3.78	4.30	0
25 psig Steam	0	0.02	0.10	0
150 psig Steam	0	0.54	2.74	0
300 psig Steam	3.97	3.97	4.21	0
600 psig Steam	0	0	0.56	3.35
Cooling Water	2.78	3.26	5.38	2.40
Process Water	1.08	1.08	1.15	1.65
Electricity	7.72	7.88	8.67	15.7
Lignin	-36.1	-36.1	-38.3	-62.3
Furfural	0	-9.31	-30.2	0
Methanol	0.00	-0.74	-1.11	0
Corn Steep Liquor	5.01	5.01	5.22	6.08
Purchased Cellulase Enzyme	24.7	24.7	26.2	30.1
Diammonium Phosphate	0.56	0.56	0.58	0.68
Zymo	2.05	1.95	2.07	2.58
Ethanol Make Up	0	0	0	0.02
Total Variable Operating Costs	**114**	**105**	**102**	**105**

The costs listed in Table 5.2.1.3 are calculated as a percentage of the total cost for the four cases as shown in Figure 5.2.1.1. The woodchips constitute the largest portion of the operating costs, in all four cases, followed by the enzymes. The sale of byproducts, especially furfural for case 2 and case 3 and lignin, reduce the operating costs significantly.

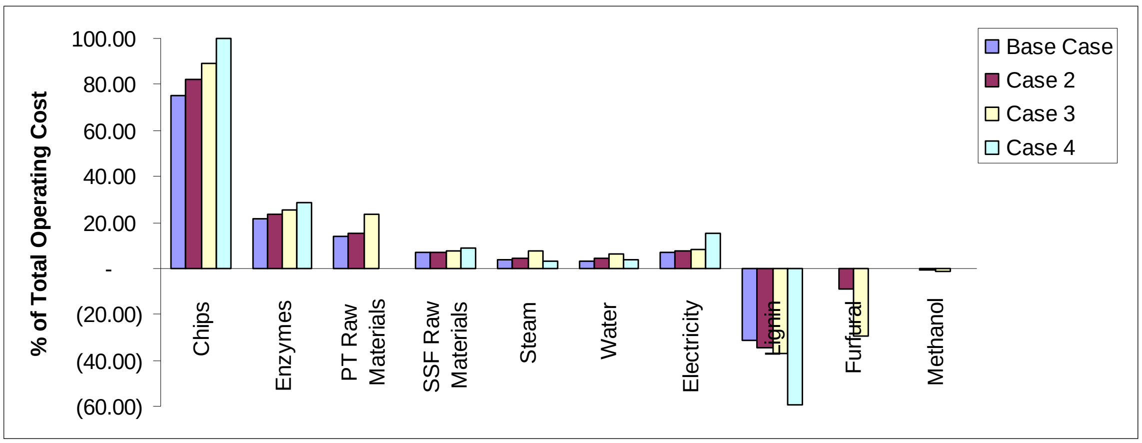

Figure 5.2.1.1: Comparison of Raw Material Costs as a Percentage of Total Operating Cost

A discounted cash flow analysis was performed for the four cases over an operating period of 20 years and the Minimum Ethanol Selling Price (MESP3) was calculated such that the Net Present Value (NPV) was zero with an Internal Rate of Return (IRR) of 10 percent. The MESP3 values for the four cases are shown in Table 5.2.1.4, along with the ethanol produced, capital costs and the operating costs. Indirect costs, to account for contingency, are added to the installed costs to get the total project investment needed. Operating costs are given in terms of millions of dollars required per year and also in terms of the cost in cents for every gallon of ethanol produced.

Table 5.2.1.4: Comparison of Economics for Pretreatment, SSF

	Base Case	**Case 2**	**Case 3**	**Case 4**
MESP3	$1.26	$1.19	$1.18	$1.21
Ethanol Production (MM Gal. / Year)	22.3	22.3	21.0	18.4
<u>Capital Costs</u>				
Total Installed Equipment Cost	$13,300,000	$15,170,000	$17,160,000	$14,840,000
Indirect Costs	$400,000	$450,000	$520,000	$450,000
Total Project Investment	$13,700,000	$15,620,000	$17,680,000	$15,290,000
<u>Operating Costs (cents/gal)</u>				
Raw Materials	135	135	150	144
Utilities	15.6	16.8	22.8	23.1
Sale of Byproducts	-36.1	-46.2	-69.7	-62.3
Capital Depreciation	3.0	3.4	4.1	4.0
<u>Operating Costs (MM$/yr)</u>				
Raw Materials	$30,120,000	$30,120,000	$31,470,000	$26,500,000
Utilities	$3,470,000	$3,740,000	$4,800,000	$4,250,000
Sale of Byproducts	-$8,060,000	-$10,310,000	-$14,660,000	-$11,440,000
Capital Depreciation	$670,000	$760,000	$860,000	$740,000

Table 5.2.1.5: Comparison of Overall Economics

	Overall MESP	Operating Costs		Total Project Investment
	$/gal	Cents/gal	MM$/yr	MM$/Yr
Base Case	1.79	147	33	40
Case 2	1.73	138	31	43
Case 3	1.76	138	29.	46
Case 4	1.91	147	27.1	45

The Base case has the lowest capital costs, raw material and utilities costs and the highest production of ethanol as can be seen in Table 5.2.1.4. But this is compensated for by the sale of furfural and methanol in Cases 2 and 3 and by increased lignin recovery in Case 4.

This is reflected in sale of byproducts in Table 5.2.1.4. Case 2 and Case 3 are the most viable, but this viability is highly sensitive to the selling price of furfural. Case 4 loses its viability due to the decreased production of ethanol. This lower production is due to the loss of soluble sugars in stream 208 when the pretreated slurry is washed to remove the dissolved lignin. About 20% of the cellulose is hydrolyzed along with the hemicellulose. These sugars are recovered by evaporation but over half are oligomers, which cannot be used for ethanol production as they are washed along with the dissolved lignin and then sent to distillation section for ethanol recovery.

Table 5.2.1.5 gives the economics with all the sections of the ethanol production plant for the four processes. The Case 2 with furfural and methanol recovery from the pretreatment section is the most viable. Case 3 and Case 4 have lower annual production of ethanol and this increases the minimum ethanol selling price.

5.2.2 Sensitivity to Furfural Cost

Higher investment is required for case 2 and case 3 to recover the methanol and furfural produced. When the furfural price falls, it is not economical to put in the additional investment required to recover these byproducts. The furfural cost is varied to observe how the economic viability of case 2 and case 3 changes. The MESP3 values are calculated for the cases 2 and 3 for different selling price of furfural. The base case MESP value is then subtracted from the MESP values for these two cases to see how the viability of the three cases over base case changes as shown in Figure 5.2.2.1 and Table 5.2.2.1. The price is changed to 10% and 20% above and below the current furfural price to observe the trend. The extra processing steps required to convert the xylose to furfural is viable only when the furfural price is more than \$952/ton. The base case becomes more viable only when the furfural price falls to \$259/ton, a 68% drop from current price.

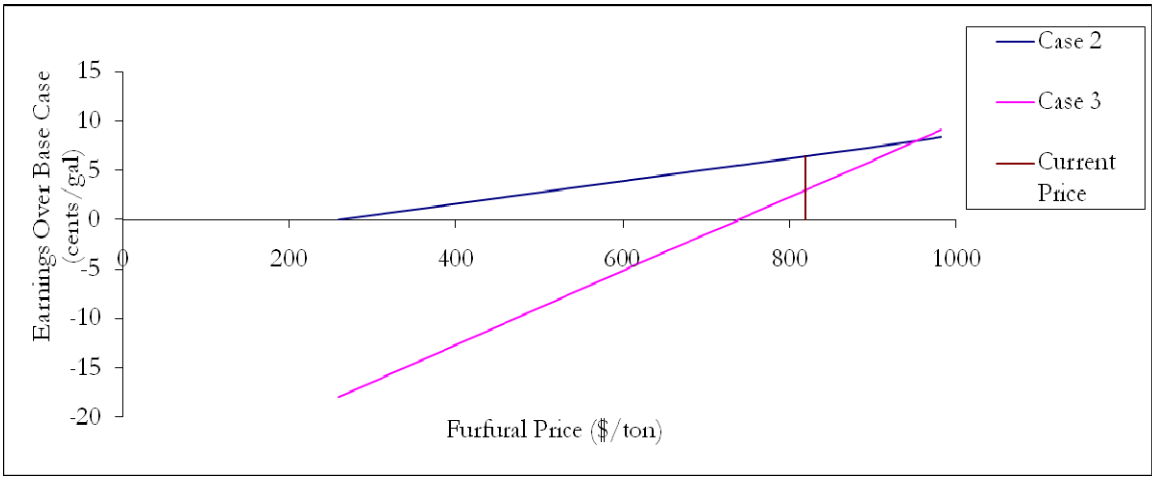

Figure 5.2.2.1: Sensitivity of MESP of Case 2 and Case 3 to Furfural Price

Table 5.2.2.1: Difference in MESP over Base Case for Varying Furfural Price

	Price of Furfural	Case 2	Case 3
	\$/ton	cents/gal	cents/gal
Price Below which Case 2 is Not Economically Viable	259	0	-18
20% Decrease in Price over Current Price	655	4.6	-3.2
10% Decrease in Price over Current Price	736	5.5	-0.1
Price Below which Case 3 is Not Economically Viable	738	5.5	0
Current Price	818	6.5	3.0
10% Increase in Price over Current Price	900	7.4	6.1
Price at which Cases 2 and 3 are Equally Viable	952	8.0	8.0
20% Increase in Price over Current Price	982	8.4	9.1

5.3 Enzyme Recycling

Once the activity loss scheme is identified, it is possible to minimize the cost by optimizing the number of parallel trains or the amount purged. The recycling of enzymes can result in savings upto 25¢ per gallon of ethanol. Figure 5.3.1 shows the reduction in average contribution of enzyme costs to MESP4 from the scenario with no recycle.

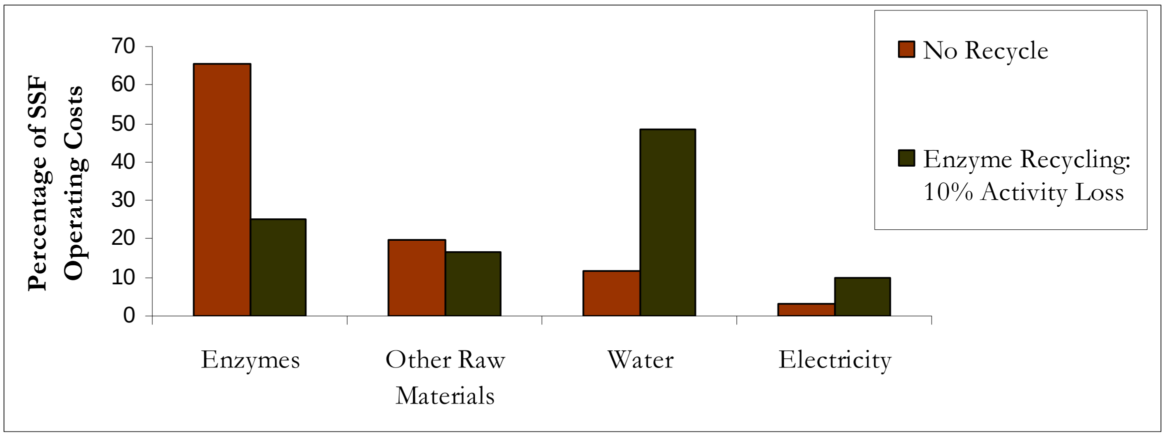

Figure 5.3.1: Average Cost Contribution of Operating Costs to MESP4

5.3.1 Comparison of Scenarios

The single train of reactors is more cost effective than the parallel trains of reactors when less than 10 percent of the activity of the cellulase is lost for each pass through the reactor as can be seen in Figure 5.3.1.1. The parallel reactors scheme represents more efficient usage of the raw materials but involves higher utilities' cost. When the activity loss is high, the cost of raw materials represents a more significant percentage of the product cost making the parallel scheme more advantageous. The SSF reactors constitute the major portion of the equipment costs. The number of reactors as well as the size of the reactors is similar for all the schemes and this is the reason the parallel reactors scheme doesn't involve higher equipment costs. Also the economies of scale are lost at this operating size and the equipment scales linearly with flowrate for the single train of reactors scheme. As expected, the multi-step and the single step deactivation schemes show similar trends as they are just different representations of the cellulase deactivation with everything else being the same. Table 5.3.1.2 shows the economics of the three schemes for different losses in enzyme activity.

Table 5.3.1.1: Comparison of the Different Scenarios

	Savings (cents/gal)		
	Single Train		Parallel Trains
% Activity Loss	Multi Step	Single Step	Single Step
0	25	26	22
5	24	24	22
10	23	22	22
15	21	21	21
20	20	19	20
25	19	18	19

Savings = MESP for the recycle scheme – MESP without recycle

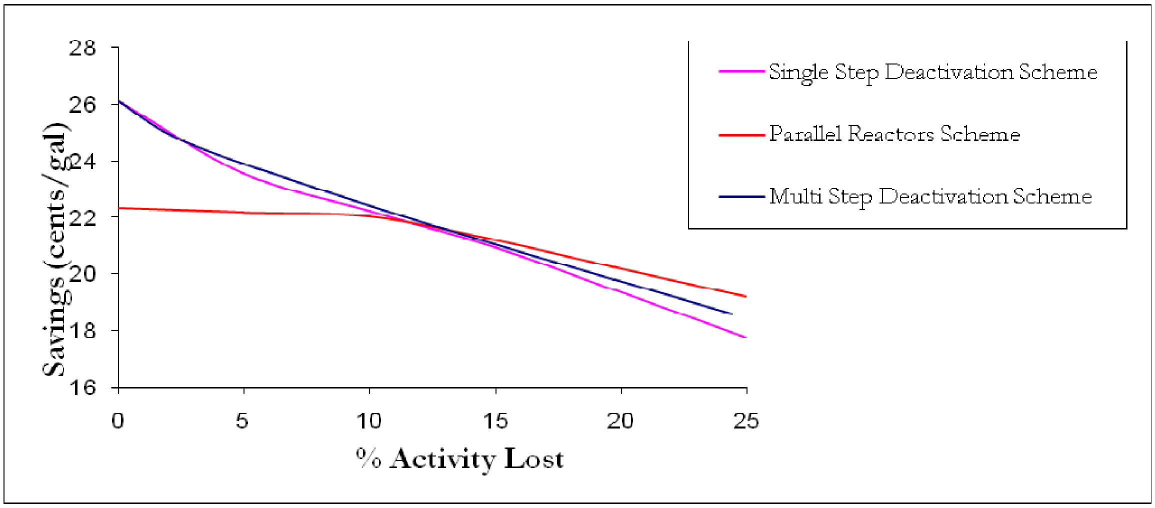

Figure 5.3.1.1: Comparison of the Different Scenarios

Table 5.3.1.2: Cellulase savings over No Recycle

	Cellulase Savings*	
% Activity Loss	Single Step	Parallel Reactors
0	74%	74%
5	70%	72%
10	71%	72%
15	67%	70%
25	55%	66%

* Cellulase Savings = Percentage decrease in cellulase cost per gallon ethanol over no recycle case.

Table 5.3.1.2: Material Balance and Economics Summary of the Recycle Schemes

Activity Loss		0%			5%		
	No Recycle	Multi-Step	Single-Step	Parallel	Multi-Step	Single-Step	Parallel
Purge %	100	25	25	0	20	25	0
No. of Trains	1	1	1	4	1	1	4
Total Project Investment ($/yr)	5,360,000	14,790,000	14,790,000	11,490,000	13,060,000	15,590,000	11,370,000
Operating Costs ($/yr)							
CSL	1,120,000	900,000	900,000	910,000	810,000	950,000	900,000
Cellulase	5,520,000	1,540,000	1,540,000	1,600,000	1,690,000	1,780,000	1,710,000
DAP	120,000	100,000	100,000	100,000	90,000	110,000	100,000
Process Water	260,000	450,000	450,000	450,000	350,000	500,000	430,000
Zymo	1,850,000	0	0	130,000	0	0	140,000
Cooling Water	440,000	300,000	300,000	110,000	240,000	320,000	110,000
Chilled Water	0	710,000	710,000	490,000	610,000	760,000	480,000
Electricity	340,000	1,020,000	1,020,000	2,930,000	860,000	1,100,000	2,880,000
Capital Depreciation	260,000	740,000	740,000	560,000	650,000	780,000	550,000
Ethanol Production (MM Gal. / Yr)	22.5	24.5	24.5	25	23.3	24.5	25.1
MESP4 ($)	0.41	0.31	0.31	0.35	0.34	0.33	0.35
Change in MESP Over No Recycle	0	0.26	0.26	0.22	0.24	0.24	0.22

Table 5.3.1.2 (Continued)

Activity Loss		10%		15%		
	No Recycle	Single-Step	Parallel	Multi-Step	Single-Step	Parallel
Purge %	100	20	0	25	20	0
No. of Trains	1	1	5	1	1	4
Total Project Investment ($/yr)	5,360,000	16,470,000	11,510,000	15,910,000	16,600,000	11,810,000
Operating Costs ($/yr)						
CSL	1,120,000	1,030,000	910,000	970,000	1,040,000	940,000
Cellulase	5,520,000	1,750,000	1,770,000	2,140,000	2,000,000	1,930,000
DAP	120,000	120,000	100,000	110,000	120,000	110,000
Process Water	260,000	530,000	440,000	500,000	530,000	450,000
Zymo	1,850,000	0	150,000	0	0	160,000
Cooling Water	440,000	350,000	110,000	330,000	350,000	110,000
Chilled Water	0	810,000	480,000	780,000	820,000	500,000
Electricity	340,000	1,190,000	2,930,000	1,130,000	1,200,000	3,050,000
Capital Depreciation	260,000	820,000	560,000	800,000	830,000	570,000
Ethanol Production (MM Gal. / Yr)	22.5	24.7	25	24.5	24.7	25.3
MESP4 ($)	0.41	0.35	0.35	0.36	0.36	0.36
Change in MESP Over No Recycle	0	0.22	0.22	0.21	0.21	0.21

Table 5.3.1.2 (Continued)

Activity Loss		25%		
	No Recycle	Multi-Step	Single-Step	Parallel
Purge %	100	25	25	0
No. of Trains	1	1	1	4
Total Project Investment ($/yr)	5,360,000	13,060,000	16,560,000	12,550,000
Operating Costs ($/yr)				
CSL	1,120,000	810,000	1,020,000	1,010,000
Cellulase	5,520,000	3,530,000	2,710,000	2,260,000
DAP	120,000	90,000	120,000	110,000
Process Water	260,000	350,000	530,000	480,000
Zymo	1,850,000	0	0	190,000
Cooling Water	440,000	240,000	350,000	120,000
Chilled Water	0	610,000	820,000	550,000
Electricity	340,000	860,000	1,200,000	3,360,000
Capital Depreciation	260,000	650,000	830,000	610,000
Ethanol Production (MM Gal. / Yr)	22.5	23.3	24.5	25.6
MESP4 ($)	0.41	0.38	0.39	0.38
Change in MESP Over No Recycle	0	0.19	0.18	0.19

5.3.2 Percentage Purge

The amount purged affects the flow rate into the SSF reactor and hence the reactor size. The reactors form a major portion of the equipment costs and small changes in their size has a significant impact on the savings. So, smaller purges imply higher flow rates through the reactor but lower loss of cellulase in the purge stream. There is an optimum purge percentage which balances out these two factors. For high activity loss scenarios, the percentage purged is higher as the viability of recycling is reduced.

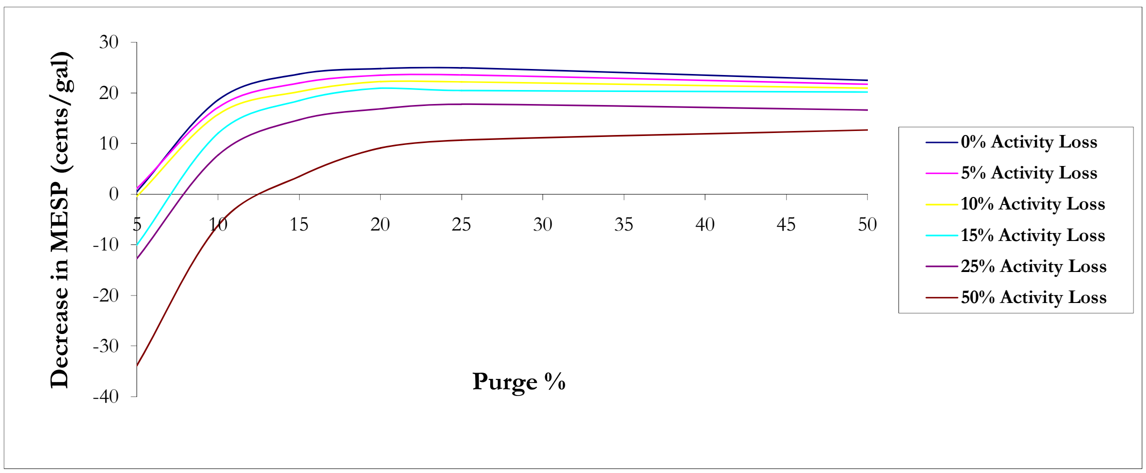

Figure 5.3.2.1: Single Step Deactivation Model- Savings vs. Percentage Purge

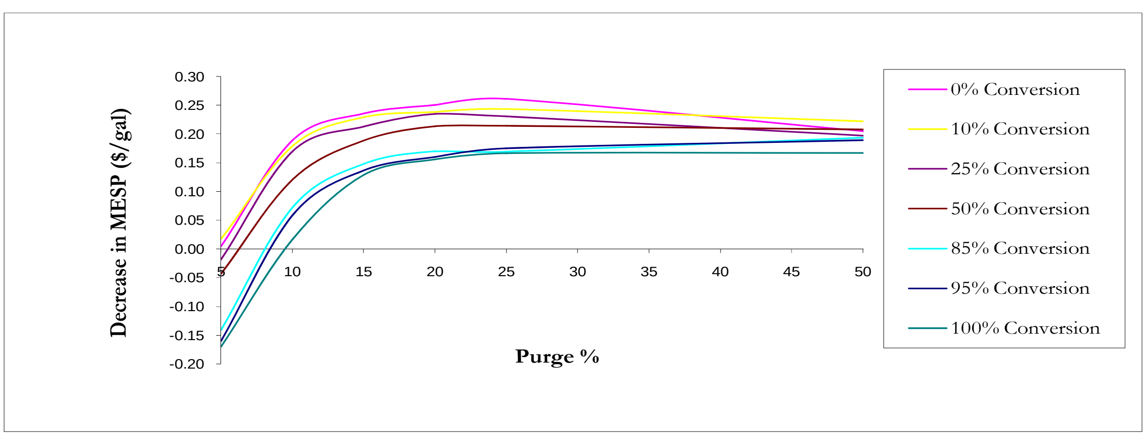

Figure 5.3.2.2: Multiple Step Deactivation Model- Savings vs. Percentage Purge

5.3.3 Number of Trains

More parallel trains imply the cellulase is used more often and the fresh cellulase to be added to the first train is lower. But economy of scale is lost and equipment costs increase. The optimal number of trains is a trade off between the decrease in cellulase cost and the increase in equipment costs with more number of trains. When the activity loss is too high, the cellulase cannot be reused often and the optimal number of parallel trains decreases.

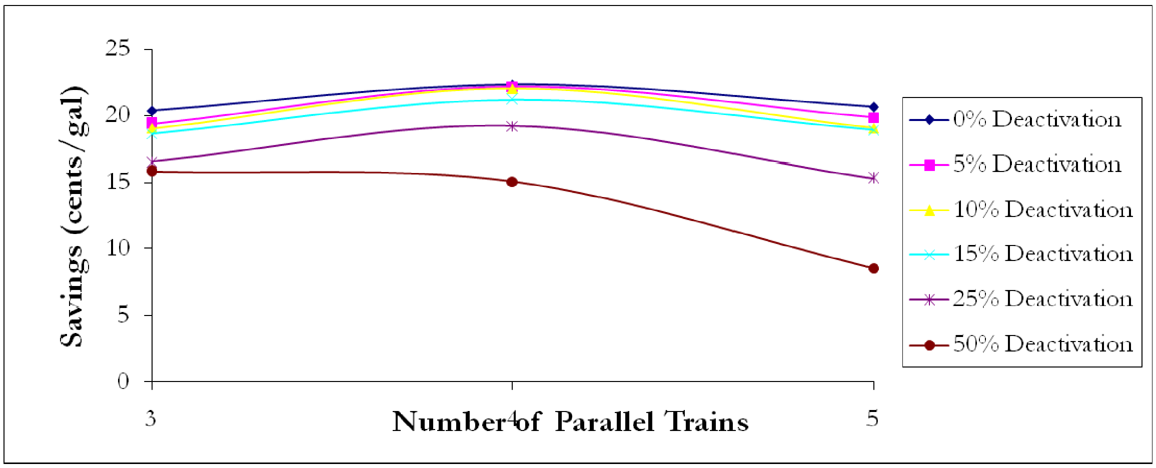

Figure 5.3.3.1: Parallel Reactors Scheme- Savings vs. Number of Parallel Trains

5.3.4 Sensitivity to Cellulase Cost

The change in cost of cellulase shifts the balance between the effect due to increased equipment costs and the amount of cellulase recycled. For the case of a single train of reactors, the cost of the cellulase affects the impact of the cellulase lost in the purge stream since the make up cellulase to be added depends on the amount purged. In the case of parallel reactors, the cellulase cost affects the number of times the cellulase is to be used. With increasing cellulase costs, the parallel train of reactors scheme becomes more viable than the single train of reactors scheme at lower enzyme activity loss and vice versa. Figures 5.3.4.1 to 5.3.4.3 show the parallel movement of the savings vs activity loss curve with changing cellulase prices.

Table 5.3.4.1: Single Step Deactivation Scheme - Sensitivity to Cellulase Cost

Single Step Deactivation Scheme										
Cellulase Cost	20% Decrease		10% Decrease		Constant Price		10% Increase		20% Increase	
% Activity Loss	Purge %	Savings ($)	Purge %	Savings ($)	Purge %	Savings ($)	Purge %	Savings ($)	Purge %	Savings ($)
0	25	0.23	25	0.24	25	0.26	20	0.28	20	0.30
5	25	0.20	25	0.22	25	0.24	20	0.25	20	0.27
10	50	0.19	25	0.21	20	0.22	20	0.24	25	0.25
15	50	0.18	20	0.19	20	0.21	20	0.23	20	0.24
25	25	0.16	25	0.16	25	0.18	25	0.19	25	0.20
50	50	0.16	50	0.16	50	0.13	50	0.16	50	0.16

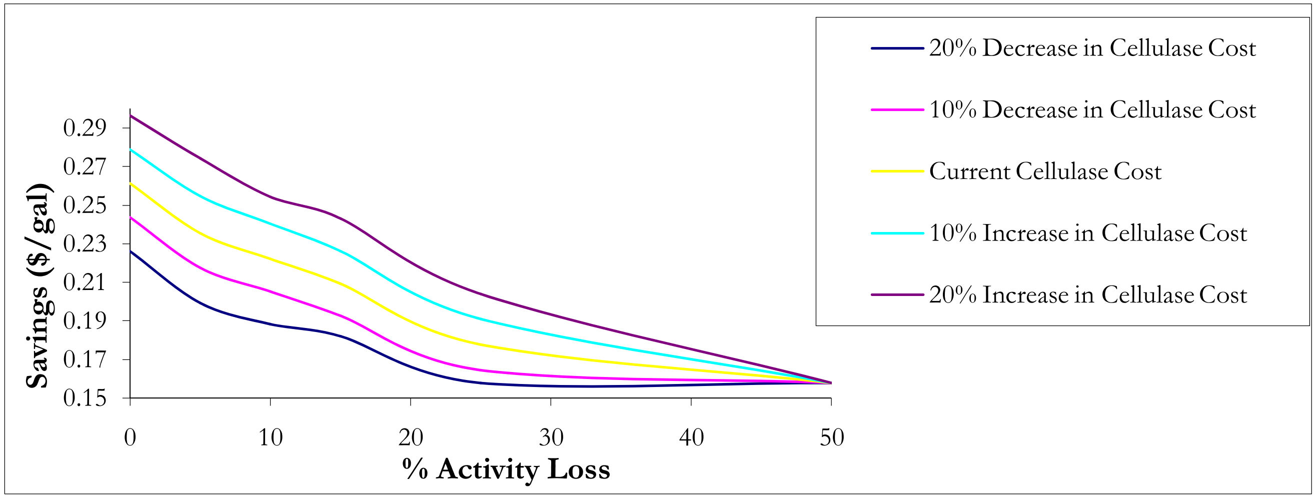

Figure 5.3.4.1: Single Step Deactivation Scheme - Sensitivity to Cellulase Cost

Table 5.3.4.2: Multiple Step Deactivation Scheme - Sensitivity to Cellulase Cost

Multiple Step Deactivation Scheme											
Cellulase Cost		20% Decrease		10% Decrease		Constant Price		10% Increase		20% Increase	
% Activity Loss	% Conversion	Purge %	Savings ($)	Purge %	Savings ($)	Purge %	Savings ($)	Purge %	Savings ($)	Purge %	Savings ($)
0	0	25	0.23	25	0.24	25	0.26	25	0.28	25	0.30
2	10	25	0.21	25	0.23	25	0.24	25	0.26	25	0.28
6	25	20	0.20	20	0.22	20	0.24	20	0.25	20	0.27
14	50	50	0.19	25	0.20	25	0.21	20	0.23	20	0.24
24	85	50	0.18	50	0.19	50	0.19	50	0.20	50	0.21
28	95	50	0.18	50	0.18	50	0.19	50	0.19	50	0.20
29	100	50	0.16	50	0.16	50	0.17	50	0.17	25	0.18

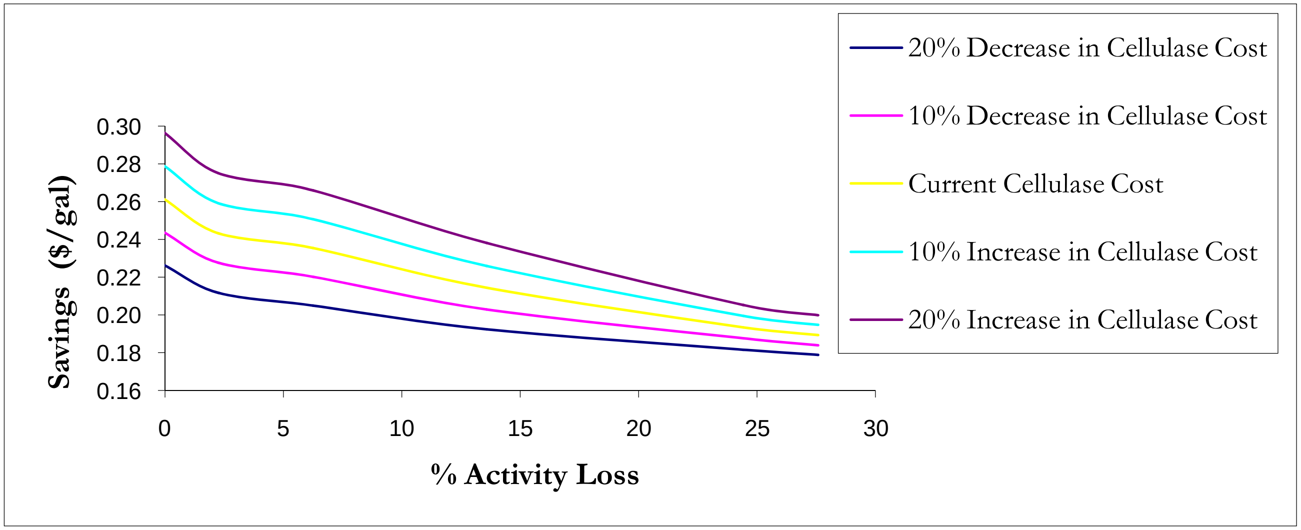

Figure 5.3.4.2: Multiple Step Deactivation Scheme - Sensitivity to Cellulase Cost

Table 5.3.4.3: Parallel Reactors Scheme - Sensitivity to Cellulase Cost

Parallel Reactors Scheme										
Cellulase Cost	20% Decrease		10% Decrease		Constant Price		10% Increase		20% Increase	
% Activity Loss	No. of Trains	Savings ($)	No. of Trains	Savings ($)	No. of Trains	Savings ($)	No. of Trains	Savings ($)	No. of Trains	Savings ($)
0	4	0.18	4	0.20	4	0.22	4	0.24	4	0.27
5	4	0.18	4	0.20	4	0.22	4	0.24	4	0.26
10	4	0.18	4	0.20	4	0.22	4	0.24	4	0.26
15	4	0.17	4	0.19	4	0.21	4	0.23	4	0.25
25	4	0.16	4	0.17	4	0.19	4	0.21	4	0.23
50	3	0.16	3	0.16	3	0.16	3	0.17	3	0.18

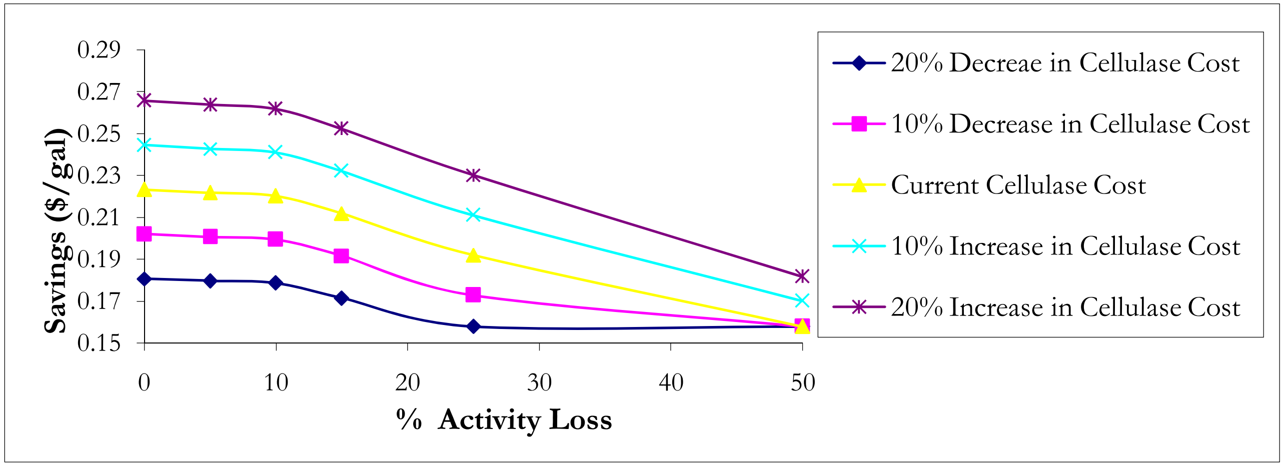

Figure 5.3.4.3: Parallel Reactors Scheme - Sensitivity to Cellulase Cost

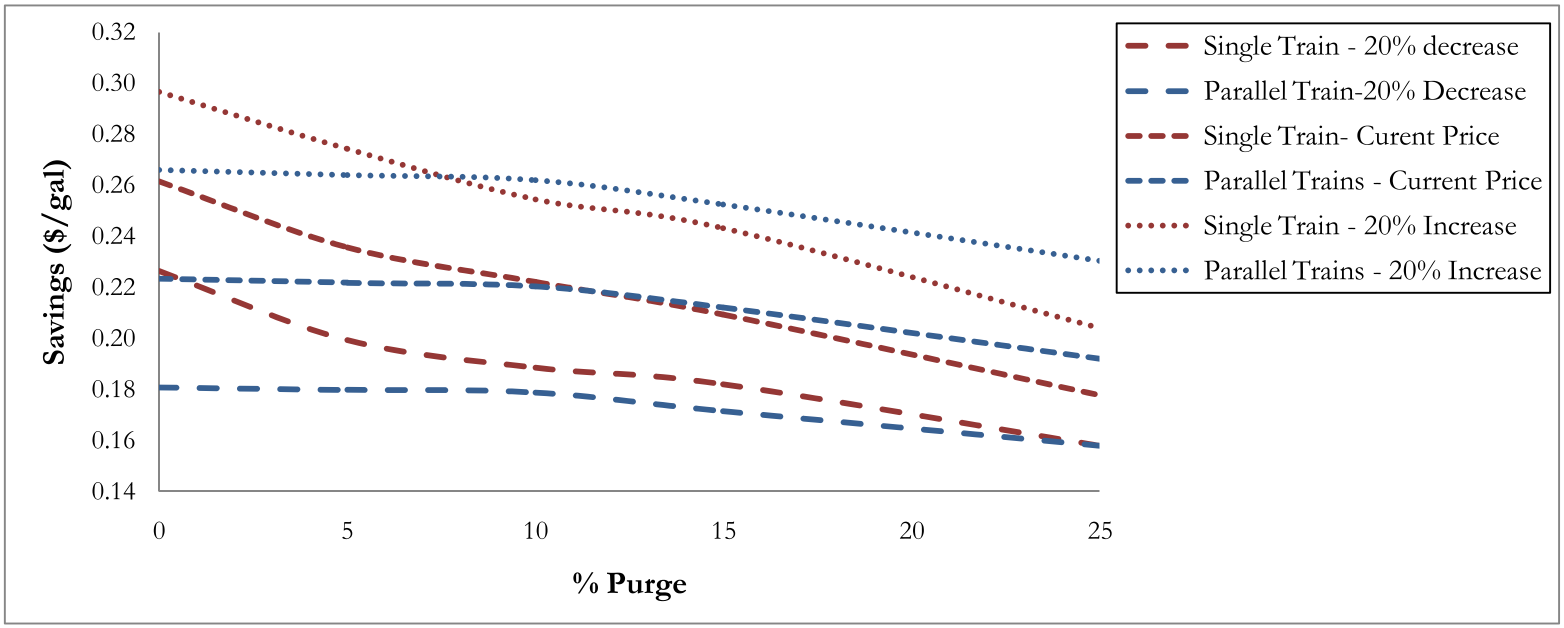

Figure 5.3.4.4: Comparison of Schemes with Change in Cellulase Cost

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 Delignification

1. The recovery of byproducts - methanol and furfural, greatly improves the economics by over 6.5¢/gal. Case 3 is viable when the furfural selling price is high to justify the increased capital costs required to convert xylose to furfural.
2. Organosolv pretreatment gives higher recovery of lignin as a byproduct but the annual production of ethanol is low at 18.4 MM gal/yr compared to 22.33 MM gal/yr produced in the base case.

6.2 Enzyme Recycling

1. The parallel trains of reactors scheme is more economical in situations where cellulase cost is a higher percentage of the overall cost. The cellulase cost contribution to the overall cost is high in cases where larger amounts of fresh cellulase are needed due to higher activity loss and when the cost of cellulase is high.
2. The single train of reactors scheme is more economical when the cellulase activity loss is less than 10%.
3. The multiple-step deactivation scheme and single-step deactivation scheme show similar economics as the only difference is the activity loss model with everything else being the same.

4. The enzyme cost contribution to the SSF operating costs reduces from around 60 percent for the case with no recycle to about 25 percent when the cellulase is recycled while the cost contribution of the utilities and depreciation increases.
5. The parallel trains' scheme involves more efficient usage of raw materials while the single train of reactors involves lower utilities' costs.
6. The parallel reactors scheme is viable as the single train of reactors doesn't have the economies of scale advantage at this operating size and hence the capital costs are comparable. The economics will change if the plant size is reduced.

6.3 Future Work

1. The SSF yields should be verified when the enzymes are recycled as some ethanol enters with the recycle stream and can cause end-product inhibition.
2. For the enzyme recycling case, possible ways to reduce the use of chilled water is needed when the enzymes present in the supernatant (309) are adsorbed by the SSF feed, as this adds a significant amount to the operating costs.
3. The effect of increasing cellulase loading on production of ethanol can be evaluated as the cellulase costs are reduced with recycling.
4. Improve the ethanol production in the Organosolv pretreatment case.

APPENDIX A

SAMPLE CALCULATIONS

Parallel Reactors Scheme: Cellulase Mass Balance

After the hydrolysis of cellulose, the enzymes distribute between the supernatant liquid and the solids. The cellulase adsorbed in the solids is denoted as C_c and that in the supernatant is denoted as C_s while the cellulase initially entering the SSF reactor is denoted as C_1. A mass balance is performed on the supernatant cellulase, C_s and the adsorbed cellulase, C_c as shown in Figure A.1 and in the Table A.1. The total active cellulase recycled, C_2 is calculated for different activity loss values. It is assumed that the adsorbed cellulase is the one which loses its activity. The calculations are repeated by replacing C_1 with C_2 to find C_3, the active cellulase sent to the third train and so on.

Table A.1: Cellulase Mass Balance

SSF	
Total	C1
Solid	Cs
Liquid	Cc=C1-Cs
Centrifuge 1	
Cake	0.98*Cc+0.28*Cs
Solid	0.98Cc
Liquid	0.28Cs
Centrate	0.02Cc+0.72*Cs
Solid	0.02Cc
Liquid	0.72Cs
Centrifuge 2	
Adsorbed	0.88*0.72*Cs+0.02Cc
Supernatant	0.12*(0.72Cs)
Out	
Cake	0.645Cs+0.0196Cc
Solid	0.98*(0.88*0.72Cs+0.02Cc)
Liquid	0.28*0.12*(0.72Cs)
Centrate	0.075Cs+0.0004Cc
Solid	0.02*(0.88*0.72Cs+0.02Cc)
Liquid	0.72*0.12*(0.72Cs)
Recycled Total	
Centrifuge 1 Cake+Centrifuge2_cake	
Solid	0.98Cc+0.98*(0.88*0.72Cs+0.02Cc)
Liq	0.28Cs+0.28*0.12*(0.72Cs)

Table A.2: Cellulase Activity Recycled for Different Loss Models

Activity Retained (95%)	
Solid	0.95*(0.621Cs+0.9996Cc)
Liquid	0.304Cs
Activity Retained (90%)	0.9C1-0.0367Cs
Solid	0.90*(0.621Cs+0.9996Cc)
Liquid	0.304Cs
Activity Retained (85%)	0.85C1-0.0178Cs
Solid	0.85*(0.621Cs+0.9996Cc)
Liquid	0.304Cs
Activity Retained (75%)	0.75C1-0.02Cs
Solid	0.75*(0.621Cs+0.9996Cc)
Liquid	0.304Cs
Activity Retained (50%)	0.5C1-0.1147Cs
Solid	0.50*(0.621Cs+0.9996Cc)
Liquid	0.304Cs
Activity Retained (100%)	1*C1-0.0746Cs
Solid	1*(0.621Cs+0.9996Cc)
Liquid	0.304Cs

Table A.3: Cellulase Entering the Parallel Trains

Cellulase into SSF	0% Activity Loss	5% Activity Loss	10% Activity Loss
C1	C1	C1	C1
C2	1*C1-0.0746Cs	0.95C1-0.056Cs	0.85C1-0.0178Cs
C3	1*C2-0.0746Cs*C2/C1	0.95C2-0.056*Cs/C1*C2	0.85C2-0.0178Cs*C2/C1
C4	1*C3-0.0746Cs*C3/C1	0.95C3-0.056*Cs/C1*C3	0.85C3-0.0178Cs*C3/C1
Cellulase into SSF	15% Activity Loss	25% Activity Loss	50% Activity Loss
C1	C1	C1	C1
C2	0.85C1-0.0178Cs	0.75C1-0.02Cs	0.5C1-0.1147Cs
C3	0.85C2-0.0178Cs*C2/C1	0.75C2-0.02Cs*C2/C1	0.5C2-0.1147Cs*C2/C1
C4	0.85C3-0.0178Cs*C3/C1	0.75C2-0.02Cs*C3/C1	0.5C3-0.1147Cs*C3/C1

Table A.4: Feed Split Depending on Cellulase Distribution between Solid and Supernatant

Activity Loss-5%				
Cellulase into SSF				
Cs/C1	C1	C2	C3	C4
0.10	1.00	0.94	0.89	0.84
0.20	1.00	0.94	0.88	0.83
0.30	1.00	0.93	0.87	0.81
0.40	1.00	0.93	0.86	0.80
0.50	1.00	0.92	0.85	0.78
0.60	1.00	0.92	0.84	0.77
0.70	1.00	0.91	0.83	0.76
0.80	1.00	0.91	0.82	0.74
0.90	1.00	0.90	0.81	0.73
Ratio of Feed Split Among the Trains				
Cs/C1	Train 1	Train 2	Train 3	Train 4
0.10	0.29	0.25	0.24	0.22
0.20	0.30	0.25	0.24	0.22
0.30	0.30	0.25	0.23	0.22
0.40	0.30	0.25	0.23	0.22
0.50	0.30	0.25	0.23	0.21
0.60	0.31	0.25	0.23	0.21
0.70	0.31	0.25	0.23	0.21
0.80	0.31	0.25	0.23	0.21
0.90	0.31	0.25	0.23	0.21
Average Split	0.30	0.25	0.23	0.21

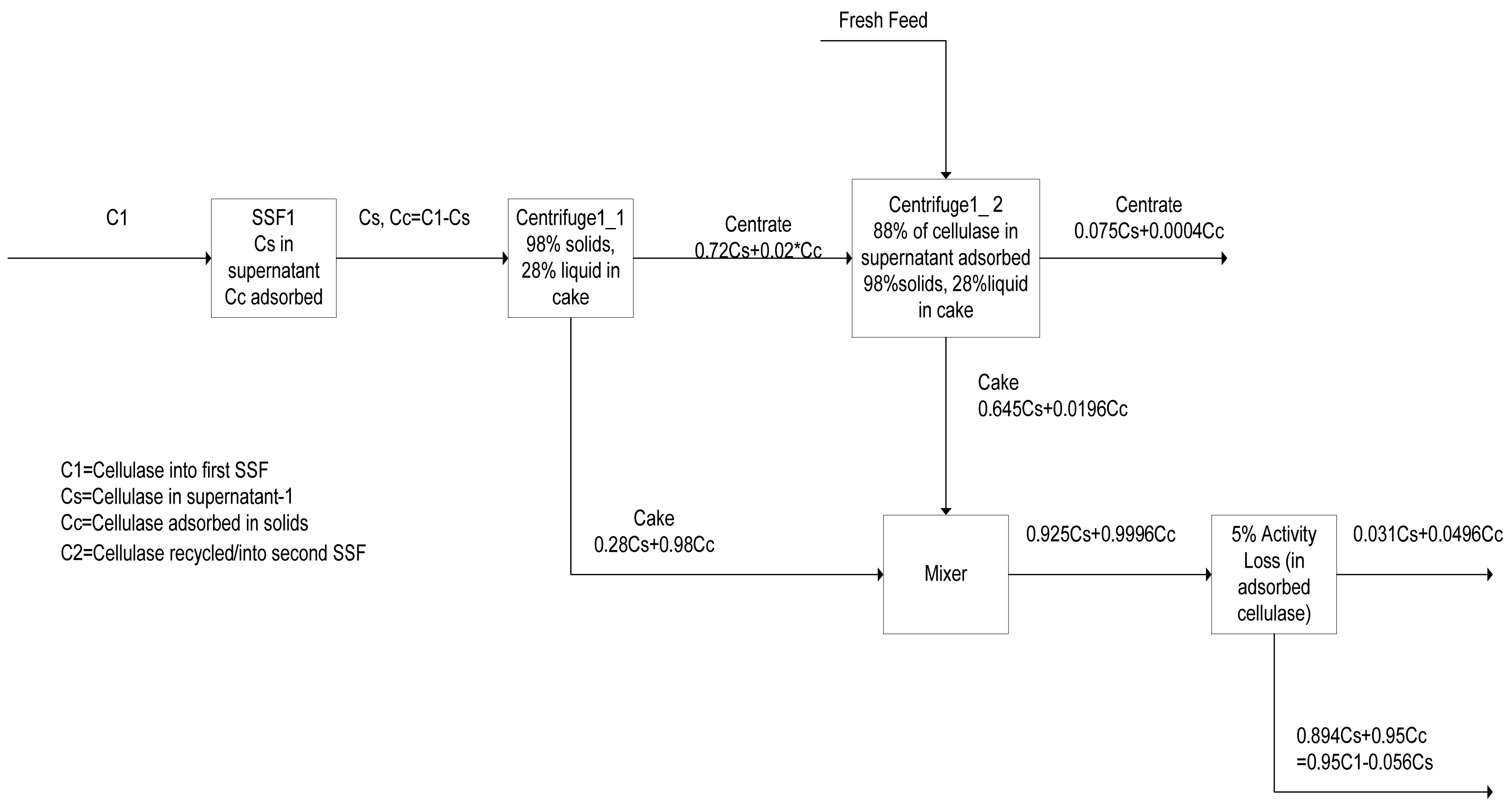

Figure A.1: Cellulase Mass Balance

APPENDIX B

PROCESS FLOW DIAGRAMS

B.1 Process Flow Diagrams: Base Case

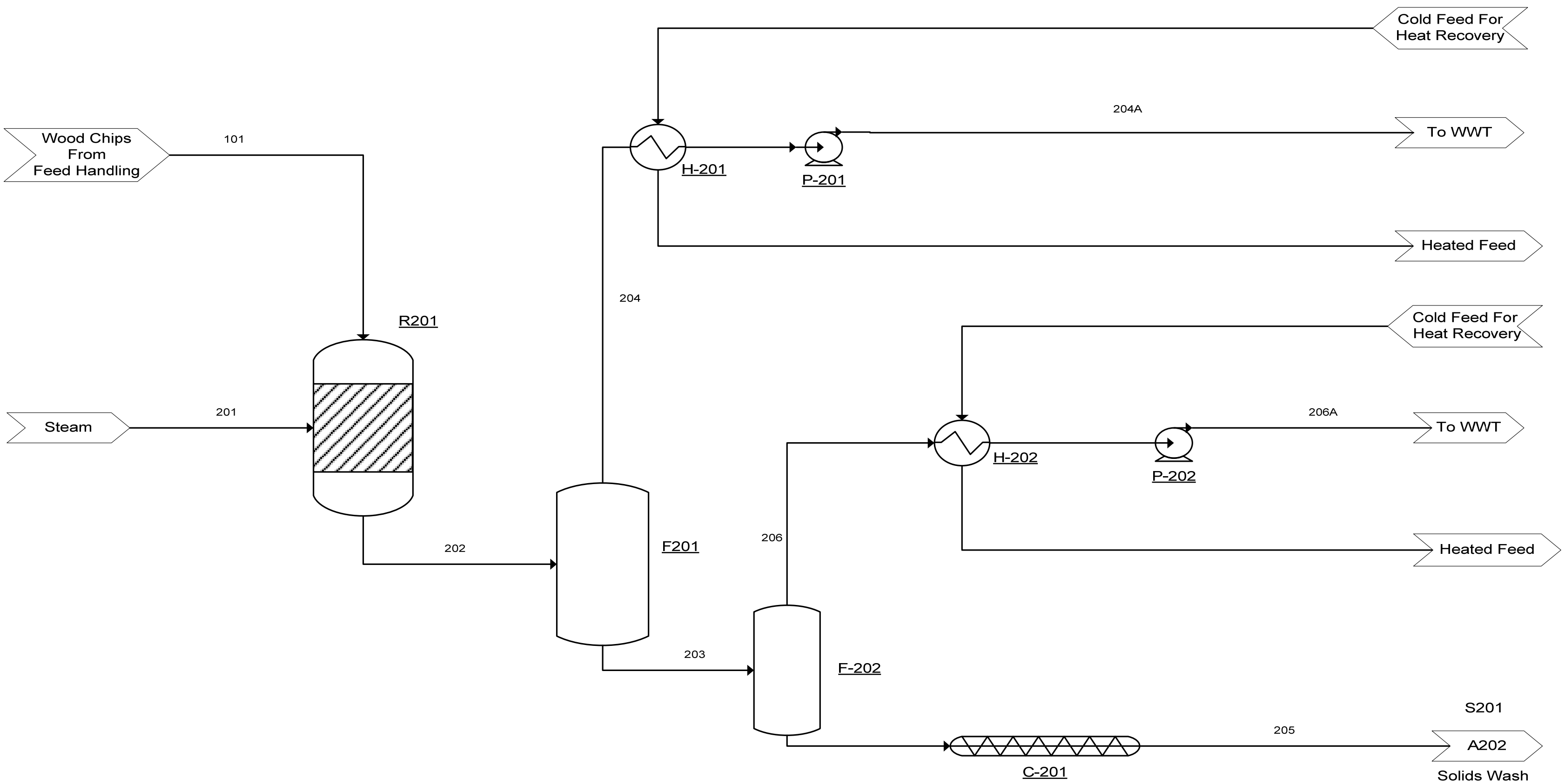

Figure B.1.1: A201- Base Case Pretreatment – Digestor and Flash System

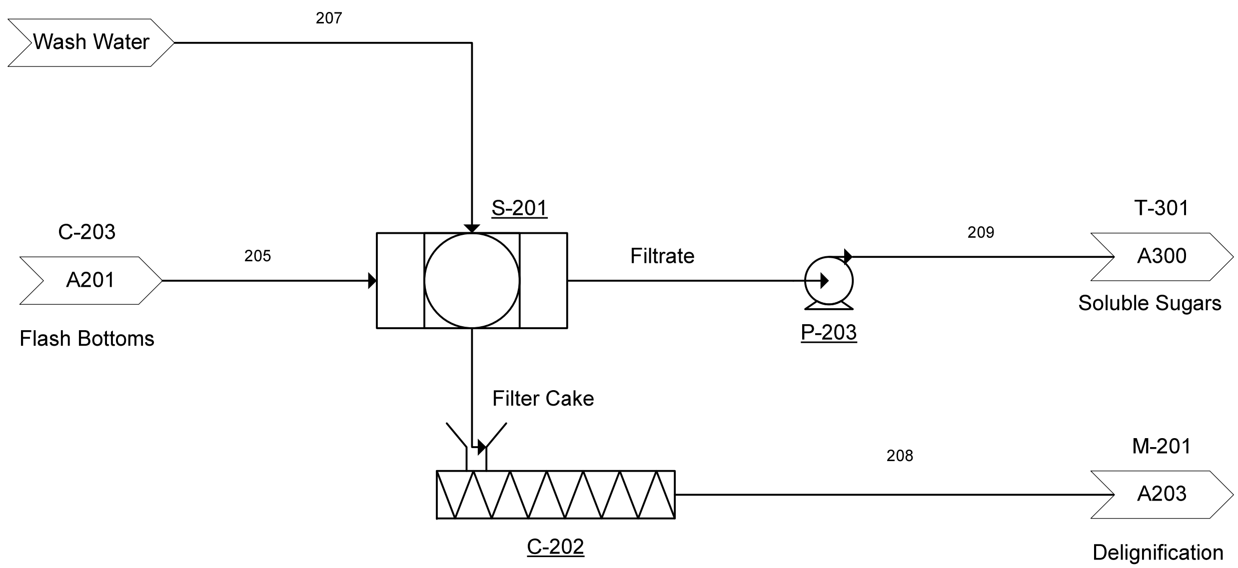

Figure B.1.2: A202- Base Case Pretreatment - Solids Wash

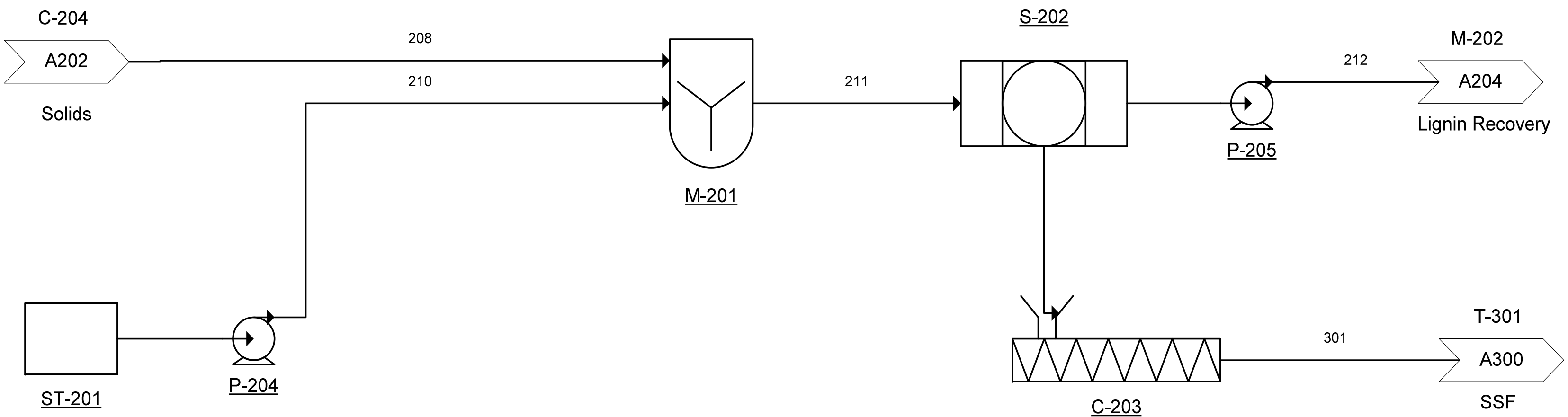

Figure B.1.3: A203- Base Case Pretreatment – Lignin Extraction

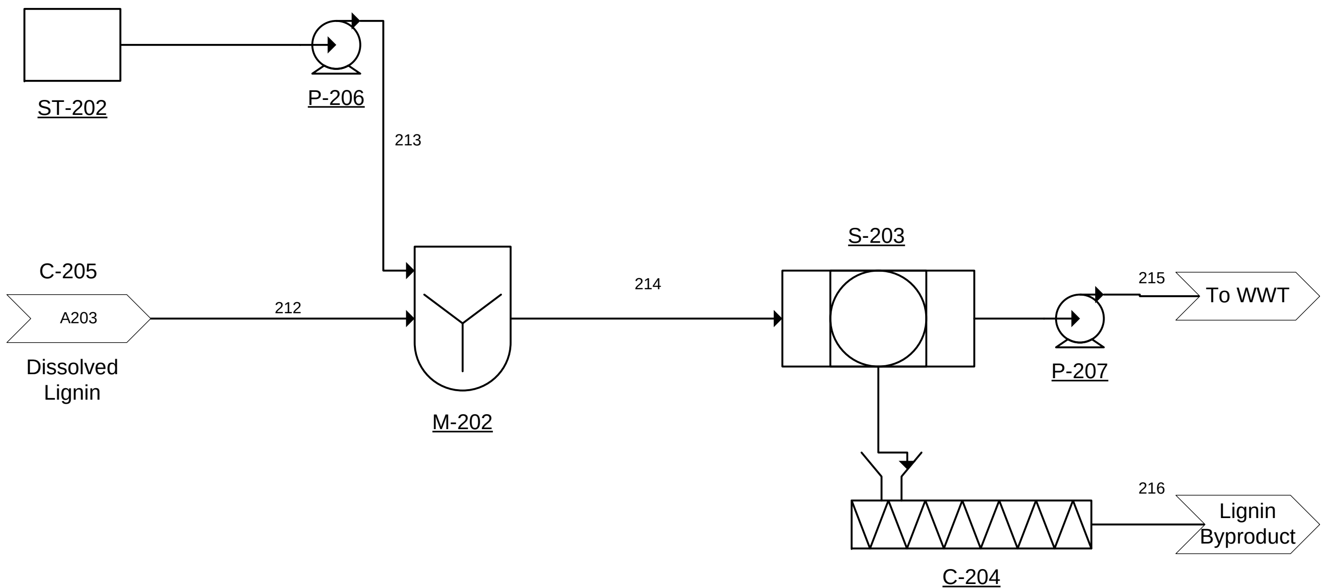

Figure B.1.4: A204- Base Case Pretreatment – Lignin Recovery

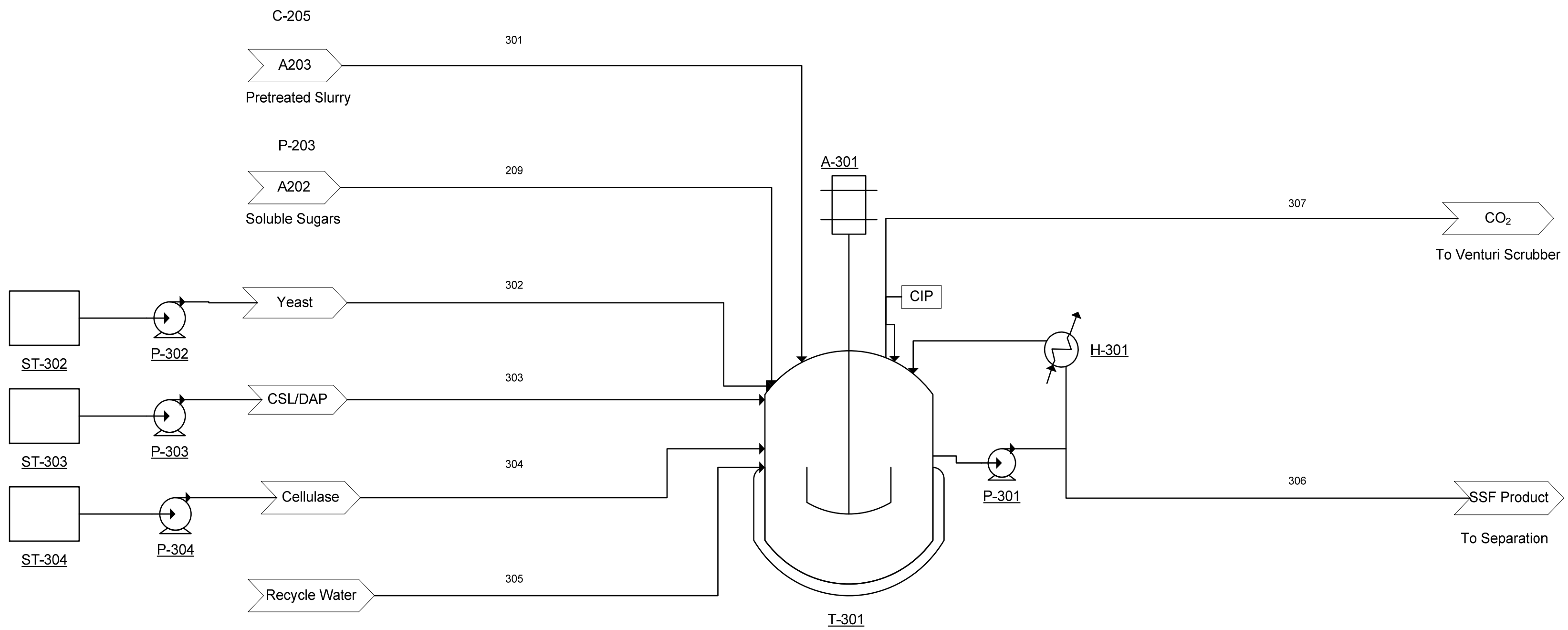

Figure B.1.5: A300- Base Case SSF

B.2 Process Flow Diagrams: Case 2

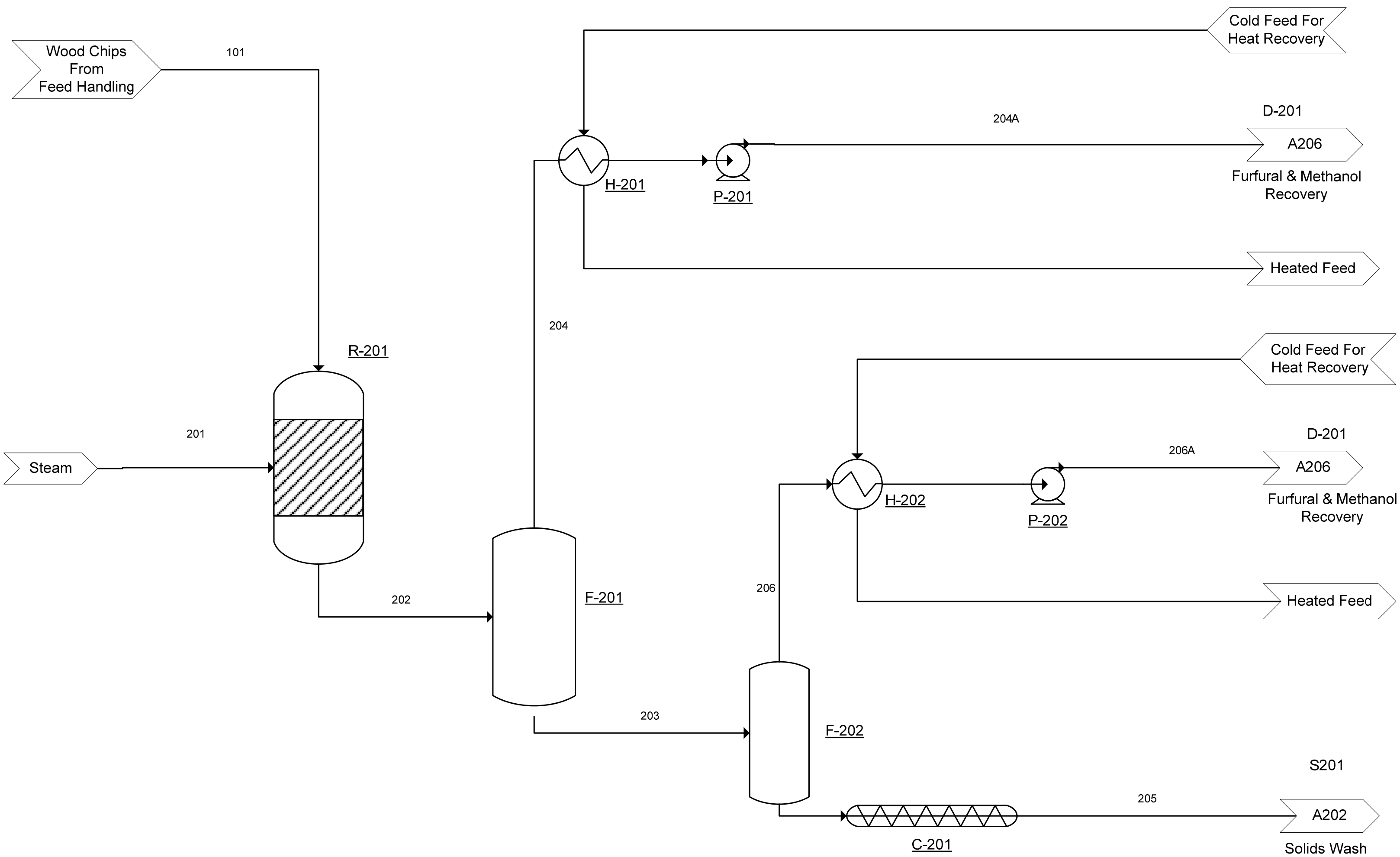

Figure B.2.1: A201- Case 2 Pretreatment - Digestor and Flash System

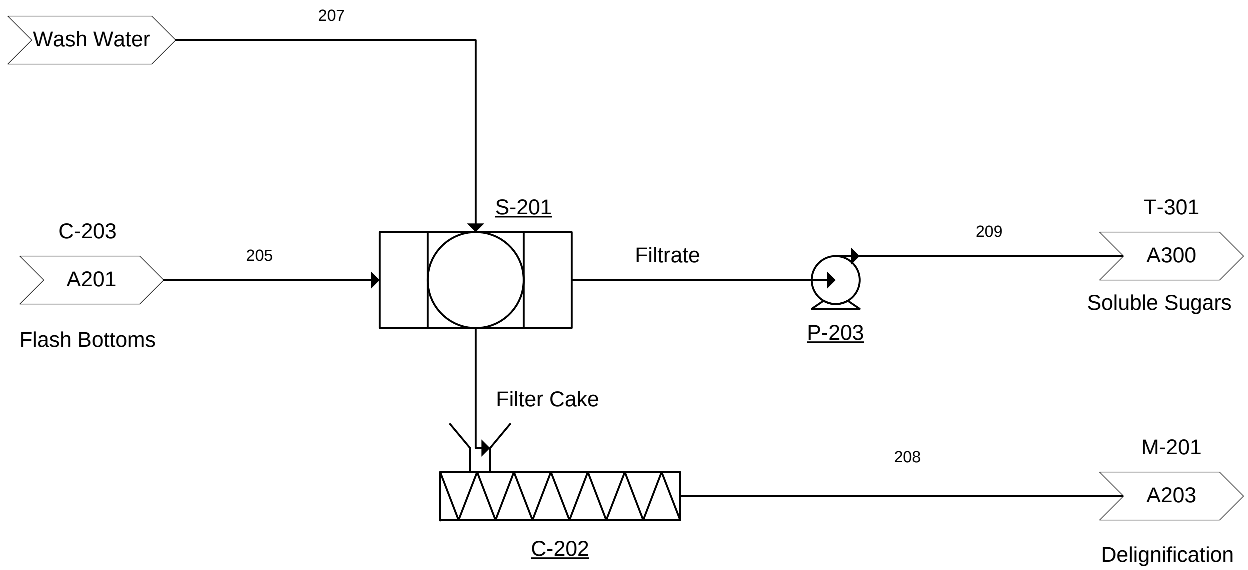

Figure B.2.2: A202- Case 2 Pretreatment - Solids Wash

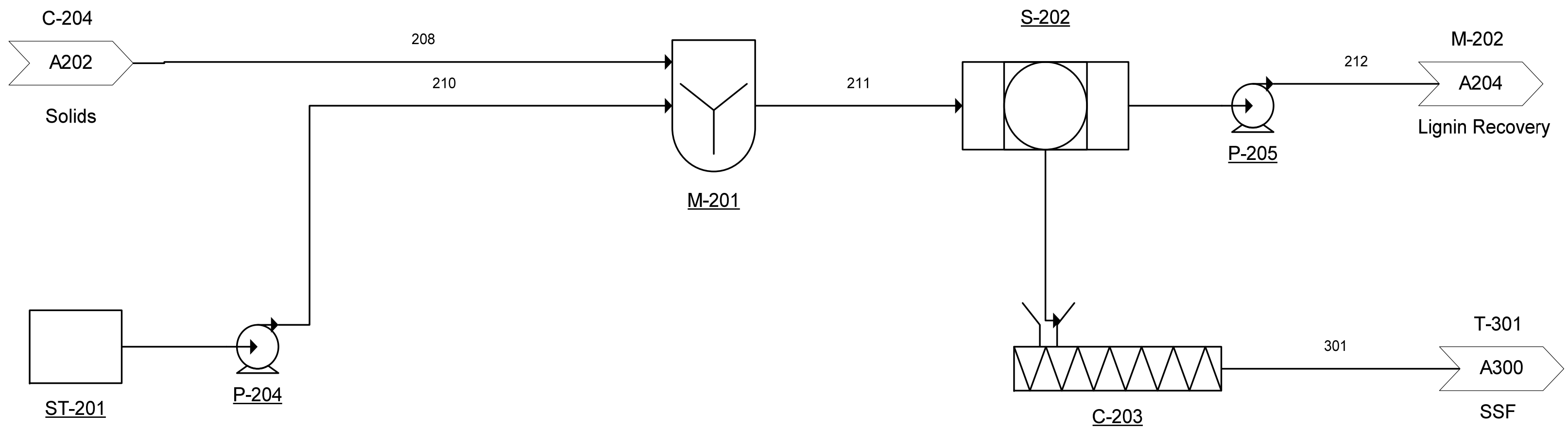

Figure B.2.3: A203- Case 2 Pretreatment – Lignin Extraction

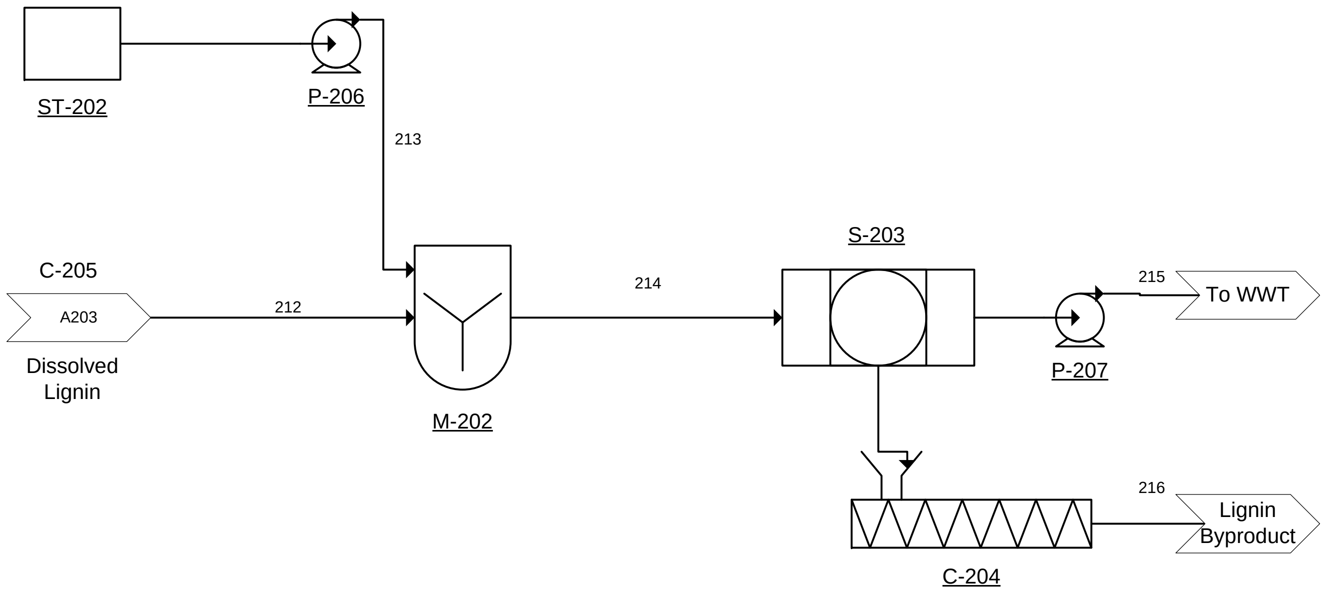

Figure B.2.4: A204- Case 2 Pretreatment – Lignin Recovery

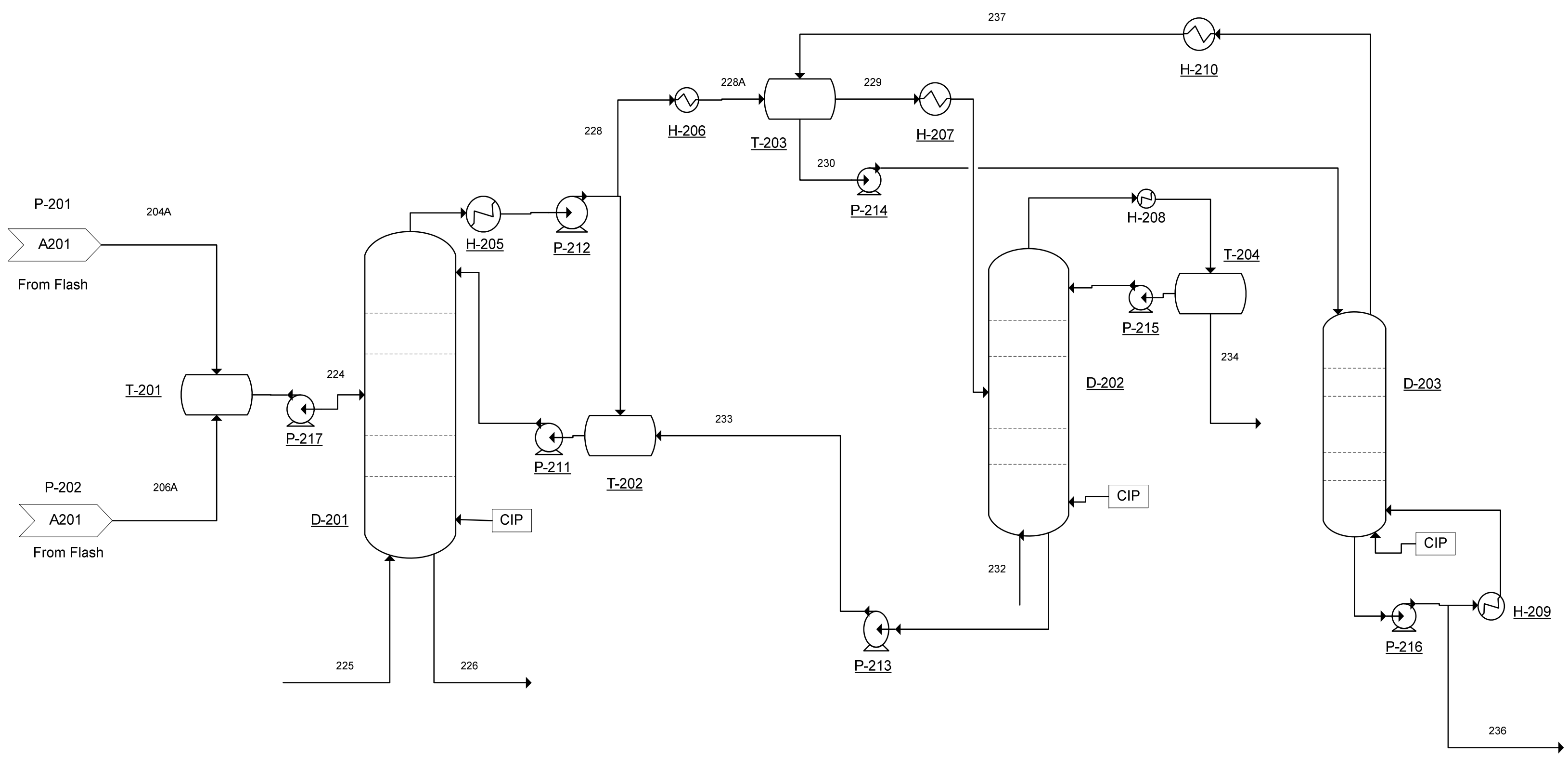

Figure B.2.5: A206- Case 2 Pretreatment – Furfural and Methanol Recovery

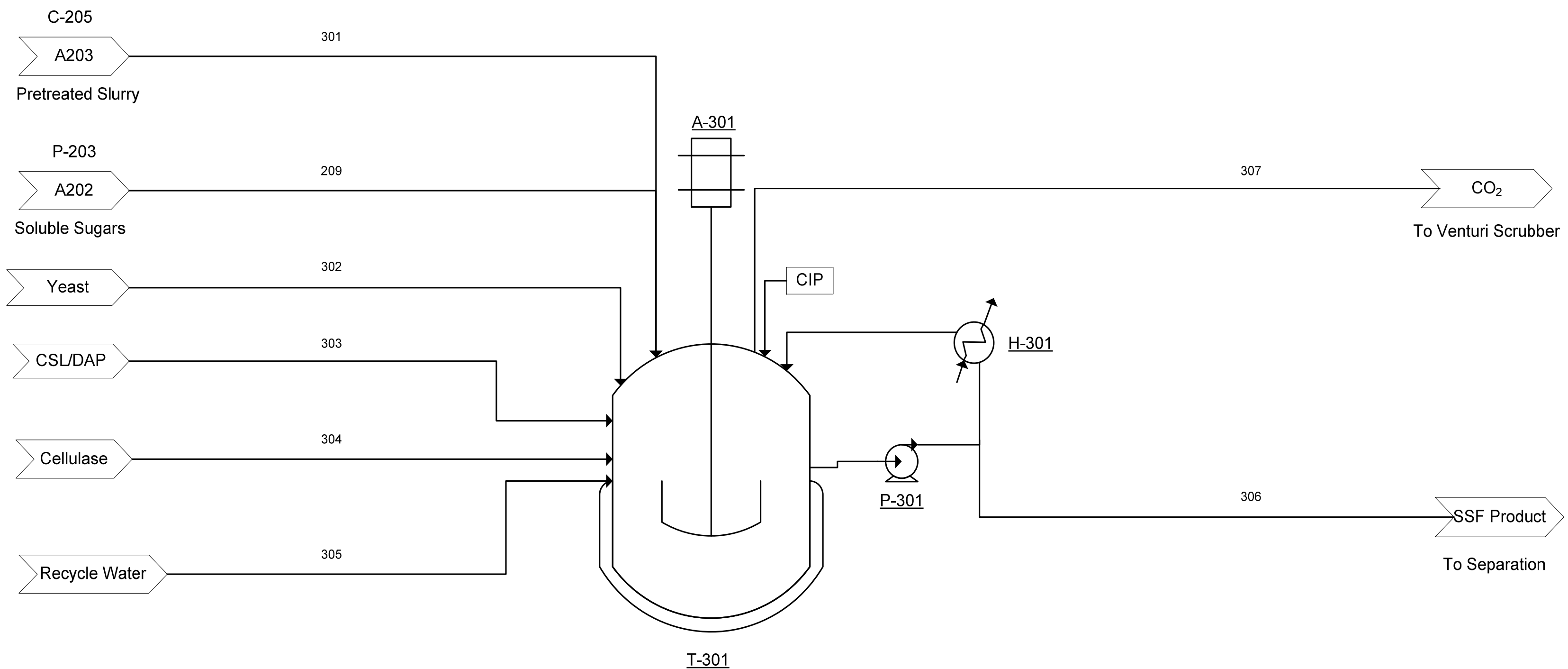

Figure B.2.6: A300- Case 2 SSF

B.3 Process Flow Diagrams: Case 3

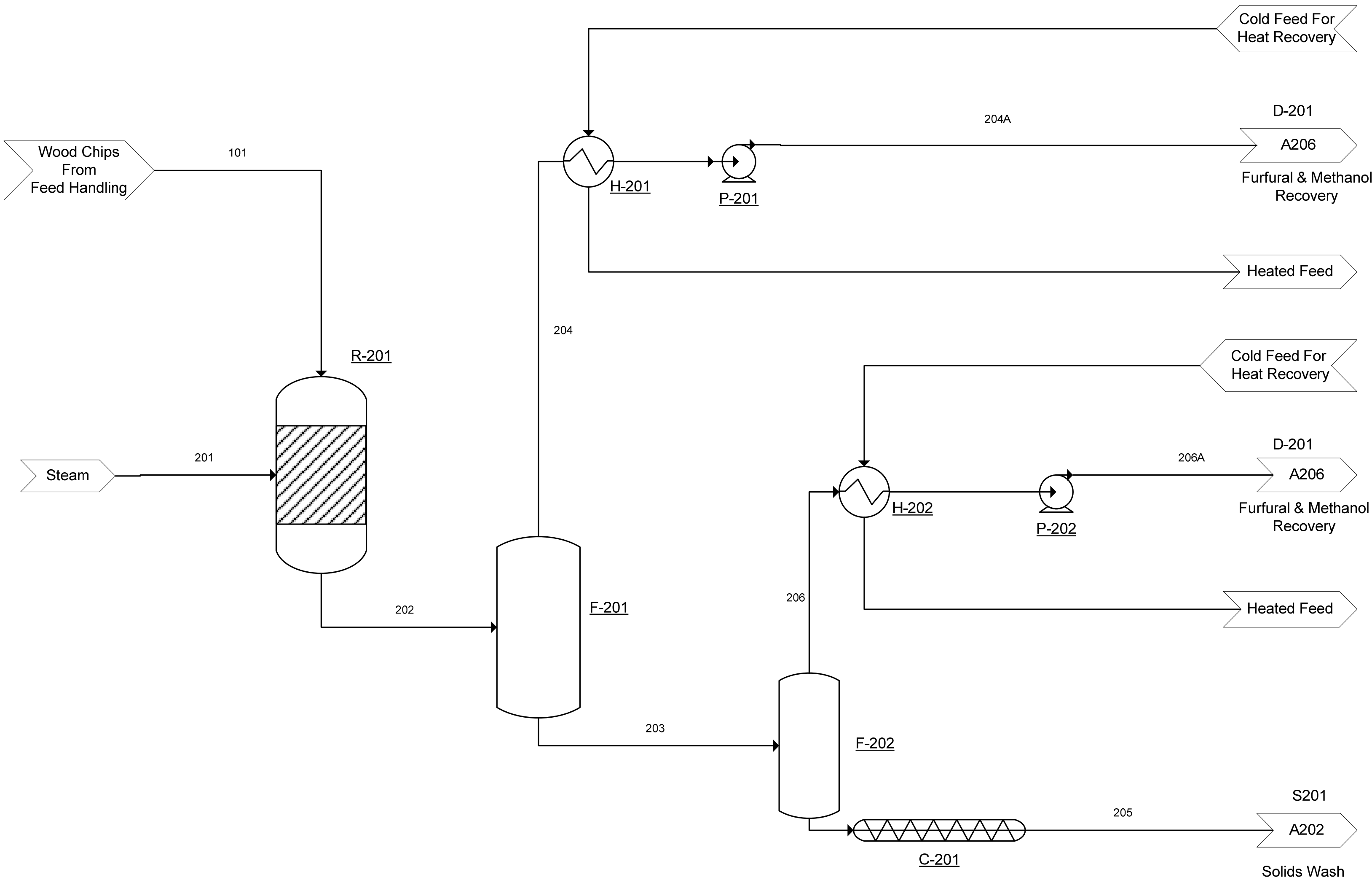

Figure B.3.1: A201- Case 3 Pretreatment - Digestor and Flash System

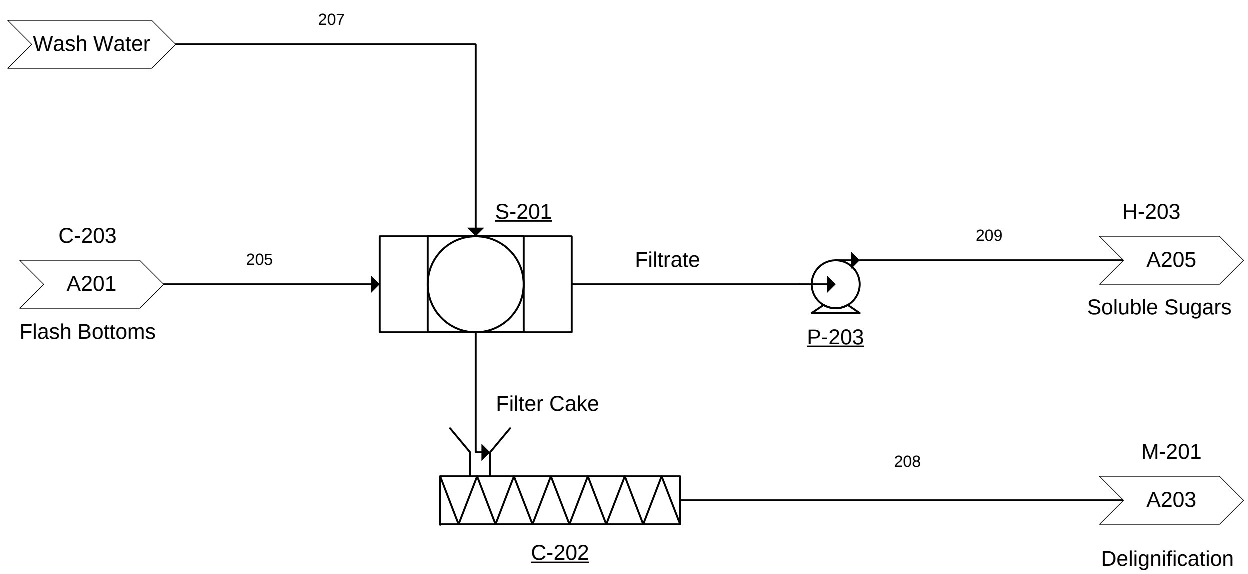

Figure B.3.2: A202- Case 3 Pretreatment - Solids Wash

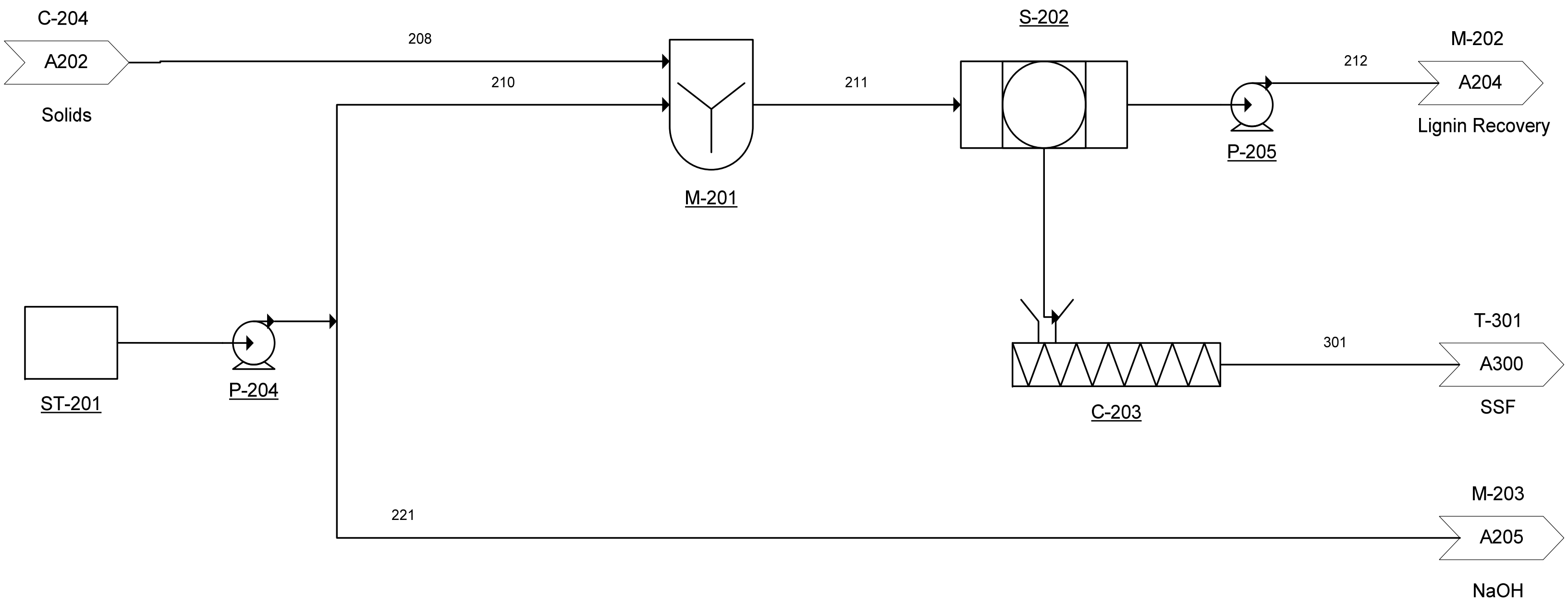

Figure B.3.3: A203- Case 3 Pretreatment – Lignin Extraction

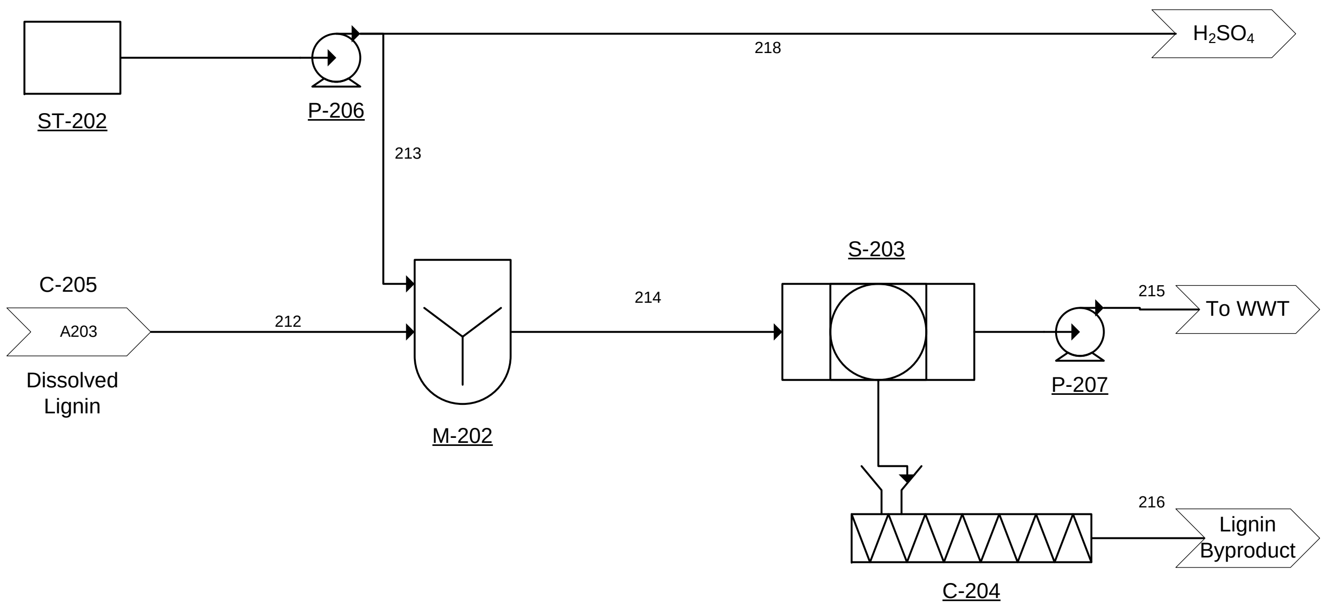

Figure B.3.4: A204- Case 3 Pretreatment – Lignin Recovery

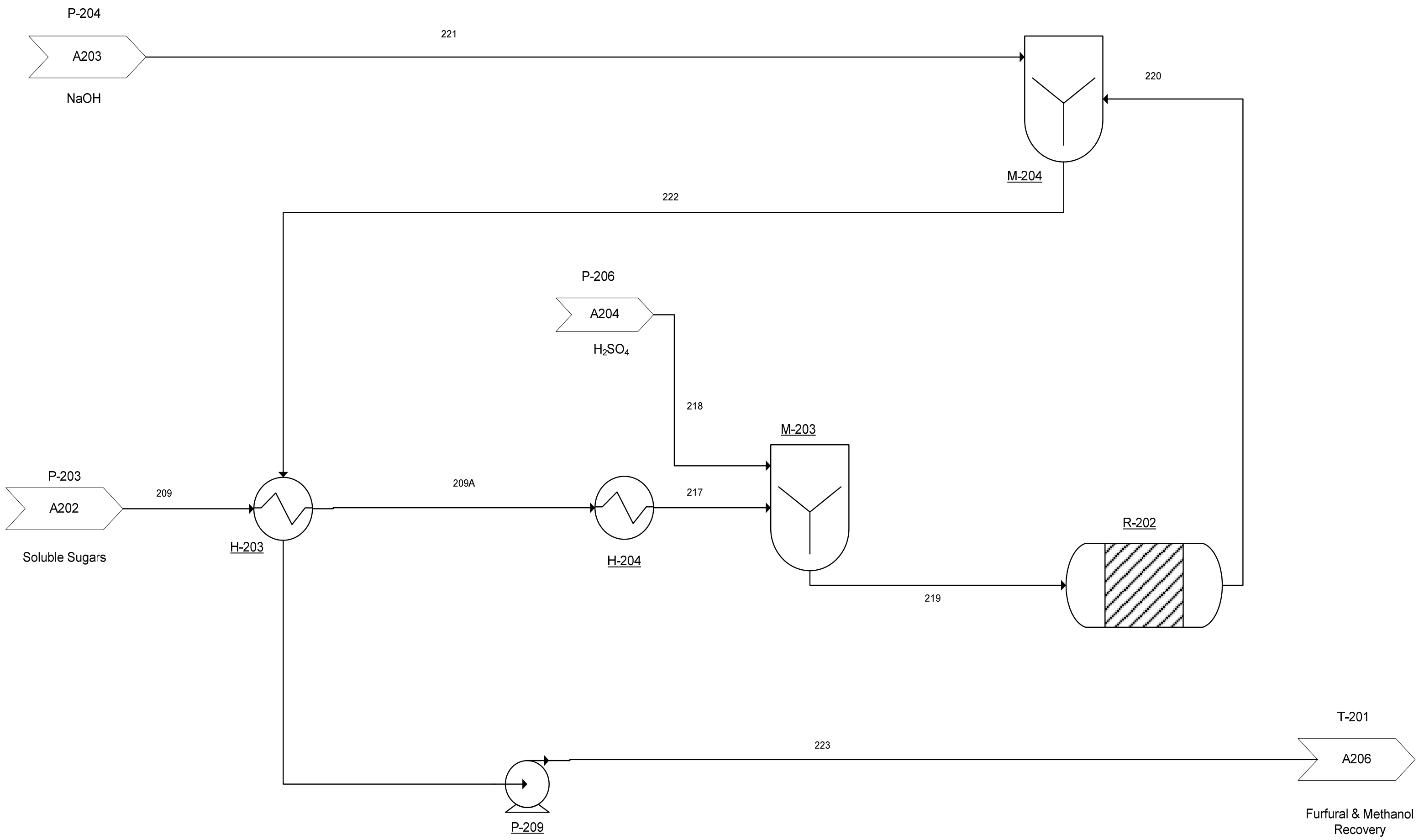

Figure B.3.5: A205- Case 3 Pretreatment – Xylose Conversion

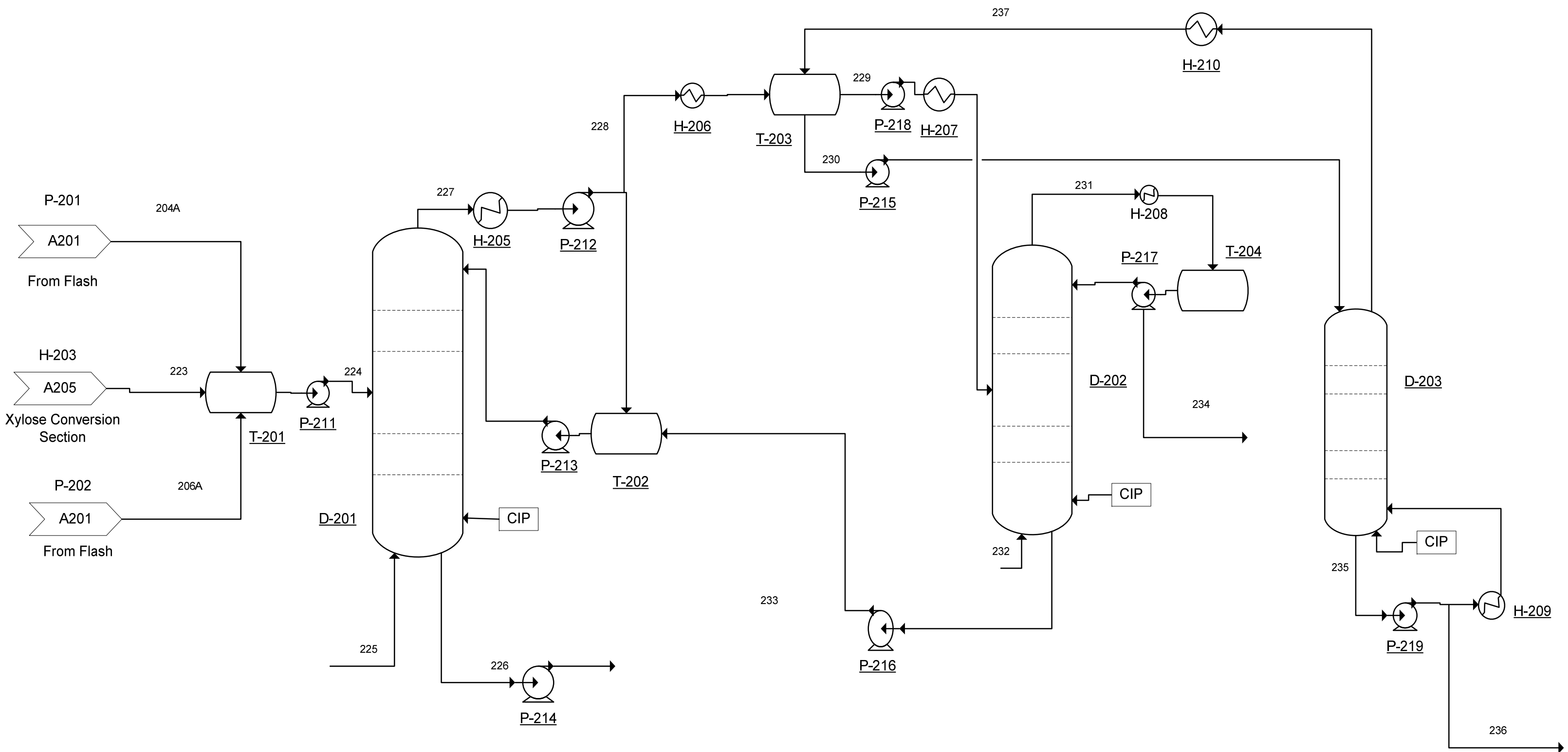

Figure B.3.6: A206- Case 3 Pretreatment – Furfural and Methanol Recovery

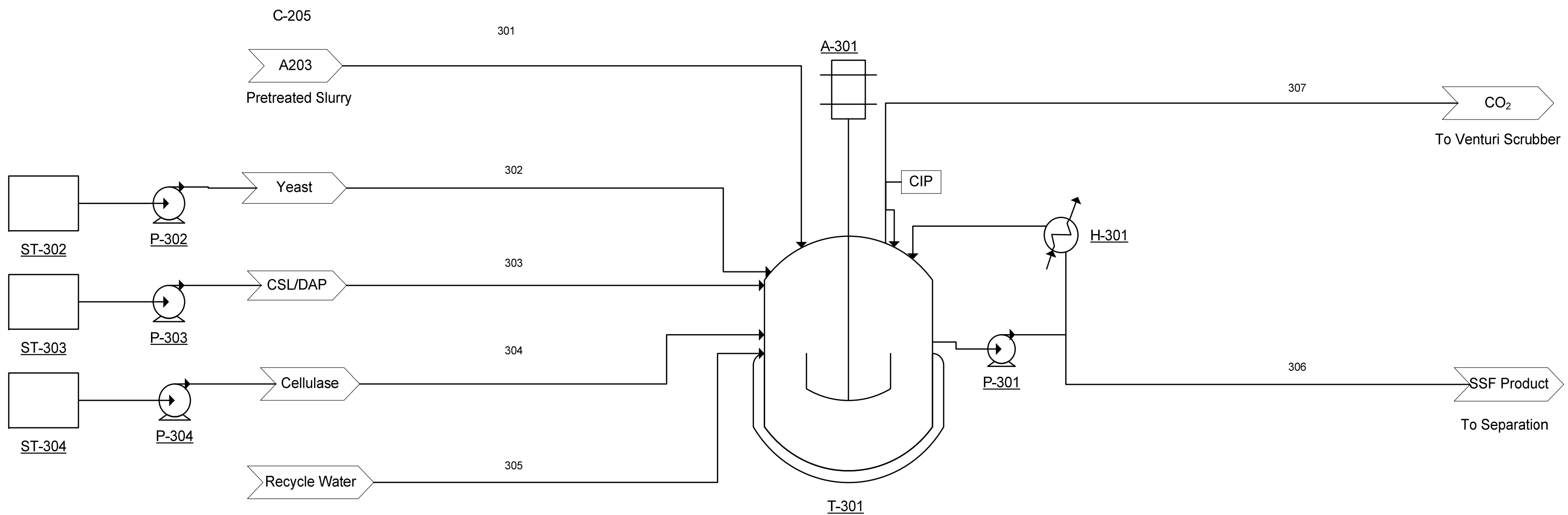

Figure B.3.7: A300- Case 3 SSF

B.4 Process Flow Diagrams: Case 4 Organosolv Pretreatment

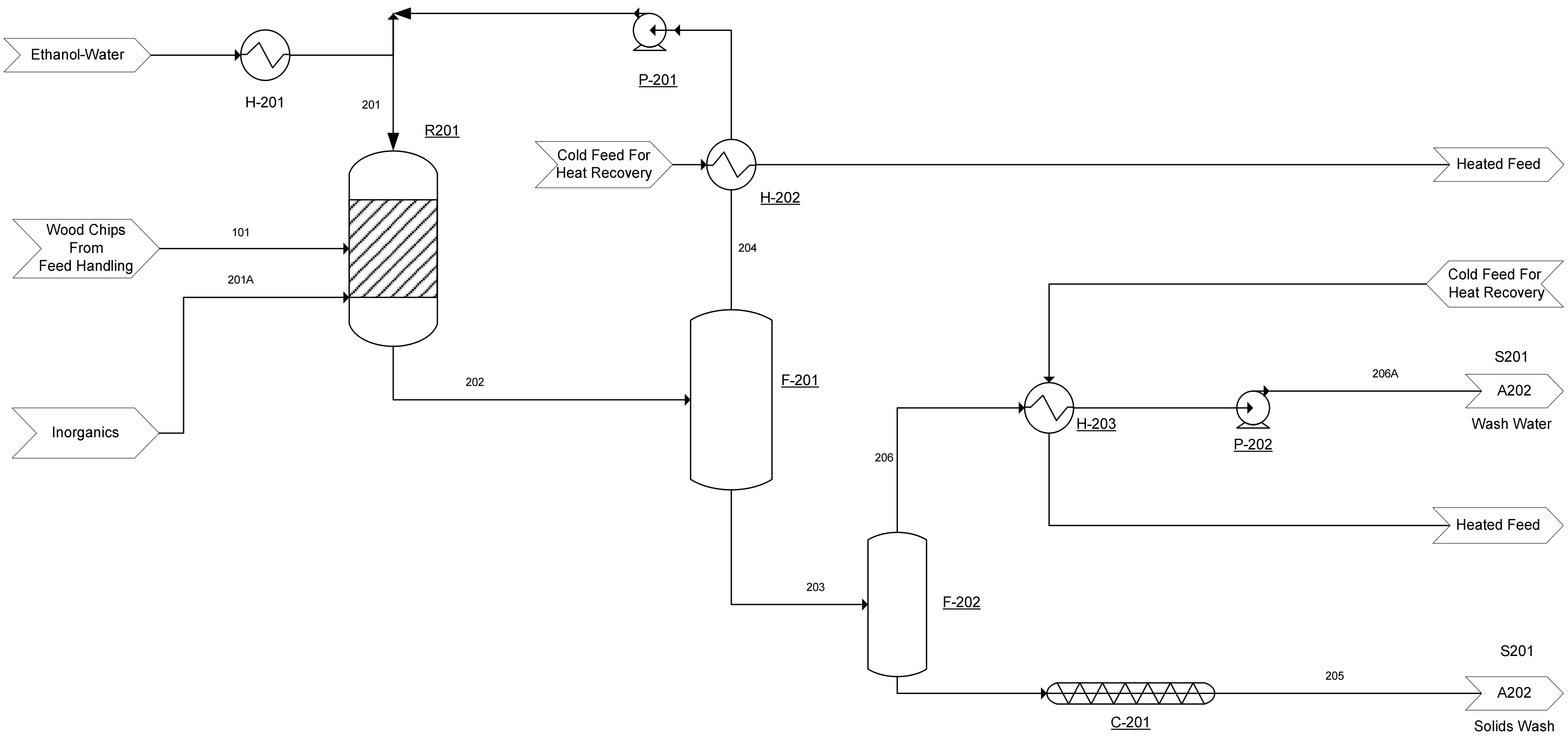

Figure B.4.1: A201- Case 4 Pretreatment - Digestor and Flash System

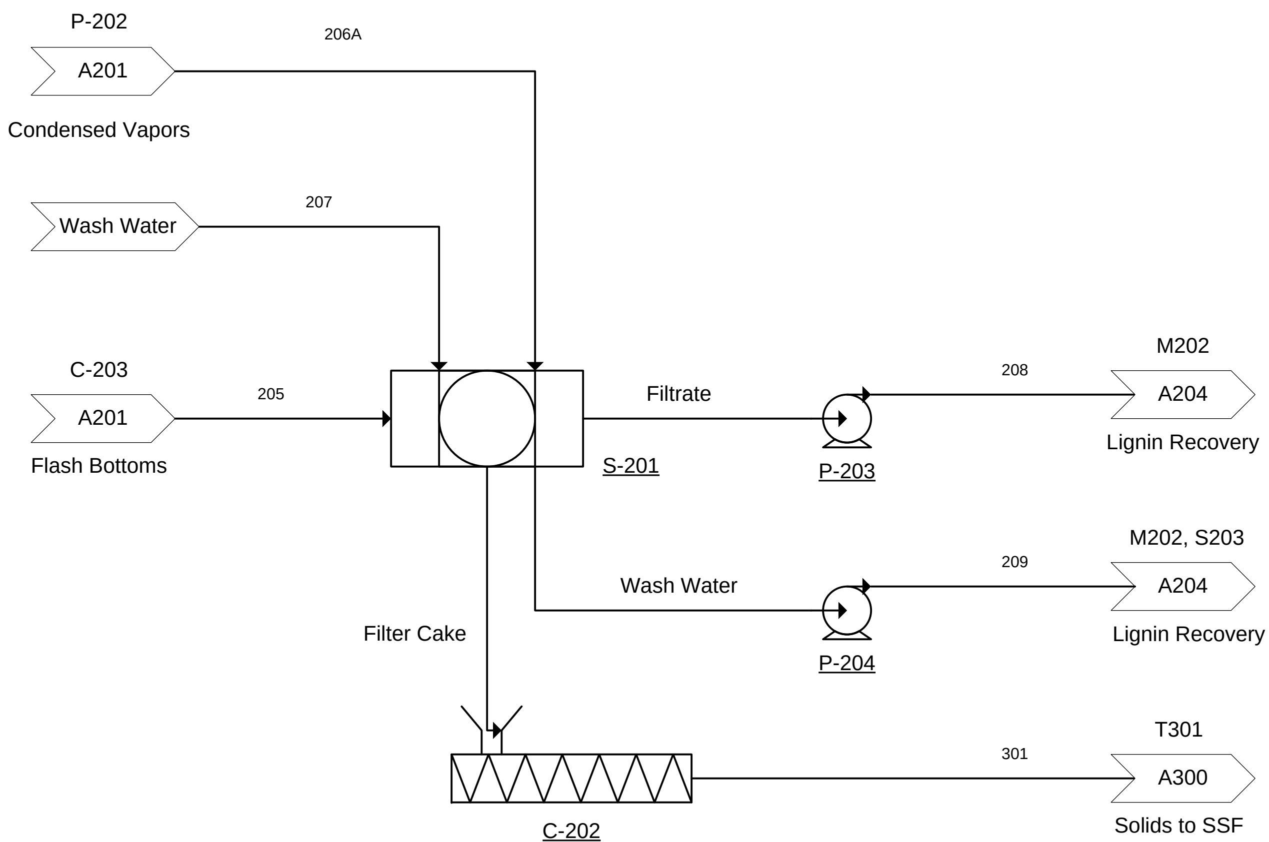

Figure B.4.2: A202- Case 4 Pretreatment - Solids Wash

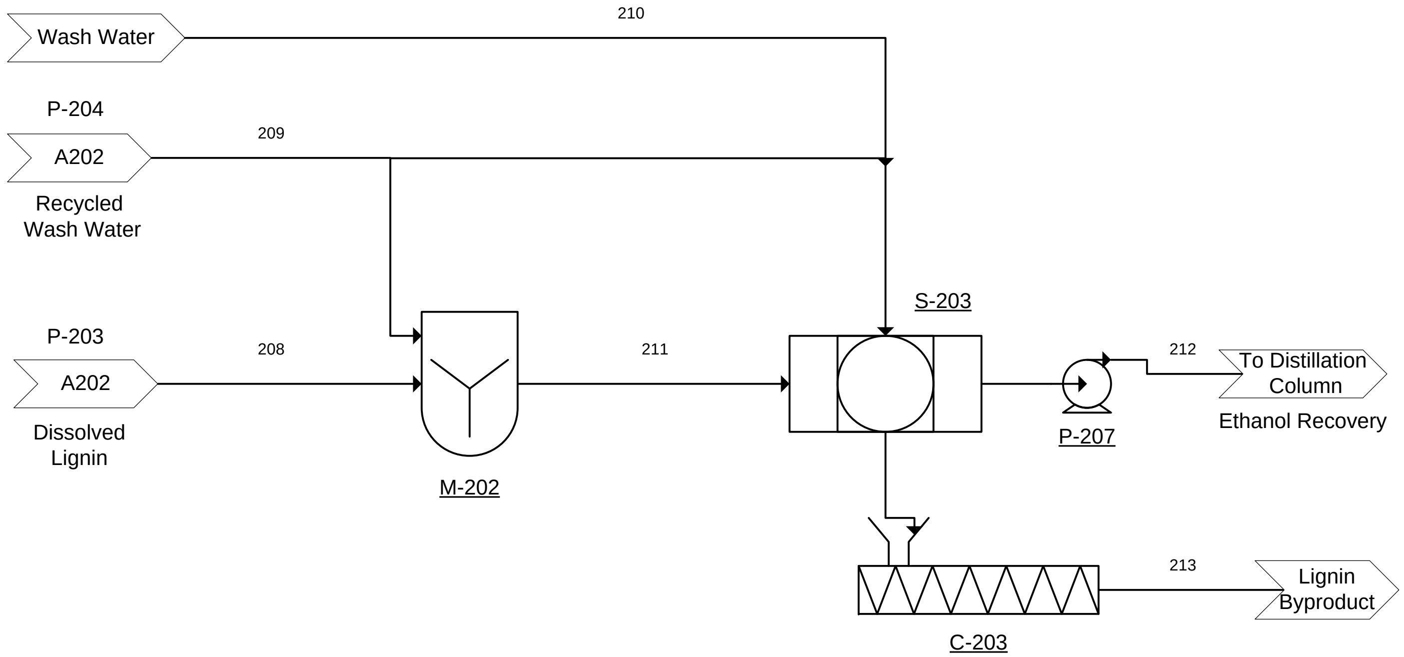

Figure B.4.3: A204- Case 4 Pretreatment – Lignin Recovery

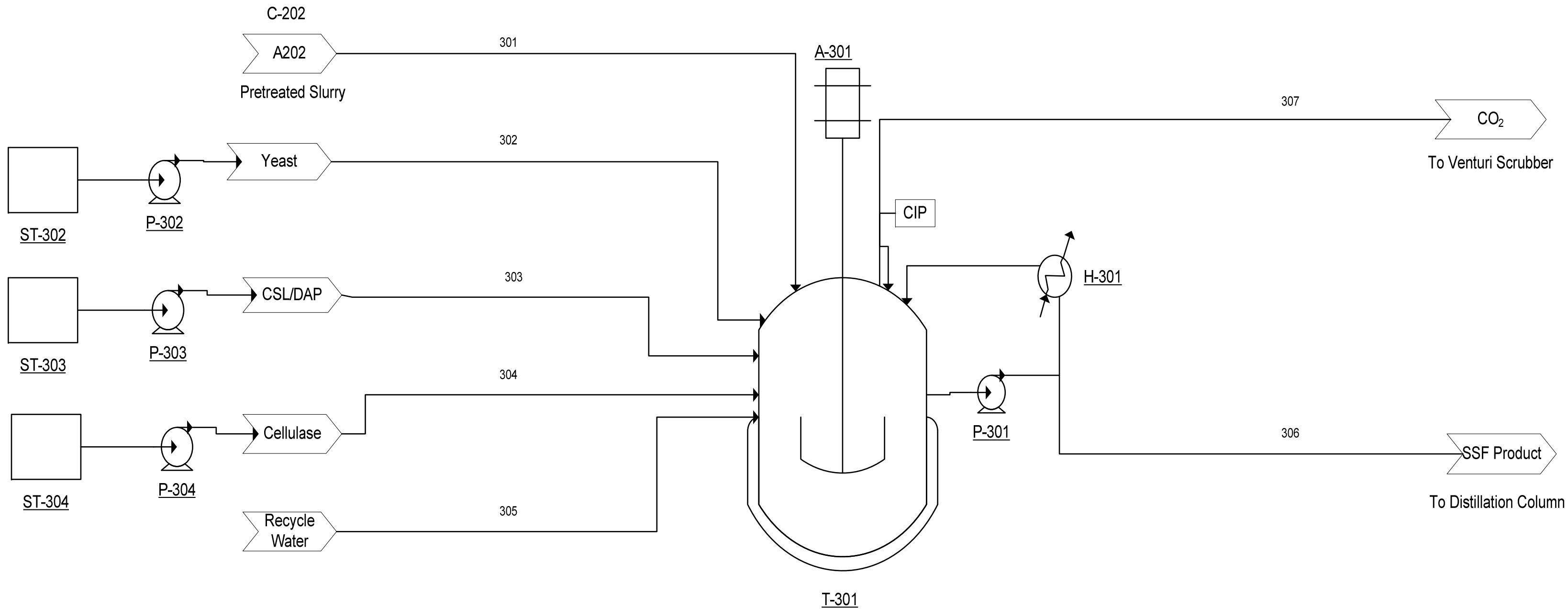

Figure B.4.4: A300- Case 4 SSF

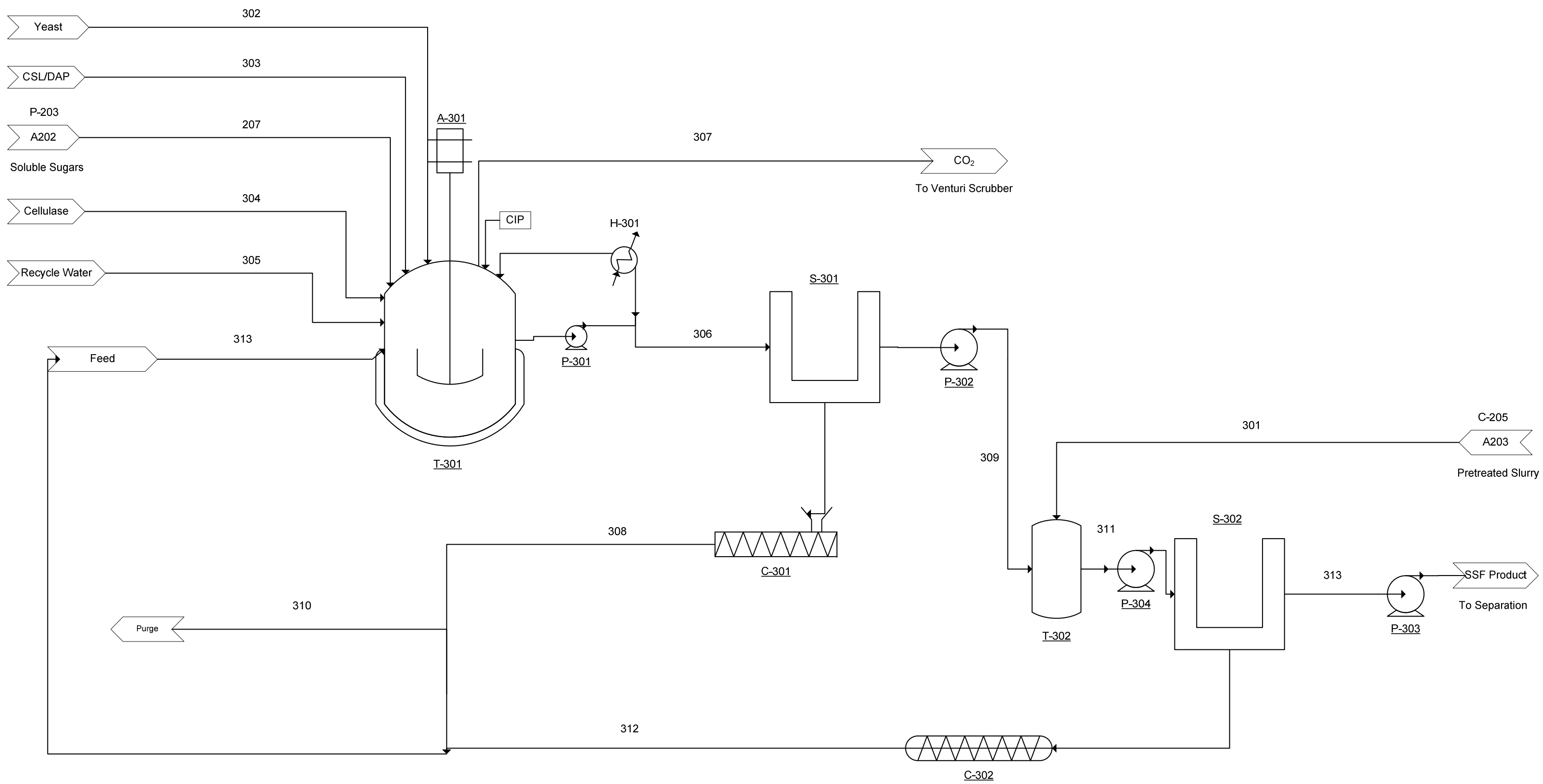

Figure B.5.1: A300 - Enzyme Recycling Scheme for Single Train of Reactors

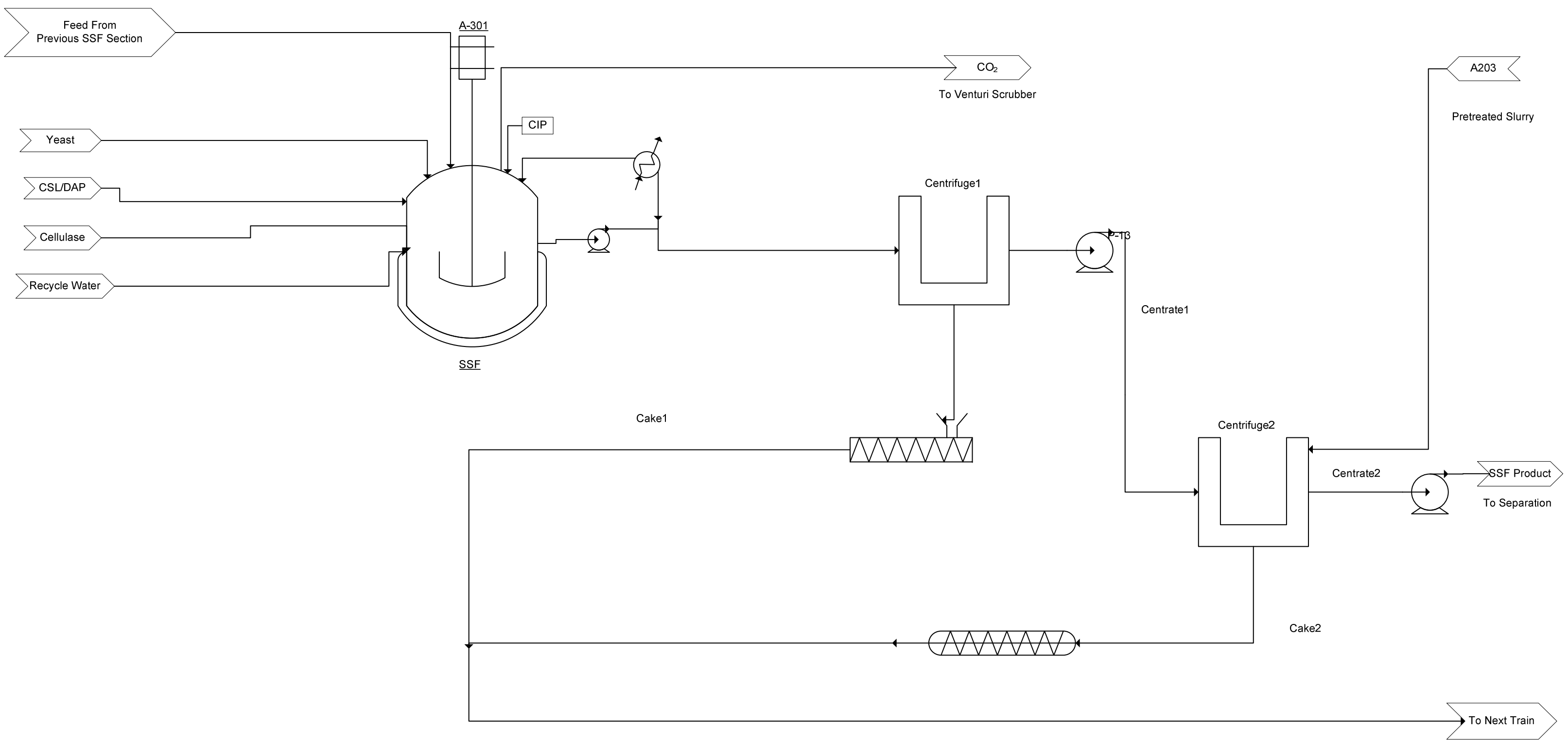

Figure B.5.2: A300 - Enzyme Recycling Scheme for Parallel Train of Reactor

APPENDIX C

MATERIAL BALANCES

<u>C.1 Base Case: Stream Results Summary</u>

Table C.1.1: Base Case Material Balance

Mass Flow lb/hr	101	201	202	203	204	204A	205	206	206A	207	208	209	210	211	212
Glucose	0	0	914	914	0	0	914	0	0	0	283	934	0	283	255
Xylose	0	0	4734	4734	0	0	4734	0	0	0	1453	4800	0	1453	1308
Arabinose	0	0	83	83	0	0	83	0	0	0	26	85	0	26	23
Galactose	0	0	581	581	0	0	581	0	0	0	174	575	0	174	157
Mannose	0	0	914	914	0	0	914	0	0	0	283	934	0	283	255
Cellulose	41667	0	40444	40444	0	0	40444	0	0	0	40444	0	0	40444	0
Xylan	14420	0	1246	1246	0	0	1246	0	0	0	1246	0	0	1246	0
Arabinan	587	0	83	83	0	0	83	0	0	0	83	0	0	83	0
Mannan	3102	0	664	664	0	0	664	0	0	0	664	0	0	664	0
Galactan	838	0	166	166	0	0	166	0	0	0	166	0	0	166	0
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Lignin (Insoluble)	22384	0	26200	26200	0	0	26200	0	0	0	26200	0	0	3668	0
Lignin (Soluble)	0	0	3613	3613	0	0	3613	0	0	0	1089	3598	0	23621	21259
Cellulase	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Zymo Yeast	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DAP	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CSL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	335	0	332	332	0	0	332	0	0	0	332	0	0	332	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	83333	55000	138371	101906	36465	36465	90004	11902	11902	194962	65202	215412	4980	70182	146076
Furfural	0	0	1329	501	827	827	290	212	212	0	217	718	0	217	195
Methanol	0	0	249	127	122	122	88	39	39	0	43	144	0	43	39
Acetic Acid	0	0	1744	1461	283	283	1362	99	99	0	366	1209	0	366	329
SSF Products	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
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NAOH	0	0	0	0	0	0	0	0	0	0	0	0	2760	2760	2484
Sodium Sulfate	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sodium Acetate	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Solid Fraction (Mass)	0.50	0.00	0.31	0.38	0.00	0.00	0.40	0.00	0.00	0.00	0.50	0.00	0.00	0.32	0.00
Total Flow lb/hr	166666	55000	221666	183970	37697	37697	171717	12252	12252	194962	138271	228408	7740	146011	172379
Total Flow cuft/hr	2357	73406	3940	2907	250671	604	2578	324460	196	3133	1960	3647	116	2089	2656
Temperature F	70	436	436	298	298	70	213	213	70	70	91	91	70	93	80
9993Pressure psi	14.7	364.7	364.7	64.7	64.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7

Table C.1.1(Continued)

Mass Flow lb/hr	213	214	214A	215	216	301	302	303	304	305	306	307
Glucose	0	255	0	220	69	28	0	0	0	0	1765	0
Xylose	0	1308	0	1130	353	145	0	0	0	0	4945	0
Arabinose	0	23	0	20	6	3	0	0	0	0	88	0
Galactose	0	157	0	135	42	17	0	0	0	0	37	0
Mannose	0	255	0	220	69	28	0	0	0	0	78	0
Cellulose	0	0	0	0	0	40444	0	0	0	0	7053	0
Xylan	0	0	0	0	0	1246	0	0	0	0	1246	0
Arabinan	0	0	0	0	0	83	0	0	0	0	83	0
Mannan	0	0	0	0	0	664	0	0	0	0	88	0
Galactan	0	0	0	0	0	166	0	0	0	0	22	0
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	2750	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	45	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	11	0
Lignin (Insoluble)	0	21259	0	0	11621	3668	0	0	0	0	3668	0
Lignin (Soluble)	0	0	0	0	0	2362	0	0	0	0	5960	0
Cellulase	0	0	0	0	0	0	0	0	12133	0	10313	0
Zymo Yeast	0	0	0	0	0	0	315	0	0	0	1141	0
DAP	0	0	0	0	0	0	0	195	0	0	181	0
CSL	0	0	0	0		0	0	1547	0	0	1500	0
Ash	0	0	0	0	0	332	0	0	0	0	332	0
Ethanol	0	0	0	0	0	0	0	0	0	0	17404	146
Water	2380	149575	92125	129263	9082	16231	66938	0	133522	178841	540478	288
Furfural	0	195	0	169	26	22	0	0	0	0	736	3
Methanol	0	39	0	34	5	4	0	0	0	0	148	0
Acetic Acid	0	329	0	285	44	37	0	0	0	0	1288	0
Lactic Acid	0	0	0	0		0	0	0	0	0	58	0
H2SO4	3050	4	0	4	704	0	0	0	0	0	0	0
CO2	0	0	0	0		0	0	0	0	0	770	15988
NAOH	0	0	0	0	0	276	0	0	0	0	276	0
Sodium Sulfate	0	4411	0	3812	1074	0		0	0	0	0	0
Sodium Acetate		0	0	0	0	0	0	0	0	0	0	0
Solid Fraction (Mass)	0.00	0.12	0.00	0.00	0.50	0.71	0.00	0.00	0.08	0.00	0.04	0.00
Total Flow lb/hr	5430	177809	92125	135291	23095	65756	67253	1742	145655	178841	602466	16431
Total Flow cuft/hr	74	2778	1479	2205	314	911	1717	30	2273	2871	9626	151700
Temperature F	70	70	68	70	70	80	68	82	68	68	86	86
Pressure psi	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7	14.7

C.2: Case 2: Streams Results Summary

Table C.2.1: Case 2 Material Balance

Mass Flow lb/hr	**101**	**201**	**202**	**203**	**204**	**204A**	**205**	**206**	**206A**	**207**	**208**	**209**	**210**	**211**	**212**
Glucose	0	0	914	914	0	0	914	0	0	0	283	934	0	283	255
Xylose	0	0	4734	4734	0	0	4734	0	0	0	1453	4800	0	1453	1308
Arabinose	0	0	83	83	0	0	83	0	0	0	26	85	0	26	23
Galactose	0	0	581	581	0	0	581	0	0	0	174	575	0	174	157
Mannose	0	0	914	914	0	0	914	0	0	0	283	934	0	283	255
Cellulose	41667	0	40444	40444	0	0	40444	0	0	0	40444	0	0	40444	0
Xylan	14420	0	1246	1246	0	0	1246	0	0	0	1246	0	0	1246	0
Arabinan	587	0	83	83	0	0	83	0	0	0	83	0	0	83	0
Mannan	3102	0	664	664	0	0	664	0	0	0	664	0	0	664	0
Galactan	838	0	166	166	0	0	166	0	0	0	166	0	0	166	0
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Lignin (Insoluble)	22384	0	26200	26200	0	0	26200	0	0	0	26200	0	0	3668	0
Lignin (Soluble)	0	0	3613	3613	0	0	3613	0	0	0	1089	3598	0	23621	21259
Cellulase	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Zymo Yeast	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DAP	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CSL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	335	0	332	332	0	0	332	0	0	0	332	0	0	332	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	83333	55000	138371	101906	36465	36465	90004	11902	11902	194962	65202	215412	4980	70182	146076
Furfural	0	0	1329	501	827	827	290	212	212	0	217	718	0	217	195
Methanol	0	0	249	127	122	122	88	39	39	0	43	144	0	43	39
Acetic Acid	0	0	1744	1461	283	283	1362	99	99	0	366	1209	0	366	329
Others	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
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NAOH	0	0	0	0	0	0	0	0	0	0	0	0	2760	2760	2484
Sodium Sulfate	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Solid Fraction (Mass)	1	0	0	0	0	0	0	0	0	0.00	1	0	0	0	0
Total Flow lb/hr	166666	55000	221666	183970	37697	37697	171717	12252	12252	194962	138271	228408	7740	146011	172379
Total Flow cuft/hr	2357	73406	3940	2907	250671	604	2578	324460	196	3133	1960	3647	116	2089	2656
Temperature F	70	436	436	298	298	70	213	213	70	70	91	91	70	93	80
Pressure psi	15	365	365	65	65	15	15	15	15	15	15	15	15	15	15

Table C.2.1 (Continued)

Mass Flow lb/hr	213	214	214A	215	215	216	224	225	226	228	228A	229	230	232	233	234	236	237
Glucose	0	255	0	220	0	69	0	0	0	0	0	0	0	0	0	0	0	0
Xylose	0	1308	0	1130	0	353	0	0	0	0	0	0	0	0	0	0	0	0
Arabinose	0	23	0	20	0	6	0	0	0	0	0	0	0	0	0	0	0	0
Galactose	0	157	0	135	0	42	0	0	0	0	0	0	0	0	0	0	0	0
Mannose	0	255	0	220	0	69	0	0	0	0	0	0	0	0	0	0	0	0
Cellulose	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
Arabinan	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannan	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactan	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Lignin (Insoluble)	0	21259	0	0	0	11621	0	0	0	0	0	0	0	0	0	0	0	0
Lignin (Soluble)	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Cellulase	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Zymo Yeast	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DAP	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CSL	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
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	2380	149575	92125	129263	10310	9082	48367	6930	56501	6004	6004	5992	78	1220	7208	1	12	67
Furfural	0	195	0	169	13	26	1039	0	447	1295	1295	703	623	0	703	0	592	31
Methanol	0	39	0	34	3	5	161	0	36	315	315	315	0	0	190	125	0	0
Acetic Acid	0	329	0	285	22	44	382	0	382	0	0	0	0	0	0	0	0	0
Others	0	0	0	0	0		0	0	0	0	0	0	0	0	0	0	0	0
H2SO4	3050	4	0	4	0	704	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
NAOH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sodium Sulfate	0	4411	0	3812	0	1074	0	0	0	0	0	0	0	0	0	0	0	0
Solid Fraction (Mass)	0	0	0.00	0	0	0.50	0	0	0	0	0	0	0	0	0	0	0	0
Total Flow lb/hr	5430	177809	92125	135291	10349	23095	49949	6930	57366	7614	7614	7010	702	1220	8101	126	605	97
Total Flow cuft/hr	74	2778	1479	2205	166	314	869	20696	1031	78871	126	117	10	3644	144	3	9	2
Temperature F	70	70	68	70	70	70	210	366	258	255	150	151	151	366	252	191	283	207
Pressure psi	15	15	15	15	15	14.7	41	165	35	35	5	35	35	165	35	35	20	15

Table C.2.1 (Continued)

Mass Flow lb/hr	**301**	**302**	**303**	**304**	**305**	**306**	**307**
Glucose	28	0	0	0	0	1765	0
Xylose	145	0	0	0	0	4945	0
Arabinose	3	0	0	0	0	88	0
Galactose	17	0	0	0	0	37	0
Mannose	28	0	0	0	0	78	0
Cellulose	40444	0	0	0	0	7053	0
Xylan	1246	0	0	0	0	1246	0
Arabinan	83	0	0	0	0	83	0
Mannan	664	0	0	0	0	88	0
Galactan	166	0	0	0	0	22	0
Glucose Oligomer	0	0	0	0	0	2750	0
Mannose Oligomer	0	0	0	0	0	45	0
Galactose Oligomer	0	0	0	0	0	11	0
Lignin (Insoluble)	3668	0	0	0	0	3668	0
Lignin (Soluble)	2362	0	0	0	0	5960	0
Cellulase	0	0	0	12133	0	10313	0
Zymo Yeast	0	315	0	0	0	1141	0
DAP	0	0	195	0	0	181	0
CSL	0	0	1547	0	0	1500	0
Ash	332	0	0	0	0	332	0
Ethanol	0	0	0	0	0	17404	146
Water	16231	66938	0	133522	111903	540478	288
Furfural	22	0	0	0	0	736	3
Methanol	4	0	0	0	0	148	0
Acetic Acid	37	0	0	0	0	1288	0
Others	0	0	0	0	0	58	0
H2SO4	0	0	0	0	0	0	0
CO2	0	0	0	0	0	770	15988
NAOH	276	0	0	0	0	276	0
Sodium Sulfate	0		0	0	0	0	0
Solid Fraction (Mass)	1	0.005	0	0	0	0	0
Total Flow lb/hr	65756	67253	1547	145655	178841	602466	16431
Total Flow cuft/hr	911	1717	27	2273	2871	9626	151700
Temperature F	80	68	95	68	68	86	86
Pressure psi	15	15	15	15	15	15	15

C.3: Case 3: Streams Results Summary

Table C.3.1: Case 3 Material Balance

Mass Flow lb/hr	101	201	202	203	204	204A	205	206	206A	207	208	209	209A	210	211
Glucose	0	0	914	914	0	0	914	0	0	0	283	934	934	0	283
Xylose	0	0	4734	4734	0	0	4734	0	0	0	1453	4800	4800	0	1453
Arabinose	0	0	83	83	0	0	83	0	0	0	26	85	85	0	26
Galactose	0	0	581	581	0	0	581	0	0	0	174	575	575	0	174
Mannose	0	0	914	914	0	0	914	0	0	0	283	934	934	0	283
Cellulose	41667	0	40444	40444	0	0	40444	0	0	0	40444	0	0	0	40444
Xylan	14420	0	1246	1246	0	0	1246	0	0	0	1246	0	0	0	1246
Arabinan	587	0	83	83	0	0	83	0	0	0	83	0	0	0	83
Mannan	3102	0	664	664	0	0	664	0	0	0	664	0	0	0	664
Galactan	838	0	166	166	0	0	166	0	0	0	166	0	0	0	166
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Lignin (Insoluble)	22384	0	26200	26200	0	0	26200	0	0	0	26200	0	0	0	3668
Lignin (Soluble)	0	0	3613	3613	0	0	3613	0	0	0	1089	3598	3598	0	23621
Cellulase	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Zymo Yeast	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DAP	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CSL	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ash	335	0	332	332	0	0	332	0	0	0	332	0	0	0	332
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	83333	55000	138371	101906	36465	36465	90004	11902	11902	194962	65202	215412	215412	4980	70182
Furfural	0	0	1329	501	827	827	290	212	212	0	217	718	718	0	217
Methanol	0	0	249	127	122	122	88	39	39	0	43	144	144	0	43
Acetic Acid	0	0	1744	1461	283	283	1362	99	99	0	366	1209	1209	0	366
Lactic Acid	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
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NAOH	0	0	0	0	0	0	0	0	0	0	0	0	0	2760	2760
Sodium Sulphate	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sodium Acetate	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Solid Fraction (Mass)	0.50	0.00	0.31	0.38	0.00	0.00	0.40	0.00	0.00	0.00	0.50	0.00	0.00	0.00	0.32
Total Flow lb/hr	166666	55000	221666	183970	37697	37697	171717	12252	12252	194962	138271	228408	228408	7740	146011
Total Flow cuft/hr	2357	73406	3940	2907	250671	604	2578	324460	196	3133	1960	3647	5708830	116	2089
Temperature F	70	436	436	298	298	70	213	213	70	70	91	91	219	70	93

Table C.3.1 (Continued)

Mass Flow lb/hr	**212**	**213**	**214**	**215**	**216**	**217**	**218**	**219**	**220**	**221**	**222**	**223**	**224**	**225**	**226**
Glucose	255	0	255	220	69	934	0	934	934	0	934	934	934	0	934
Xylose	1308	0	1308	1130	353	4800	0	4800	336	0	336	336	336	0	336
Arabinose	23	0	23	20	6	85	0	85	6	0	6	6	6	0	6
Galactose	157	0	157	135	42	575	0	575	575	0	575	575	575	0	575
Mannose	255	0	255	220	69	934	0	934	934	0	934	934	934	0	934
Cellulose	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
Arabinan	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannan	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactan	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Lignin (Insoluble)	0	0	21259	0	11621	0	0	0	0	0	0	0	0	0	0
Lignin (Soluble)	21259	0	0	0	0	3598	0	3598	3598	0	3598	3598	3598	0	3598
Cellulase	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Zymo Yeast	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DAP	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CSL	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
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Water	146076	2380	149575	129263	9082	215412	172	215583	217219	2208	219870	219870	268237	34800	306923
Furfural	195	0	195	169	26	718	0	718	3625	0	3625	3625	4664	0	2852
Methanol	39	0	39	34	5	144	0	144	144	0	144	144	305	0	128
Acetic Acid	329	0	329	285	44	1209	0	1209	1209	0	0	0	382	0	382
Lactic Acid	0	0	0	0		0	0	0	0	0	0	0	0	0	0
H2SO4	0	3050	4	4	704	0	220	220	220	0	0	0	0	0	0
CO2	0	0	0	0		0	0	0	0	0	0	0	0	0	0
NAOH	2484	0	0	0	0	0	0	0	0	1181	197	197	0	0	0
Sodium Sulphate	0	0	4411	3812	1074	0	0	0	0	0	318	318	318	0	318
Sodium Acetate	0	0	0	0	0	0	0	0	0	0	1651	1651	1651	0	1651
Solid Fraction (Mass)	0.00	0.00	0.12	0.00	0.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total Flow lb/hr	172379	5430	177809	135291	23095	228408	391	228799	228799	3389	232188	232188	281941	34800	318637
Total Flow cuft/hr	2656	74	2778	2205	314	3770480	6067	7697830	508931	115	7275220	4035	4810	103930	5705
Temperature F	80	70	70	70	70	420	415	419	415	77	355	211	187	366	258
Pressure psi			15	15	14.7	30.0	14.7	14.7	221.2	14.7	14.7	14.7	14.7	164.7	34.7

Table C.3.1 (Continued)

Mass Flow lb/hr	**228**	**228A**	**229**	**230**	**232**	**233**	**234**	**236**	**237**	**301**	**302**	**303**	**304**	**305**	**306**	**307**
Glucose	0	0	0	0	0	0	0	0	0	28	0	0	0	0	1718	0
Xylose	0	0	0	0	0	0	0	0	0	145	0	0	0	0	145	0
Arabinose	0	0	0	0	0	0	0	0	0	3	0	0	0	0	3	0
Galactose	0	0	0	0	0	0	0	0	0	17	0	0	0	0	8	0
Mannose	0	0	0	0	0	0	0	0	0	28	0	0	0	0	31	0
Cellulose	0	0	0	0	0	0	0	0	0	40444	0	0	0	0	7053	0
Xylan	0	0	0	0	0	0	0	0	0	1246	0	0	0	0	1246	0
Arabinan	0	0	0	0	0	0	0	0	0	83	0	0	0	0	83	0
Mannan	0	0	0	0	0	0	0	0	0	664	0	0	0	0	88	0
Galactan	0	0	0	0	0	0	0	0	0	166	0	0	0	0	22	0
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	2750	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	45	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	11	0
Lignin (Insoluble)	0	0	0	0	0	0	0	0	0	3668	0	0	0	0	3668	0
Lignin (Soluble)	0	0	0	0	0	0	0	0	0	2362	0	0	0	0	2362	0
Cellulase	0	0	0	0	0	0	0	0	0	0	0	0	12133	0	10313	0
Zymo Yeast	0	0	0	0	0	0	0	0	0	0	315	0	0	0	1086	0
DAP	0	0	0	0	0	0	0	0	0	0	0	191	0	0	179	0
CSL	0	0	0	0	0	0	0	0	0	0	0	1520	0	0	1476	0
Ash	0	0	0	0	0	0	0	0	0	332	0	0	0	0	332	0
Ethanol	0	0	0	0	0	0	0	0	0	0	0	0	0	0	16271	130
Water	25442	25442	333	25405	3920	29327	1	37	296	16231	66938	0	133522	359855	542548	270
Furfural	4006	4006	1945	2193	0	2193	0	1813	132	22	0	0	0	0	22	0
Methanol	483	483	0	483	0	306	177	0	0	4	0	0	0	0	4	0
Acetic Acid	0	0	0	0	0	0	0	0	0	37	0	0	0	0	77	0
Lactic Acid	0	0	0	0	0	0	0	0	0	0	0	0	0	0	54	0
H2SO4	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	757	14903
NAOH	0	0	0	0	0	0	0	0	0	276	0	0	0	0	276	0
Sodium Sulphate	0	0	0	0	0	0	0	0	0	0						
Sodium Acetate	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Solid Fraction (Mass)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.71		0.00	0.08	0.00	0.04	0.00
Total Flow lb/hr	29931	29931	2278	28081	3920	31827	178	1850	428	65756	67253	1711	145655	359855	592629	15308
Total Flow cuft/hr	527	493	33	468	11707	566	4	29	7	911	1717	29	2273	5288	9514	141358
Temperature F	252	150	151	151	366	254	191	283	207	80	68	68	68	68	86	86
Pressure psi	34.7	4.3	35.0	35.0	164.7	34.7	34.7	20.0	15.0	15	15	15	15	15	15	15

C.4: Case 4: Streams Results Summary

Table C.4.1: Case 4 Material Balance

Mass Flow lb/hr	101	201	202	203	204	205	206	206A	207	208	209	210	211	212	213
Glucose	0	0	6453	6453	0	6453	0	0	0	6377	72	0	6481	6451	1
Xylose	0	0	4282	4282	0	4282	0	0	0	4232	48	0	4301	4281	1
Arabinose	0	0	293	293	0	293	0	0	0	290	3	0	295	293	0
Galactose	0	0	352	352	0	352	0	0	0	348	4	0	353	352	0
Mannose	0	0	2053	2053	0	2053	0	0	0	2029	23	0	2062	2053	0
Cellulose	41667	0	34786	34786	0	34786	0	0	0	0	0	0	0	0	0
Xylan	14420	0	9738	9738	0	9738	0	0	0	0	0	0	0	0	0
Arabinan	587	0	235	235	0	235	0	0	0	0	0	0	0	0	0
Mannan	3102	0	939	939	0	939	0	0	0	0	0	0	0	0	0
Galactan	838	0	469	469	0	469	0	0	0	0	0	0	0	0	0
Glucose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Mannose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Galactose Oligomer	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Lignin (Insoluble)	22384	0	3520	3520	0	3520	0	0	0	0	0	0	16380	0	16380
Lignin (Soluble)	0	0	19182	19182	0	19182	0	0	0	18787	28	0	2448	2436	0
Cellulase	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Zymo Yeast	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DAP	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
CSL	0	0	352	352	0	352	0	0	0	352	0	0	352	352	0
Ash	335	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Ethanol	0	200000	199975	106775	93200	66102	40673	40673	0	99560	9207	0	108849	109197	23
Water	83333	216667	299992	259171	40821	238530	20641	20641	618774	541046	216823	84562	782173	897259	16354
Furfural	0	0	176	143	33	115	28	28	0	141	3	0	144	144	0
Methanol	0	0	59	34	24	23	12	12	0	34	1	0	35	35	0
Acetic Acid	0	0	469	444	25	428	16	16	0	438	6	0	446	445	0
Lactic Acid	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
CO2	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
NAOH	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Sodium Sulphate	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
MAQ	0	3344	3344	3344	0	3344	0	0	0	3304	37	0	3358	3343	1
Solid Fraction (Mass)	0.50	0.00	0.08	0.11	0.00	0.13	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.50
Total Flow lb/hr	166666	420011	586667	452564	134103	391194	61369	61369	618774	676938	226254	84562	927677	1026640	32761
Total Flow cuft/hr	2357	4003960	13795	8536	358985	6695	960683	1241	10015	11635	3737	1369	15805	17496	442
Temperature F	70	225	446	289	289	189	189	176	85	145	103	85	140	134	96
Pressure psi	15	26.0	689.3	96.0	96.0	14.7	14.7	14.7	15.0	14.7	15.0	15.0	15.0	15.0	15.0

Table C.4.1 (Continued)

Mass Flow lb/hr	301	302	303	304	305	306	307
Glucose	0	0	0	0	0	1476	0
Xylose	0	0	0	0	0	0	0
Arabinose	0	0	0	0	0	0	0
Galactose	0	0	0	0	0	21	0
Mannose	0	0	0	0	0	42	0
Cellulose	34786	0	0	0	0	6067	0
Xylan	9738	0	0	0	0	9738	0
Arabinan	235	0	0	0	0	235	0
Mannan	939	0	0	0	0	124	0
Galactan	469	0	0	0	0	62	0
Glucose Oligomer	0	0	0	0	0	2365	0
Mannose Oligomer	0	0	0	0	0	64	0
Galactose Oligomer	0	0	0	0	0	32	0
Lignin (Insoluble)	3885	0	0	0	0	3885	0
Lignin (Soluble)	0	0	0	0	0	0	0
Cellulase	0	0	0	12133	0	10313	0
Zymo Yeast	0	341	0	0	0	1017	0
DAP	0	0	211	0	0	200	0
CSL	0	0	1675	0	0	1636	0
Ash	0	0	0	0	0	0	0
Ethanol	61	0	0	0	0	14338	90
Water	49991	72463	0	133522	407649	608713	234
Furfural	0	0	0	0	0	0	0
Methanol	0	0	0	0	0	0	0
Acetic Acid	0	0	0	0	0	35	0
Lactic Acid	0	0	0	0	0	48	0
H2SO4	0	0	0	0	0	0	0
CO2	0	0	0	0	0	829	12889
NAOH	0	0	0	0	0	0	0
Sodium Sulphate	0	0	0	0	0	0	0
MAQ	0	0	0	0	0	0	0
Solid Fraction (Mass)	0.50	0.005	0.00	0.08	0.00	0.05	0.00
Total Flow lb/hr	100106	72804	1886	145655	407649	663060	13217
Total Flow cuft/hr	1468	4	32	2273	6875	10595	122077
Temperature F	95	68	84	68	68	86	86
Pressure psi	15.0	14.7	14.7	14.7	14.7	14.7	14.7

APPENDIX D

INDIVIDUAL EQUIPMENT DESCRIPTION AND COSTS SUMMARY

D.1 Base Case Equipment Description and Costs

Table D.1.1: Base Case Equipment Costs

Tag	Name	Qty	MoC	Description	Eqpt Cost	Installed Cost
R-201	Pretreatment Reactor	1	SS304	Screw Reactor, RT - 4 min	1,090,167	2,496,482
F-201	First Flash Vessel	1	SS304	4840 gal, 50 psig, 6.5'x19.5'	133,700	252,800
F-202	Second Flash Vessel	1	SS304	4716 gal, 0 psig, 6.5'x19'	49,000	216,800
H-201	First Flash Overhead Vapor Condenser	1	CS/SS304	Fixed Shell-Tube, 1168 sf,	51,000	156,200
H-202	Second Flash Overhead Vapor Condenser	1	CS/SS304	Fixed Shell-Tube, 1410 sf,	44,300	146,600
P-201	First Flash Overhead Condensate Pump	1	CS	Centrifugal, 76 gpm, 150' Head	4,300	27,300
P-202	Second Flash Overhead Condensate Pump	1	CS	Centrifugal, 25 gpm, 150' Head	4,100	23,100
C-201	Second Flash Bottoms Conveyor	1	SS304	Screw, 18"x10'	23,100	41,700
A201				**Section Total**	**1,399,667**	**3,360,982**
-						
S-201	Filter Press System for Solids Wash	1	SS316	Pneumapress, 320 gpm 400 sq ft	1,625,083	1,851,893
C-202	Washed Solids Conveyor to Delignification	1	Belt	Belt, 50"x 15'	42,200	58,100
P-203	Soluble Sugars Pump	1	CS	Centrifugal, 45 gpm, 70' Head	20,971	81,158
A202				**Section Total**	**1,688,254**	**1,991,151**
-						
M-201	NaOH Mixing Tank	1	CS	5 min RT, 1285 gal, vertical agitated vessel, 15 hp	36,300	142,100
ST-201	NaOH Storage Tank	1	CS	64000 gal, 3 days RT, 95% wv, Flat Btm Storage	57,738	80,833
P-204	NaOH Pump	1	CS	Centrifugal 14 gpm, 150' head	6,900	19,319
S-202	Filter Press for Lignin Seperation	1	SS316	Pneumapress, 260 gpm,162 sf	951,184	998,743
C-203	Delignified Solids Conveyor	1	Belt	50"x30' Belt	46,400	67,700
P-205	Dissolved Lignin Pump	1	CS	Centrifugal, 90 gpm, 150' head	9,200	49,300
A203				**Section Total**	**1,107,721**	**1,357,995**
ST-202	H2SO4 Storage Tank	2	Plastic	7387 gal, 1 day RT, 90% vv	21,746	30,444
P-206	H2SO4 Pump	2	SS316	4.5 gpm, 245 ft. head	22,045	61,726
M-202	H2SO4 Mixer	1	SS316	Static, 8"x36"	8,200	88,900
S-203	Lignin Seperation Press	1	SS316	Pneumapress, 95 gpm,120 sf	789,626	829,107
C-204	Lignin Byproduct Conveyor	1	CS	Screw, 10"x100'	17,800	40,600
P-207	Waste Water Pump	1	CS	Centrifugal, 76 gpm, 150' Head	4,300	27,300
A204				**Section Total**	**863,717**	**1,078,078**

Table D.1.1 Continued

Tag	Name	Qty	MoC	Description	Eqpt Cost	Installed Cost
T-301	SSF Reactor	6	SS304	Flat-Btm Tanks 1000000 gal each	3,348,003	4,017,603
P-301	SSF Recirculation Pump	6	SS304	Centrifugal, 2135 gpm, 150' Head	126,689	354,729
H-301	SSF Cooler	6	SS304	Plate-Frame, 1450 sf	60,864	127,814
ST-302	Yeast Storage	1	SS304	Flat-Btm, 610000 gal, 95% wv, 3 Days RT	198,630	278,082
ST-303	CSL/DAP Storage	1	SS304	Flat-Btm, 18200 gal, 95% wv, 3 Days RT	31,419	43,986
ST-304	Cellulase Storage	2	SS304	Flat-Btm, 645000 gal ea, 95% wv, 3 Days RT	408,233	571,526
P-302	Yeast Pump	3	CS	Rotary Lobe, 45 gpm, 150' Head	35,743	85,494
P-303	CSL/DAP Pump	1	CS	Centrifugal, 4 gpm, 150' Head	10,636	25,440
P-304	Cellulase Pump	1	CS	Centrifugal, 283 gpm, 150' Head	4,777	6,687
A300				**Section Total**	**4,224,993**	**5,511,361**
-						
				Total Cost	**9,284,352**	**13,299,567**

D.2 Case 2 Additional Equipment Description and Costs

Table D.2.1: Case 2 Equipment Costs

Tag	Name	Qty	MoC	Description	Eqpt Cost	Installed Cost
T-201	Collection Tank for Condensed Flash Overhead Streams	1	SS304	Agitated Tank, 688 Gal	69,000	172,400
D-201	Furfural Stripping Column	1	A515/SS304	41 Sieve Trays, 4'x94'	187,500	419,600
H-205	Furfural Stripping Column Overhead Condenser	1	CS/SS304	Fixed Shell-Tube, 760 sf	33,400	131,600
T-202	Stripping Column Reflux Drum	1	SS	Horizontal Drum, 100 Gal	12,300	88,100
H-206	Heat Exchanger	1	SS	Double Pipe, 63 sf	19,400	58,400
T-203	Decanter	1	SS	Decanter, 635 gal	22,100	119,100
H-207	Methanol Column Feed Heater	1	SS	Double Pipe, 47 sf	13,300	51,800
D-202	Methanol Stripping Column	1		13 Trays, 1,5'x13'	36,700	171,400
H-208	Methanol Column Overhead Condenser	1	SS	Fixed Shell-Tube, 294 sf	22,900	100,400
T-204	Methanol Column Reflux Drum	1	SS	Horizontal Drum, 50 Gal	5,400	80,400
H-210	Dehydration Column Overhead Condenser	1	SS	Double Pipe, 10 sf	5,300	38,100
D-203	Furfural Dehydration Column	1	SS304	Packed, 1.5'x26'	38,800	130,900
H-209	Dehydration Column Reboiler	1	SS	Double Pipe, 20 sf	2,500	40,400
P-211	Stripping Column Feed Pump	1	SS	Centrifugal, 108 gpm, 150' Head	5,700	39,100
P-212	Stripping Column Overhead Pump	1	SS	Centrifugal, 27 gpm, 150' Head	5,000	29,200
P-213	Stripping Column Reflux Pump	1	SS	Centrifugal, 12 gpm, 150' Head	4,400	28,500
P-214	Stripping Column Bottoms Pump	1	SS	Centrifugal, 130 gpm, 150' Head	5,800	39,200
P-215	Furfural Dehydration Column Feed Pump	1	SS	Centrifugal, 2 gpm, 150' Head	4,400	23,400
P-216	Methanol Recovery Column Bottoms Pump	1	SS	Centrifugal, 18 gpm, 150' Head	4,400	28,500
P-217	Methanol Recovery Column Reflux Pump	1	SS	Centrifugal, 8 gpm, 150' Head	4,400	24,500
P-218	Methanol Recovery Column Feed Pump	1	SS	Centrifugal, 15 gpm, 150' Head	4,400	28,600
P-219	Furfural Product Pump	1	SS	Centrifugal, 2 gpm, 150' Head	4,400	23,400
A206				**Section Total**	**511,500**	**1,867,000**
-						
				Other Sections	**9,284,352**	**13,299,567**
				Total Cost	**9,795,852**	**15,166,567**

D.3 Case 3 -Additional Equipment Description and Costs

Table D.3.1: Case 3 Equipment Costs

Tag	Name	Qty	MoC	Description	Eqpt Cost	Installed Cost
H-203	Heat Interchange	2	CS	Floating Head, 6500 sf each	216,600	468,400
H-204	Soluble Sugars Heat Exchanger	1	CS	Floating Head, 1760 sf	41,700	126,500
M-203	Acid Mixer	1	SS316	Static Mixer, 6"x24", 12 elements	10,400	77,600
R-202	Reactor	1	Hastelloy Clad	Plug Flow, 3'x60', 15 Trays	275,400	701,000
P-209	Neutralized Product Pump	2	CS	Centrifugal, 250 gpm, 150' Head	11,600	68,100
M-204	Neutralizing Tank	1	SS316	Static Mixer, 6"x24", 12 elements	10,400	77,600
A205				**Section Total**	**719,834**	**1,785,188**
T-201	Collection Tank for Condensed Flash Overhead Streams	1	CS	6000 Gal Vessel	27,900	161,800
D-201	Furfural Stripping Column	1	A515/SS304	52 Trays, 7'x116'	303,800	562,700
H-205	Furfural Stripping Column Overhead Condenser	1	CS/SS304	Fixed Shell-Tube, 3800 sf	16,900	68,400
T-202	Stripping Column Reflux Drum	1	SS	Horizontal Drum, 750 Gal	12,300	76,900
H-206	Heat Exchanger	1	SS	Fixed Shell-Tube, 153 sf	14,700	58,700
T-203	Decanter	1	SS	Decanter, 1250 gal	16,100	105,500
H-207	Methanol Column Feed Heater	1	SS	Fixed Shell-Tube, 120 sf	14,900	59,000
D-202	Methanol Stripping Column	1	A515/SS304	13 Trays, 2'x38'	32,100	148,700
H-208	Methanol Column Overhead Condenser	1	SS	Fixed Shell-Tube, 944 sf	15,500	66,200
T-204	Methanol Column Reflux Drum	1	SS	Horizontal Drum, 475 Gal	10,900	73,400
H-210	Dehydration Column Overhead Condenser	1	SS	Double Pipe, 20 sf	2,400	40,300
D-203	Furfural Dehydration Column	1	SS304	Packed, 2'x26', Packing Ht 12'	28,500	106,400
H-209	Dehydration Column Reboiler	1	SS	Kettle Type, 38 sf	34,400	77,300
P-211	Stripping Column Feed Pump	3	SS	Centrifugal, 190 gpm ea, 150' Head	16,500	96,800
P-212	Stripping Column Overhead Pump	1	SS	Centrifugal, 112 gpm, 150' Head	4,700	28,400
P-214	Stripping Column Bottoms Pump	3	SS	Centrifugal, 225 gpm, 150' Head	17,100	101,800
P-215	Furfural Dehydration Column Feed Pump	1	SS	Centrifugal, 4 gpm, 150' Head	4,300	27,200
P-216	Methanol Recovery Column Bottoms Pump	1	SS	Centrifugal, 70 gpm, 150' Head	4,300	27,300
P-217	Methanol Recovery Column Reflux Pump	1	SS	Centrifugal, 25 gpm, 150' Head	4,100	23,100
P-218	Methanol Recovery Column Feed Pump	1	SS	Centrifugal, 60 gpm, 150' Head	4,200	24,200
P-219	Furfural Product Pump	1	SS	Centrifugal, 5 gpm, 150' Head	3,500	20,300
A206				**Section Total**	**593,300**	**1,978,600**
				Other Sections	**9,284,352**	**13,299,567**
				Total Cost	**10,597,486**	**17,063,356**

D.4: Case 4 Additional Equipment Description and Costs

Table D.4.1: Case 4 Equipment Costs

Tag	Name	Qty	MoC	Description	Eqpt Cost	Installed Cost
C-201	Pretreated Hydrolyzate Conveyor	1	SS304	Screw, 16"x33'	44,100	63,400
F-201	First Flash Vessel	1	SS304	7600 Gal Liq Vol, 80 psig, 6'x17.2'	283,700	445,500
F-202	Second Flash Vessel	1	SS304	6190 Gal Liq Vol, 0 psig, 7'x21.5'	65,300	215,900
H-201A	Ethanol-Water Heat Exchanger	1	A285C/SS304	Fixed Shell-Tube, 4610 sf	165,800	384,400
H-201B	Ethanol-Water Vaporizer	1	A285C/SS304	Floating Shell-Tube, 796 sf	37,200	129,200
H-202	First Flash Vapors Condenser	1	A285C/SS304	Floating Shell-Tube, 4020 sf	116,200	251,000
H-203	Second Flash Vapors Condenser	1	A285C/SS304	Floating Shell-Tube, 2545 sf	81,000	211,200
P-200	MAQ Pump	1	SS	Centrifugal, 5 gpm, 150' Head	3,500	20,300
P-201	First Flash Condensed Vapors Pump	1	SS	Centrifugal, 340 gpm, 150' Head	8,300	51,500
P-202	Second Flash Condensed Vapors Pump	1	SS	Centrifugal, 154 gpm, 150' Head	6,300	40,000
R-201	Pretreatment Reactor	1	SS316	Screw Reactor	1,090,167	2,496,482
ST-200	MAQ Storage	1	SS304	Flat-Btm Storage, 25000 Gal, 3 Days RT, 90%wv	25,012	35,017
A201				**Section Total**	**1,901,567**	**4,308,882**
C-202	Pulp Conveyor	1	CS	Belt, 25"x60'	38,900	68,600
P-203	Filtrate Pump	3	CS	Centrifugal, 480 gpm, 150' Head	23,400	127,800
P-204	Wash Water Pump	2	CS	Centrifugal, 170 gpm, 150' Head	10,200	58,300
S-201	Solids Wash	2	SS316	Pneumapress System,300 gpm, 375 sq ft	2,655,737	3,041,614
A202				**Section Total**	**2,728,237**	**3,296,314**
C-204	Lignin Conveyor	1	CS	Screw, 7" Dia x 33' Long	7,300	20,700
M-202	Precipitation Tank	1	SS304	Agitated Tank, 12735 gal, 8.5' Dia x 30' Ht	211,400	344,800
P-207	Ethanol Filtrate Pump	5	SS	Centrifugal, 430 gpm, 150' Head	46,500	309,100
S-203	Lignin Seperation Press	1	SS316	Pneumapress, 132 gpm, 170 sq ft	969,718	1,018,204
A204				**Section Total**	**1,234,918**	**1,692,804**
				A300	**4,224,993**	**5,511,361**
				Total Cost	**10,089,714**	**14,809,362**

D.5: Enzyme Recycling A300 Equipment Description and Costs

Table D.5.1: Enzyme Recycling with 15% Deactivation and 15% Purge: A300 Equipment Costs

Tag	Name	Qty	MoC	Description	Eqpt Cost	Installed Cost
A-301	SSF Agitators	20	SS304	Fixed Prop, Side Mounted, 95 HP	243,065	291,678
C-301	First Centrifuge Cake Conveyor	1	SS316	Screw, 28"x45'	91,100	118,430
C-302	Second Centrifuge Cake Conveyor	1	SS316	Screw, 25"x45'	79,539	103,401
C-303	Purge Conveyor	1	SS316	Screw, 18"x30'	23,546	30,610
H-301	SSF Heat Exchanger	10	SS304	Plate-Frame, 3470 sf, 300 BTU/sf-hr	182,813	383,906
P-301	SSF Recirculation and Transfer Pumps	10	SS304	Centrifugal, 1525 gpm, 150' Head	229,532	642,689
P-302	First Centrate Pump	1	SS304	Centrifugal, 1700 gpm, 150' Head	20,730	58,044
P-303	Second Centrate Pump	1	SS304	Centrifugal, 1875 gpm, 150' Head	16,675	46,690
P-304	Mixing Tank Transfer Pump	1	SS304	Centrifugal, 1300 gpm, 150' Head	22,498	62,995
S-301	SSF Product Centrifuge	4	SS304	Centrifuge, 50"x20"	3,640,497	4,368,596
S-302	SSF Feed Centrifuge	3	SS304	Centrifuge, 50"x20"	2,718,104	3,261,725
T-301	SSF Tank	10	SS304	Flat-Btm Storage, 1.2MM Gal, RT 7.2 Hrs	5,972,994	7,167,593
T-302	Centrate Cellulase Recovery Tank	1	SS304	Flat-Btm Storage, 128000 Gal, RT 1 Hr	1,378,573	2,026,502
				Total	14,619,667	18,562,861

APPENDIX E

COST OF CHEMICALS

Table E.1: Raw Materials' and Utilities' Costs

Raw Material	Cost	Cost Basis	Cost Year	Source
Hardwood chips	55.00	\$/ODT	2002	Vendor Quote
NaOH	267.30	\$/ton	2006	Reporter, 2002
H2SO4	66	\$/ton	2006	Reporter, 2002
Lignin	83.06	\$/ton	2003	Wingren, Galbe et al., 2003
Furfural	0.50	\$/lb	2005	Vendor Quote
Methanol	1.14	\$/gal	2006	Reporter, 2002
Corn Steep Liquor	0.08	\$/lb	2002	Aden, Ruth et al., 2002
Cellulase Enzyme	0.05	\$/lb	2002	Wingren, Galbe et al., 2003
Diammonium Phosphate	0.07	\$/lb	2002	Aden, Ruth et al., 2002
Zymo Yeast	0.17	\$/lb	2002	Wingren, Galbe et al., 2003
25 psig Steam	3.24	\$/ton	2002	Peter and Timmerhaus, 2002
150 psig Steam	3.51	\$/ton	2002	Peter and Timmerhaus, 2002
300 psig Steam	3.83	\$/ton	2002	Peter and Timmerhaus, 2002
600 psig Steam	4.48	\$/ton	2002	Peter and Timmerhaus, 2002
Cooling Water	0.14	\$/m3	2002	Peter and Timmerhaus, 2002
Chilled Water	1.00	\$/ton-day	2002	Seider, Seader et al., 2003
Process Water	0.31	\$/m3	2002	Peter and Timmerhaus, 2002
Electricity	0.06	\$/KW	2002	Peter and Timmerhaus, 2002

APPENDIX F

ENZYME RECYCLING RESULTS

Table F.1.1: Single Step Deactivation Scheme Raw Material Usage

Single Step Deactivation Scheme	No Recycle	0 % Activity Loss					
Purge %		5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40,444	47,511	47,030	46,558	46,095	45,641	43,501
Cellulase In (lb/hr)	12,133	14,252	14,109	13,967	13,828	13,692	13,050
Fresh (lb/hr)	12,133	734	1,416	2,087	2,744	3,387	6,426
Recycled (lb/hr)	-	13,519	12,693	11,881	11,085	10,305	6,624
Decellulase In (lb/hr)	-	-	-	-	-	-	-
Tot Cellulase In (lb/hr)	12,133	14,252	14,109	13,967	13,828	13,692	13,050
Cellulase Out (lb/hr)	10,313	14,252	14,109	13,967	13,828	13,692	13,050
Decellulase Out (lb/hr)	1,820	-	-	-	-	-	-
Total Cellulase Out (lb/hr)	12,133	14,252	14,109	13,967	13,828	13,692	13,050
Avg. % Activity Loss ((In-Out)/(In))				0%			
Fresh DAP Added (lb/hr)	209	419	259	210	193	180	152
Fresh CSL Added (lb/hr)	1,659	3,296	2,024	1,636	1,503	1,405	1,188
DAP Recycled (lb/hr)	-	353	201	151	131	115	71
CSL Recycled (lb/hr)	-	2,842	1,631	1,237	1,070	943	588
Total DAP In (lb/hr)	209	772	459	361	323	295	223
Total CSL In (lb/hr)	1,659	6,138	3,655	2,872	2,573	2,348	1,775
Total DAP Out (lb/hr)	196	755	443	344	307	279	208
Total CSL Out (lb/hr)	1,613	6,078	3,596	2,813	2,515	2,290	1,720
Fresh Zymo	330	-	-	-	-	-	-
Ethanol wt% into SSF	0.0%	1.3%	1.9%	2.2%	2.3%	2.4%	2.3%
Ethanol Produced (lb/hr)	16,973	20,074	19,862	19,653	19,449	19,249	18,305
Ethanol in Purge (lb/hr)	-	526	1,010	1,460	1,878	2,267	3,850
Ethanol Production (MM Gal. / Yr)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.1.1 (Continued)

Single Step Deactivation Scheme	No Recycle	5 % Activity Loss					
Purge %		5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40,444	47,511	47,030	46,558	46,095	45,641	43,501
Cellulase In (lb/hr)	12,133	14,253	14,109	13,967	13,828	13,692	13,050
Fresh (lb/hr)	12,133	1,409	2,051	2,681	3,298	3,903	6,757
Recycled (lb/hr)	-	12,845	12,058	11,287	10,531	9,790	6,293
Decellulase In (lb/hr)	-	13,172	6,315	3,976	2,794	2,083	673
Tot Cellulase In (lb/hr)	12,133	27,425	20,424	17,943	16,622	15,775	13,723
Cellulase Out (lb/hr)	10,313	13,541	13,403	13,269	13,137	13,008	12,398
Decellulase Out (lb/hr)	1,820	13,884	7,021	4,674	3,485	2,767	1,325
Total Cellulase Out (lb/hr)	12,133	27,425	20,424	17,943	16,622	15,775	13,723
Avg. % Activity Loss ((In-Out)/(In))				5%			
Fresh DAP Added (lb/hr)	209	404	259	214	195	183	153
Fresh CSL Added (lb/hr)	1,659	3,177	2,024	1,672	1,520	1,427	1,195
DAP Recycled (lb/hr)	-	340	201	155	132	117	71
CSL Recycled (lb/hr)	-	2,738	1,631	1,265	1,083	958	591
Total DAP In (lb/hr)	209	743	460	369	327	300	225
Total CSL In (lb/hr)	1,659	5,915	3,656	2,938	2,603	2,385	1,787
Total DAP Out (lb/hr)	196	727	443	353	311	283	209
Total CSL Out (lb/hr)	1,613	5,854	3,596	2,879	2,545	2,327	1,732
Fresh Zymo	330	-	-	-	-	-	-
Ethanol wt% into SSF	0.0%	1.3%	1.9%	2.2%	2.3%	2.4%	2.3%
Ethanol Produced (lb/hr)	16,973	20,074	19,862	19,653	19,449	19,249	18,305
Ethanol in Purge (lb/hr)	-	525	1,011	1,460	1,878	2,268	3,850
Ethanol Production (MM Gal. / Yr)	22.5	25.54	25.27	25.01	24.75	24.49	23.29

Table F.1.1 (Continued)

Single Step Deactivation Scheme	No Recycle	10 % Activity Loss					
Purge %		5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40,444	47,511	47,030	46,558	46,095	45,641	43,501
Cellulase In (lb/hr)	12,133	14,253	14,109	13,967	13,828	13,692	13,050
Fresh (lb/hr)	12,133	2,085	2,686	3,275	3,852	4,418	7,088
Recycled (lb/hr)	-	12,169	11,423	10,693	9,976	9,274	5,962
Decellulase In (lb/hr)	-	26,267	12,635	7,952	5,587	4,165	1,345
Tot Cellulase In (lb/hr)	12,133	40,520	26,744	21,919	19,416	17,858	14,396
Cellulase Out (lb/hr)	10,313	12,828	12,698	12,571	12,446	12,323	11,745
Decellulase Out (lb/hr)	1,820	27,692	14,046	9,349	6,970	5,535	2,650
Total Cellulase Out (lb/hr)	12,133	40,520	26,744	21,919	19,416	17,858	14,396
Avg. % Activity Loss ((In-Out)/(In))				10%			
Fresh DAP Added (lb/hr)	209	404	259	219	197	186	154
Fresh CSL Added (lb/hr)	1,659	3,177	2,024	1,707	1,537	1,448	1,203
DAP Recycled (lb/hr)	-	340	201	158	134	118	72
CSL Recycled (lb/hr)	-	2,738	1,631	1,292	1,096	972	595
Total DAP In (lb/hr)	209	744	459	377	331	304	226
Total CSL In (lb/hr)	1,659	5,915	3,655	2,999	2,633	2,420	1,798
Total DAP Out (lb/hr)	196	727	443	360	315	288	211
Total CSL Out (lb/hr)	1,613	5,855	3,596	2,940	2,575	2,362	1,743
Fresh Zymo	330	-	-	-	-	-	-
Ethanol wt% into SSF	0.0%	1.3%	1.9%	2.1%	2.3%	2.3%	2.3%
Ethanol Produced (lb/hr)	16973	20074	19861	19653	19449	19249	18305
Ethanol in Purge (lb/hr)	0	525	1010	1460	1878	2268	3850
Ethanol Production (MM Gal. / Yr)	22.5	25.54	25.27	25.01	24.75	24.49	23.29

Table F.1.1 (Continued)

Single Step Deactivation Scheme	No Recycle	15 % Activity Loss					
Purge %		5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40,444	47,511	47,030	46,558	46,095	45,641	43,501
Cellulase In (lb/hr)	12,133	14,253	14,109	13,967	13,828	13,692	13,050
Fresh (lb/hr)	12,133	2,761	3,320	3,869	4,406	4,933	7,420
Recycled (lb/hr)	-	11,493	10,789	10,098	9,422	8,759	5,631
Decellulase In (lb/hr)	-	39,455	18,963	11,928	8,381	6,248	2,018
Tot Cellulase In (lb/hr)	12,133	53,709	33,072	25,896	22,209	19,940	15,068
Cellulase Out (lb/hr)	10,313	12,115	11,993	11,872	11,754	11,639	11,093
Decellulase Out (lb/hr)	1,820	41,593	21,079	14,023	10,455	8,302	3,976
Total Cellulase Out (lb/hr)	12,133	53,709	33,072	25,896	22,209	19,940	15,068
Avg. % Activity Loss ((In-Out)/(In))				15%			
Fresh DAP Added (lb/hr)	209	468	285	223	199	192	155
Fresh CSL Added (lb/hr)	1,659	3,686	2,232	1,739	1,554	1,498	1,210
DAP Recycled (lb/hr)	-	396	223	162	135	123	72
CSL Recycled (lb/hr)	-	3,184	1,804	1,318	1,108	1,007	599
Total DAP In (lb/hr)	209	864	507	384	335	315	227
Total CSL In (lb/hr)	1,659	6,870	4,036	3,057	2,662	2,505	1,809
Total DAP Out (lb/hr)	196	847	491	368	318	299	212
Total CSL Out (lb/hr)	1,613	6,810	3,976	2,998	2,603	2,447	1,754
Fresh Zymo	330	-	-	-	-	-	-
Ethanol wt% into SSF	0.0%	1.1%	1.7%	2.1%	2.3%	2.3%	2.3%
Ethanol Produced (lb/hr)	16973	20074	19861	19653	19449	19249	18305
Ethanol in Purge (lb/hr)	0	525	1011	1460	1878	2271	3850
Ethanol Production (MM Gal. / Yr)	22.5	25.54	25.27	25.01	24.75	24.49	23.29

Table F.1.1 (Continued)

Single Step Deactivation Scheme	No Recycle	25 % Activity Loss					
Purge %		5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40,444	47,511	47,030	46,558	46,095	45,641	43,501
Cellulase In (lb/hr)	12,133	14,253	14,109	13,967	13,828	13,692	13,050
Fresh (lb/hr)	12,133	4,113	4,590	5,057	5,515	5,964	8,082
Recycled (lb/hr)	-	10,141	9,519	8,910	8,314	7,729	4,968
Decellulase In (lb/hr)	-	65,761	31,604	19,880	13,968	10,413	3,363
Tot Cellulase In (lb/hr)	12,133	80,014	45,713	33,848	27,796	24,106	16,414
Cellulase Out (lb/hr)	10,313	10,690	10,582	10,475	10,371	10,269	9,788
Decellulase Out (lb/hr)	1,820	69,324	35,132	23,372	17,425	13,836	6,626
Total Cellulase Out (lb/hr)	12,133	80,014	45,713	33,848	27,796	24,106	16,414
Avg. % Activity Loss ((In-Out)/(In))				25%			
Fresh DAP Added (lb/hr)	209	474	297	241	212	196	172
Fresh CSL Added (lb/hr)	1,659	3,739	2,333	1,888	1,656	1,534	1,348
DAP Recycled (lb/hr)	-	402	233	176	145	126	81
CSL Recycled (lb/hr)	-	3,231	1,887	1,435	1,184	1,032	671
Total DAP In (lb/hr)	209	876	530	418	357	323	254
Total CSL In (lb/hr)	1,659	6,970	4,220	3,322	2,839	2,566	2,019
Total DAP Out (lb/hr)	196	859	514	401	340	306	238
Total CSL Out (lb/hr)	1,613	6,910	4,160	3,263	2,781	2,508	1,963
Fresh Zymo	330	-	-	-	-	-	-
Ethanol wt% into SSF	0.0%	1.1%	1.7%	2.0%	2.1%	2.2%	2.1%
Ethanol Produced (lb/hr)	16973	20074	19861	19653	19449	19249	18305
Ethanol in Purge (lb/hr)	0	525	1011	1461	1880	2269	3854
Ethanol Production (MM Gal. / Yr)	22.5	25.54	25.27	25.01	24.75	24.49	23.29

Table F.1.1 (Continued)

Single Step Deactivation Scheme	No Recycle	50 % Activity Loss					
Purge %		5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40,444	47,511	47,030	46,558	46,095	45,641	43,501
Cellulase In (lb/hr)	12,133	14,253	14,109	13,967	13,828	13,692	13,050
Fresh (lb/hr)	12,133	7,493	7,763	8,027	8,286	8,540	9,738
Recycled (lb/hr)	-	6,760	6,346	5,940	5,542	5,152	3,312
Decellulase In (lb/hr)	-	131,525	63,209	39,761	27,936	20,826	6,727
Tot Cellulase In (lb/hr)	12,133	145,778	77,318	53,728	41,764	34,519	19,777
Cellulase Out (lb/hr)	10,313	7,127	7,054	6,984	6,914	6,846	6,525
Decellulase Out (lb/hr)	1,820	138,652	70,263	46,745	34,850	27,673	13,252
Total Cellulase Out (lb/hr)	12,133	145,779	77,318	53,728	41,764	34,519	19,777
Avg. % Activity Loss ((In-Out)/(In))				50%			
Fresh DAP Added (lb/hr)	209	612	372	289	239	222	177
Fresh CSL Added (lb/hr)	1,659	4,831	2,928	2,265	1,869	1,734	1,383
DAP Recycled (lb/hr)	-	522	295	214	165	144	84
CSL Recycled (lb/hr)	-	4,190	2,381	1,731	1,342	1,172	689
Total DAP In (lb/hr)	209	1,134	667	502	404	365	260
Total CSL In (lb/hr)	1,659	9,021	5,309	3,996	3,211	2,906	2,072
Total DAP Out (lb/hr)	196	1,117	651	486	387	349	245
Total CSL Out (lb/hr)	1,613	8,961	5,250	3,937	3,153	2,848	2,017
Fresh Zymo	330	-	-	-	-	-	-
Ethanol wt% into SSF	0.0%	0.9%	1.4%	1.7%	2.0%	2.1%	2.1%
Ethanol Produced (lb/hr)	16973	20074	19861	19653	19449	19249	18305
Ethanol in Purge (lb/hr)	0	526	1012	1463	1881	2271	3856
Ethanol Production (MM Gal. / Yr)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.1.2: Single Step Deactivation Scheme – Economics Summary

Single Step Deactivation Scheme	No Recycle	0 % Activity Loss					
Purge %		5	10	15	20	25	50
Capital Costs ($/yr)							
Total Installed Equipment Cost ($/yr)	5,200,397	28,490,000	20,080,000	17,010,000	15,730,000	14,930,000	12,400,000
Added Costs ($/yr)	160,000	850,000	600,000	510,000	470,000	450,000	380,000
Total Project Investment ($/yr)	5,360,000	29,340,000	20,680,000	17,520,000	16,200,000	15,380,000	12,780,000
Operating Costs (cents/gal)							
CSL	5.0	8.6	5.3	4.4	4.0	3.8	3.4
Cellulase	24.5	1.3	2.5	3.8	5.0	6.3	12.5
DAP	0.6	1.0	0.6	0.5	0.5	0.4	0.4
Process Water	1.1	5.5	3.2	2.4	2.2	2.0	1.5
Zymo	1.8	0.0	0.0	0.0	0.0	0.0	0.0
Cooling Water	1.8	3.1	1.9	1.5	1.4	1.3	1.0
Chilled Water	0.0	6.9	4.3	3.5	3.2	3.1	2.5
Electricity	1.4	10.4	6.5	5.2	4.7	4.4	3.6
Capital Depreciation	1.2	5.8	4.1	3.5	3.3	3.1	2.7
Operating Costs ($/yr)							
CSL	1,120,000	2,200,000	1,350,000	1,090,000	1,000,000	940,000	790,000
Cellulase	5,520,000	330,000	640,000	950,000	1,250,000	1,540,000	2,920,000
DAP	120,000	250,000	150,000	120,000	110,000	110,000	90,000
Process Water	260,000	1,400,000	800,000	610,000	530,000	490,000	350,000
Zymo	1,850,000	-	-	-	-	-	-
Cooling Water	440,000	800,000	480,000	380,000	340,000	310,000	240,000
Chilled Water	-	1,760,000	1,100,000	880,000	800,000	750,000	590,000
Electricity	340,000	2,660,000	1,630,000	1,290,000	1,160,000	1,080,000	840,000
Capital Depreciation	260,000	1,470,000	1,030,000	880,000	810,000	770,000	640,000
MESP4 ($/gal)	0.41	0.57	0.38	0.33	0.32	0.32	0.35
Change in MESP Over No Recycle	-	0.00	0.19	0.24	0.25	0.25	0.23
Max Savings ($/gal)		0.25					
Purge %		25					

Table F.1.2 (Continued)

Single Step Deactivation Scheme	No Recycle	5 % Activity Loss					
Purge %		5	10	15	20	25	50
Capital Costs ($/yr)							
Total Installed Equipment Cost ($/yr)	5,200,397	27,790,000	20,240,000	17,330,000	15,860,000	15,140,000	12,470,000
Indirect Costs ($/yr)	160,000	830,000	600,000	520,000	470,000	450,000	380,000
Total Project Investment ($/yr)	5,360,000	28,620,000	20,840,000	17,850,000	16,330,000	15,590,000	12,850,000
Operating Costs (cents/gal)							
CSL	5.0	8.3	5.3	4.5	4.1	3.9	3.4
Cellulase	24.5	2.5	3.7	4.9	6.1	7.2	13.2
DAP	0.6	0.9	0.6	0.5	0.5	0.4	0.4
Process Water	1.1	5.2	3.2	2.5	2.2	2.0	1.5
Zymo	1.8	0.0	0.0	0.0	0.0	0.0	0.0
Cooling Water	1.8	3.0	1.9	1.6	1.4	1.3	1.0
Chilled Water	0.0	6.6	4.4	3.6	3.3	3.1	2.5
Electricity	1.4	10.1	6.5	5.3	4.7	4.5	3.6
Capital Depreciation	1.2	5.6	4.1	3.6	3.3	3.2	2.7
Operating Costs ($/yr)							
CSL	1,120,000	2,120,000	1,350,000	1,120,000	1,010,000	950,000	800,000
Cellulase	5,520,000	640,000	930,000	1,220,000	1,500,000	1,780,000	3,070,000
DAP	120,000	240,000	150,000	130,000	120,000	110,000	90,000
Process Water	260,000	1,330,000	800,000	620,000	530,000	500,000	350,000
Zymo	1,850,000	-	-	-	-	-	-
Cooling Water	440,000	780,000	490,000	390,000	340,000	320,000	240,000
Chilled Water	-	1,690,000	1,110,000	900,000	810,000	760,000	590,000
Electricity	340,000	2,570,000	1,650,000	1,330,000	1,170,000	1,100,000	840,000
Capital Depreciation	260,000	1,430,000	1,040,000	890,000	820,000	780,000	640,000
MESP4 ($/gal)	0.41	0.56	0.40	0.35	0.34	0.33	0.35
Change in MESP Over No Recycle	-	0.01	0.17	0.22	0.24	0.24	0.22
Max Savings ($/gal)	-	0.24					
Purge %		25					

Table F.1.2 (Continued)

Single Step Deactivation Scheme	No Recycle	10 % Activity Loss					
Purge %		5	10	15	20	25	50
Capital Costs ($/yr)							
Total Installed Equipment Cost ($/yr)	5,200,397	27,970,000	20,390,000	17,640,000	15,990,000	15,340,000	12,540,000
Indirect Costs ($/yr)	160,000	840,000	620,000	530,000	480,000	460,000	380,000
Total Project Investment ($/yr)	5,360,000	28,810,000	21,010,000	18,170,000	16,470,000	15,800,000	12,920,000
Operating Costs (cents/gal)							
CSL	5.0	8.3	5.3	4.6	4.1	3.9	3.4
Cellulase	24.5	3.7	4.8	6.0	7.1	8.2	13.8
DAP	0.6	0.9	0.6	0.5	0.5	0.4	0.4
Process Water	1.1	5.2	3.2	2.5	2.2	2.0	1.5
Zymo	1.8	0.0	0.0	0.0	0.0	0.0	0.0
Cooling Water	1.8	3.1	2.0	1.6	1.4	1.3	1.0
Chilled Water	0.0	6.7	4.4	3.7	3.3	3.2	2.6
Electricity	1.4	10.2	6.6	5.5	4.8	4.6	3.6
Capital Depreciation	1.2	5.6	4.2	3.6	3.3	3.2	2.8
Operating Costs ($/yr)							
CSL	1,120,000	2,120,000	1,350,000	1,140,000	1,030,000	970,000	800,000
Cellulase	5,520,000	950,000	1,220,000	1,490,000	1,750,000	2,010,000	3,220,000
DAP	120,000	240,000	150,000	130,000	120,000	110,000	90,000
Process Water	260,000	1,330,000	800,000	630,000	530,000	500,000	350,000
Zymo	1,850,000	-	-	-	-	-	-
Cooling Water	440,000	790,000	500,000	400,000	350,000	330,000	240,000
Chilled Water	-	1,710,000	1,120,000	930,000	810,000	770,000	600,000
Electricity	340,000	2,600,000	1,670,000	1,360,000	1,190,000	1,120,000	850,000
Capital Depreciation	260,000	1,440,000	1,050,000	910,000	820,000	790,000	650,000
MESP4 ($/gal)	0.41	0.58	0.41	0.37	0.35	0.35	0.36
Change in MESP Over No Recycle	-	0.00	0.16	0.20	0.22	0.22	0.21
Max Savings ($/gal)	-	0.22					
Purge %		20					

Table F.1.2 (Continued)

Single Step Deactivation Scheme	No Recycle	15 % Activity Loss					
Purge %		5	10	15	20	25	50
Capital Costs ($/yr)							
Total Installed Equipment Cost ($/yr)	5,200,397	31,360,000	21,470,000	17,950,000	16,120,000	15,690,000	12,610,000
Indirect Costs ($/yr)	160,000	940,000	640,000	540,000	480,000	470,000	380,000
Total Project Investment ($/yr)	5,360,000	32,300,000	22,110,000	18,490,000	16,600,000	16,160,000	12,990,000
Operating Costs (cents/gal)							
CSL	5.0	9.6	5.9	4.6	4.2	4.1	3.5
Cellulase	24.5	4.9	6.0	7.0	8.1	9.2	14.5
DAP	0.6	1.1	0.7	0.5	0.5	0.5	0.4
Process Water	1.1	6.2	3.4	2.6	2.2	2.1	1.5
Zymo	1.8	0.0	0.0	0.0	0.0	0.0	0.0
Cooling Water	1.8	3.6	2.1	1.6	1.4	1.4	1.0
Chilled Water	0.0	7.8	4.7	3.8	3.3	3.3	2.6
Electricity	1.4	11.9	7.1	5.6	4.9	4.7	3.7
Capital Depreciation	1.2	6.3	4.4	3.7	3.4	3.3	2.8
Operating Costs ($/yr)							
CSL	1,120,000	2,460,000	1,490,000	1,160,000	1,040,000	1,000,000	810,000
Cellulase	5,520,000	1,260,000	1,510,000	1,760,000	2,000,000	2,240,000	3,380,000
DAP	120,000	280,000	170,000	130,000	120,000	110,000	90,000
Process Water	260,000	1,600,000	870,000	640,000	530,000	510,000	350,000
Zymo	1,850,000	-	-	-	-	-	-
Cooling Water	440,000	920,000	530,000	410,000	350,000	340,000	240,000
Chilled Water	-	2,000,000	1,200,000	950,000	820,000	800,000	600,000
Electricity	340,000	3,040,000	1,790,000	1,400,000	1,200,000	1,160,000	860,000
Capital Depreciation	260,000	1,620,000	1,110,000	920,000	830,000	810,000	650,000
MESP4 ($/gal)	0.41	0.67	0.45	0.39	0.36	0.37	0.37
Change in MESP Over No Recycle	-	-0.10	0.12	0.18	0.21	0.20	0.20
Max Savings ($/gal)	-	0.21					
Purge %		20					

Table F.1.2 (Continued)

Single Step Deactivation Scheme	No Recycle	25 % Activity Loss					
Purge %		5	10	15	20	25	50
Capital Costs ($/yr)							
Total Installed Equipment Cost ($/yr)	5,200,397	31,460,000	22,390,000	18,640,000	17,100,000	16,080,000	13,590,000
Indirect Costs ($/yr)	160,000	940,000	670,000	560,000	510,000	480,000	410,000
Total Project Investment ($/yr)	5,360,000	32,400,000	23,060,000	19,200,000	17,610,000	16,560,000	14,000,000
Operating Costs (cents/gal)							
CSL	5.0	9.8	6.2	5.0	4.5	4.2	3.9
Cellulase	24.5	7.3	8.3	9.2	10.1	11.1	15.8
DAP	0.6	1.1	0.7	0.6	0.5	0.5	0.4
Process Water	1.1	6.2	3.6	2.7	2.4	2.2	1.8
Zymo	1.8	0.0	0.0	0.0	0.0	0.0	0.0
Cooling Water	1.8	3.6	2.2	1.7	1.5	1.4	1.2
Chilled Water	0.0	7.8	5.0	4.0	3.6	3.4	2.8
Electricity	1.4	12.0	7.5	5.9	5.3	4.9	4.1
Capital Depreciation	1.2	6.3	4.6	3.8	3.6	3.4	3.0
Operating Costs ($/yr)							
CSL	1,120,000	2,490,000	1,560,000	1,260,000	1,100,000	1,020,000	900,000
Cellulase	5,520,000	1,870,000	2,090,000	2,300,000	2,510,000	2,710,000	3,680,000
DAP	120,000	280,000	180,000	140,000	130,000	120,000	100,000
Process Water	260,000	1,570,000	920,000	670,000	590,000	530,000	410,000
Zymo	1,850,000	-	-	-	-	-	-
Cooling Water	440,000	930,000	570,000	440,000	380,000	350,000	270,000
Chilled Water	-	2,000,000	1,270,000	990,000	890,000	820,000	660,000
Electricity	340,000	3,060,000	1,900,000	1,470,000	1,300,000	1,200,000	950,000
Capital Depreciation	260,000	1,620,000	1,150,000	960,000	880,000	830,000	700,000
MESP4 ($/gal)	0.41	0.70	0.49	0.42	0.40	0.39	0.40
Change in MESP Over No Recycle	-	-0.13	0.08	0.15	0.17	0.18	0.17
Max Savings ($/gal)		0.18					
Purge %		25					

Table F.1.2 (Continued)

Single Step Deactivation Scheme	No Recycle	50 % Activity Loss					
Purge %		5	10	15	20	25	50
Capital Costs ($/yr)							
Total Installed Equipment Cost ($/yr)	5,200,397	37,220,000	25,740,000	21,170,000	18,220,000	17,050,000	13,930,000
Indirect Costs ($/yr)	160,000	1,120,000	770,000	640,000	550,000	510,000	420,000
Total Project Investment ($/yr)	5,360,000	38,340,000	26,510,000	21,810,000	18,770,000	17,560,000	14,350,000
Operating Costs (cents/gal)							
CSL	5.0	12.6	7.7	6.0	5.0	4.7	4.0
Cellulase	24.5	13.3	14.0	14.6	15.2	15.9	19.0
DAP	0.6	1.4	0.9	0.7	0.6	0.5	0.4
Process Water	1.1	7.8	4.4	3.2	2.5	2.3	1.8
Zymo	1.8	0.0	0.0	0.0	0.0	0.0	0.0
Cooling Water	1.8	4.6	2.8	2.1	1.7	1.6	1.2
Chilled Water	0.0	9.9	6.0	4.7	3.9	3.6	2.9
Electricity	1.4	15.2	9.1	7.0	5.8	5.3	4.2
Capital Depreciation	1.2	7.5	5.3	4.4	3.8	3.6	3.1
Operating Costs ($/yr)							
CSL	1,120,000	3,220,000	1,950,000	1,510,000	1,250,000	1,160,000	920,000
Cellulase	5,520,000	3,410,000	3,530,000	3,650,000	3,770,000	3,880,000	4,430,000
DAP	120,000	360,000	220,000	170,000	140,000	130,000	100,000
Process Water	260,000	2,000,000	1,120,000	810,000	620,000	560,000	410,000
Zymo	1,850,000	-	-	-	-	-	-
Cooling Water	440,000	1,180,000	700,000	520,000	420,000	380,000	280,000
Chilled Water	-	2,520,000	1,530,000	1,180,000	960,000	880,000	680,000
Electricity	340,000	3,870,000	2,310,000	1,760,000	1,430,000	1,300,000	980,000
Capital Depreciation	260,000	1,920,000	1,330,000	1,090,000	940,000	880,000	720,000
MESP4 ($/gal)	0.41	0.91	0.63	0.54	0.48	0.46	0.44
Change in MESP Over No Recycle	-	-0.34	-0.06	0.03	0.09	0.11	0.13
Max Savings ($/gal)	-	0.13					
Purge %		50					

Table F.2.1: Multi-Step Deactivation Scheme – Raw Material Usage

Multi Step Deactivation Scheme	No Recycle	0% Conversion					
Purge %	NA	5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40444	47511	47030	46558	46095	45641	43501
Cellulase In (lb/hr)	12133	14252	14109	13967	13828	13692	13050
Fresh (lb/hr)	12133	734	1416	2087	2744	3387	6426
Recycled (lb/hr)	0	13519	12693	11881	11085	10305	6624
Cellu40 In (lb/hr)	0	0	0	0	0	0	0
Cellu30 In (lb/hr)	0	0	0	0	0	0	0
Cellu20 In (lb/hr)	0	0	0	0	0	0	0
Cellu10 In (lb/hr)	0	0	0	0	0	0	0
Decellulase In (lb/hr)	0	0	0	0	0	0	0
Tot Cellulase In (lb/hr)	12133	14252	14109	13967	13828	13692	13050
Cellulase Out (lb/hr)	10313	14252	14109	13967	13828	13692	13050
Cellu40 Out (lb/hr)	0	0	0	0	0	0	0
Cellu30 Out (lb/hr)	0	0	0	0	0	0	0
Cellu20 Out (lb/hr)	0	0	0	0	0	0	0
Cellu10 Out (lb/hr)	0	0	0	0	0	0	0
Decellulase Out (lb/hr)	1820	0	0	0	0	0	0
Total Cellulase Out (lb/hr)	12133	14252	14109	13967	13828	13692	13050
% Activity Added (Fresh /Total)		5.15		14.94	19.84	24.74	49.24
Activity Loss (Fresh Cellulase Weight)		0	0	0	0	0	0
Avg. % Activity Loss ((In-Out)/(In))		0.00					
Fresh DAP Added (lb/hr)	209	419	256	211	191	173	168
Fresh CSL Added (lb/hr)	1659	3296	2004	1645	1485	1348	1310
DAP Recycled (lb/hr)	0	353	199	152	129	110	79
CSL Recycled (lb/hr)	0	2842	1614	1244	1057	902	651
Total DAP In (lb/hr)	209	772	455	363	320	283	246
Total CSL In (lb/hr)	1659	6138	3618	2889	2542	2250	1961
Total DAP Out (lb/hr)	196	755	438	347	303	267	231
Total CSL Out (lb/hr)	1613	6078	3559	2830	2483	2192	1906
Fresh Zymo Added (lb/hr)	330	0	0	0	0	0	0
Ethanol wt% into SSF	0.0%	0.7%	1.1%	1.3%	1.4%	1.5%	1.2%
Ethanol Produced (lb/hr)	16973	20074	19862	19653	19449	19249	18305
Ethanol in product stream	16848	19459	18707	18021	17384	16782	14269
Ethanol in Purge (lb/hr)	0	526	1010	1460	1878	2266	3854
Ethanol Production (MM Gal. / Year)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.2.1 (Continued)

Multi Step Deactivation Scheme	No Recycle	10% Conversion					
Purge %	NA	5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40444	47511	47030	46558	46095	45641	43501
Cellulase In (lb/hr)	12133	7440	9020	10036	10713	11178	12080
Fresh (lb/hr)	12133	1088	1717	2353	2984	3607	6561
Recycled (lb/hr)	0	6352	7303	7683	7729	7571	5519
Cellu40 In (lb/hr)	0	4825	4262	3641	3083	2607	1129
Cellu30 In (lb/hr)	0	3130	2014	1321	887	608	105
Cellu20 In (lb/hr)	0	2030	952	479	255	142	10
Cellu10 In (lb/hr)	0	1316	450	174	73	33	1
Decellulase In (lb/hr)	0	2433	403	99	30	10	0
Tot Cellulase In (lb/hr)	12133	21174	17101	15749	15042	14579	13325
Cellulase Out (lb/hr)	10313	6696	8118	9032	9642	10060	10872
Cellu40 Out (lb/hr)	0	5087	4738	4280	3846	3464	2224
Cellu30 Out (lb/hr)	0	3299	2239	1553	1107	808	208
Cellu20 Out (lb/hr)	0	2140	1058	563	318	188	19
Cellu10 Out (lb/hr)	0	1388	500	204	92	44	2
Decellulase Out (lb/hr)	1820	2565	448	116	37	13	0
Total Cellulase Out (lb/hr)	12133	21174	17101	15749	15042	14579	13325
% Activity Added (Fresh /Total)	0.00	7.63	12.17	16.85	21.58	26.34	50.28
Activity Loss (Fresh Cellulase Weight)	0	375	334	313	300	291	266
Avg. % Activity Loss ((In-Out)/(In))	0	2.26					
Fresh DAP Added (lb/hr)	209	404	259	212	194	182	153
Fresh CSL Added (lb/hr)	1659	3177	2025	1652	1510	1415	1191
DAP Recycled (lb/hr)	0	340	201	153	131	116	71
CSL Recycled (lb/hr)	0	2737	1631	1250	1076	949	589
Total DAP In (lb/hr)	209	743	460	365	325	297	224
Total CSL In (lb/hr)	1659	5914	3656	2902	2586	2364	1780
Total DAP Out (lb/hr)	196	727	443	348	309	281	208
Total CSL Out (lb/hr)	1613	5854	3596	2843	2528	2306	1725
Fresh Zymo Added (lb/hr)	330	0	0	0	0	0	0
Ethanol wt% into SSF	0.0%	0.7%	1.1%	1.3%	1.4%	1.4%	1.3%
Ethanol Produced (lb/hr)	16973	20074	19862	19653	19449	19249	18305
Ethanol in product stream	16848	19455	18709	18022	17387	16793	14255
Ethanol in Purge (lb/hr)	0	525	1011	1460	1878	2268	3850
Ethanol Production (MM Gal. / Year)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.2.1 (Continued)

Multi Step Deactivation Scheme	No Recycle	25% Conversion					
Purge %	NA	5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40444	47511	47030	46558	46095	45641	43501
Cellulase In (lb/hr)	12133	5999	6992	7839	8552	9148	10940
Fresh (lb/hr)	12133	1731	2275	2838	3410	3984	6775
Recycled (lb/hr)	0	4268	4718	5001	5141	5164	4165
Cellu40 In (lb/hr)	0	4930	4834	4604	4297	3952	2242
Cellu30 In (lb/hr)	0	4052	3342	2704	2159	1707	459
Cellu20 In (lb/hr)	0	3330	2311	1588	1085	737	94
Cellu10 In (lb/hr)	0	2737	1598	933	545	319	19
Decellulase In (lb/hr)	0	12608	3577	1328	551	242	5
Tot Cellulase In (lb/hr)	12133	33656	22654	18997	17190	16105	13759
Cellulase Out (lb/hr)	10313	4499	5244	5879	6414	6861	8205
Cellu40 Out (lb/hr)	0	5197	5374	5413	5361	5251	4416
Cellu30 Out (lb/hr)	0	4271	3715	3179	2694	2268	905
Cellu20 Out (lb/hr)	0	3511	2569	1867	1354	980	185
Cellu10 Out (lb/hr)	0	2885	1776	1097	680	423	38
Decellulase Out (lb/hr)	1820	13292	3976	1561	687	322	10
Total Cellulase Out (lb/hr)	12133	33656	22654	18997	17190	16105	13759
% Activity Added (Fresh /Total)	0.00	12.14	16.12	20.32	24.66	29.10	51.91
Activity Loss (Fresh Cellulase Weight)	0	1052	954	883	832	793	688
Avg. % Activity Loss ((In-Out)/(In))	0	6.26					
Fresh DAP Added (lb/hr)	209	419	256	217	193	186	169
Fresh CSL Added (lb/hr)	1659	3298	2004	1692	1506	1453	1318
DAP Recycled (lb/hr)	0	353	199	157	131	119	79
CSL Recycled (lb/hr)	0	2844	1614	1281	1072	976	655
Total DAP In (lb/hr)	209	772	455	374	324	305	248
Total CSL In (lb/hr)	1659	6142	3618	2973	2578	2429	1973
Total DAP Out (lb/hr)	196	755	438	357	308	289	233
Total CSL Out (lb/hr)	1613	6082	3559	2914	2520	2371	1918
Fresh Zymo Added (lb/hr)	330	0	0	0	0	0	0
Ethanol wt% into SSF	0.0%	0.7%	1.1%	1.3%	1.4%	1.4%	1.2%
Ethanol Produced (lb/hr)	16973	20074	19862	19653	19449	19249	18305
Ethanol in product stream	16848	19460	18709	18026	17387	16797	14270
Ethanol in Purge (lb/hr)	0	526	1010	1460	1878	2268	3854
Ethanol Production (MM Gal. / Year)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.2.1 (Continued)

Multi Step Deactivation Scheme	No Recycle	50% Conversion					
Purge %	NA	5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40444	47511	47030	46558	46095	45641	43501
Cellulase In (lb/hr)	12133	5406	5971	6518	7043	7542	9596
Fresh (lb/hr)	12133	2842	3285	3746	4220	4704	7161
Recycled (lb/hr)	0	2564	2686	2772	2823	2838	2436
Cellu40 In (lb/hr)	0	4878	4881	4824	4711	4550	3264
Cellu30 In (lb/hr)	0	4401	3991	3570	3151	2745	1110
Cellu20 In (lb/hr)	0	3970	3262	2642	2108	1656	378
Cellu10 In (lb/hr)	0	3582	2667	1955	1410	999	128
Decellulase In (lb/hr)	0	33002	11949	5566	2848	1520	66
Tot Cellulase In (lb/hr)	12133	55239	32721	25075	21272	19014	14543
Cellulase Out (lb/hr)	10313	2703	2985	3259	3522	3771	4798
Cellu40 Out (lb/hr)	0	5142	5426	5671	5877	6046	6430
Cellu30 Out (lb/hr)	0	4639	4436	4197	3931	3648	2187
Cellu20 Out (lb/hr)	0	4186	3627	3106	2630	2201	744
Cellu10 Out (lb/hr)	0	3776	2965	2298	1759	1328	253
Decellulase Out (lb/hr)	1820	34793	13282	6543	3553	2020	130
Total Cellulase Out (lb/hr)	12133	55239	32721	25075	21272	19014	14543
% Activity Added (Fresh /Total)	0.00	19.94	23.28	26.82	30.52	34.36	54.87
Activity Loss (Fresh Cellulase Weight)	0	2224	2077	1951	1842	1749	1448
Avg. % Activity Loss ((In-Out)/(In))	0	13.58					
Fresh DAP Added (lb/hr)	209	419	284	222	198	187	154
Fresh CSL Added (lb/hr)	1659	3298	2229	1732	1548	1459	1204
DAP Recycled (lb/hr)	0	353	222	161	135	119	72
CSL Recycled (lb/hr)	0	2844	1801	1313	1104	980	596
Total DAP In (lb/hr)	209	772	507	383	333	306	226
Total CSL In (lb/hr)	1659	6142	4030	3045	2652	2438	1801
Total DAP Out (lb/hr)	196	755	490	366	317	290	211
Total CSL Out (lb/hr)	1613	6082	3971	2986	2594	2380	1745
Fresh Zymo Added (lb/hr)	330	0	0	0	0	0	0
Ethanol wt% into SSF	0.0%	0.7%	1.0%	1.2%	1.3%	1.3%	1.3%
Ethanol Produced (lb/hr)	16973	20074	19862	19653	19449	19249	18305
Ethanol in product stream	16848	19461	18720	18031	17391	16799	14257
Ethanol in Purge (lb/hr)	0	526	1011	1461	1879	2268	3850
Ethanol Production (MM Gal. / Year)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.2.1 (Continued)

Multi Step Deactivation Scheme	No Recycle	85% Conversion					
Purge %	NA	5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40444	47511	47030	46558	46095	45641	43501
Cellulase In (lb/hr)	12133	5144	5481	5827	6181	6542	8394
Fresh (lb/hr)	12133	4412	4741	5083	5438	5804	7755
Recycled (lb/hr)	0	732	740	743	743	739	639
Cellu40 In (lb/hr)	0	4836	4845	4829	4787	4718	3920
Cellu30 In (lb/hr)	0	4546	4282	4002	3708	3402	1831
Cellu20 In (lb/hr)	0	4274	3785	3317	2871	2453	855
Cellu10 In (lb/hr)	0	4018	3346	2749	2224	1769	399
Decellulase In (lb/hr)	0	63021	25483	13302	7637	4574	350
Tot Cellulase In (lb/hr)	12133	85838	47222	34024	27408	23458	15749
Cellulase Out (lb/hr)	10313	772	822	874	927	981	1259
Cellu40 Out (lb/hr)	0	5098	5385	5677	5972	6268	7723
Cellu30 Out (lb/hr)	0	4792	4760	4705	4625	4520	3607
Cellu20 Out (lb/hr)	0	4505	4208	3899	3582	3260	1684
Cellu10 Out (lb/hr)	0	4235	3719	3231	2774	2351	787
Decellulase Out (lb/hr)	1820	66436	28327	15638	9528	6078	689
Total Cellulase Out (lb/hr)	12133	85838	47222	34024	27408	23458	15749
% Activity Added (Fresh /Total)	0.00	30.96	33.60	36.39	39.32	42.39	
Activity Loss (Fresh Cellulase Weight)	0	3879	3696	3523	3361	3210	2618
Avg. % Activity Loss ((In-Out)/(In))	0	24.41					
Fresh DAP Added (lb/hr)	209	480	299	240	212	200	156
Fresh CSL Added (lb/hr)	1659	3782	2348	1877	1659	1560	1217
DAP Recycled (lb/hr)	0	407	235	175	145	128	73
CSL Recycled (lb/hr)	0	3269	1900	1426	1186	1051	603
Total DAP In (lb/hr)	209	886	534	415	358	328	229
Total CSL In (lb/hr)	1659	7051	4248	3302	2845	2612	1820
Total DAP Out (lb/hr)	196	869	517	399	341	312	213
Total CSL Out (lb/hr)	1613	6991	4189	3243	2786	2554	1765
Fresh Zymo Added (lb/hr)	330	0	0	0	0	0	0
Ethanol wt% into SSF	0.0%	0.6%	0.9%	1.2%	1.2%	1.3%	1.3%
Ethanol Produced (lb/hr)	16973	20074	19862	19653	19449	19249	18305
Ethanol in product stream	16848	19472	18728	18039	17404	16805	14259
Ethanol in Purge (lb/hr)	0	526	1012	1461	1880	2269	3851
Ethanol Production (MM Gal. / Year)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.2.1 (Continued)

Multi Step Deactivation Scheme	No Recycle	95% Conversion					
Purge %	NA	5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40444	47511	47030	46558	46095	45641	43501
Cellulase In (lb/hr)	12133	5104	5403	5714	6035	6367	8140
Fresh (lb/hr)	12133	4862	5160	5471	5794	6128	7934
Recycled (lb/hr)	0	242	243	243	242	240	207
Cellu40 In (lb/hr)	0	4828	4835	4822	4788	4730	4028
Cellu30 In (lb/hr)	0	4568	4327	4070	3798	3514	1993
Cellu20 In (lb/hr)	0	4321	3872	3435	3013	2611	986
Cellu10 In (lb/hr)	0	4088	3465	2899	2391	1940	488
Decellulase In (lb/hr)	0	71674	29493	15679	9176	5606	478
Tot Cellulase In (lb/hr)	12133	94584	51395	36619	29202	24768	16113
Cellulase Out (lb/hr)	10313	255	270	286	302	318	407
Cellu40 Out (lb/hr)	0	5090	5375	5669	5973	6285	7935
Cellu30 Out (lb/hr)	0	4815	4810	4785	4739	4670	3926
Cellu20 Out (lb/hr)	0	4556	4304	4038	3759	3469	1942
Cellu10 Out (lb/hr)	0	4310	3852	3408	2982	2577	961
Decellulase Out (lb/hr)	1820	75558	32785	18433	11447	7448	941
Total Cellulase Out (lb/hr)	12133	94584	51395	36619	29202	24768	16113
% Activity Added (Fresh /Total)	0.00	34.11	36.57	39.17	41.90	44.75	60.79
Activity Loss (Fresh Cellulase Weight)	0	4353	4161	3979	3805	3641	2971
Avg. % Activity Loss ((In-Out)/(In))	0	27.56					
Fresh DAP Added (lb/hr)	209	487	303	243	214	202	156
Fresh CSL Added (lb/hr)	1659	3843	2379	1905	1674	1578	1221
DAP Recycled (lb/hr)	0	413	238	178	147	130	73
CSL Recycled (lb/hr)	0	3322	1926	1448	1197	1063	605
Total DAP In (lb/hr)	209	901	541	421	361	332	230
Total CSL In (lb/hr)	1659	7165	4305	3353	2871	2641	1826
Total DAP Out (lb/hr)	196	884	524	405	344	316	214
Total CSL Out (lb/hr)	1613	7104	4245	3294	2813	2583	1771
Fresh Zymo Added (lb/hr)	330	0	0	0	0	0	0
Ethanol wt% into SSF	0.0%	0.6%	0.9%	1.1%	1.2%	1.3%	1.3%
Ethanol Produced (lb/hr)	16973	20074	19862	19653	19449	19249	18305
Ethanol in product stream	16848	19473	18730	18041	17406	16807	14260
Ethanol in Purge (lb/hr)	0	526	1012	1462	1880	2269	3851
Ethanol Production (MM Gal. / Year)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.2.1 (Continued)

Multi Step Deactivation Scheme	No Recycle	100% Conversion					
Purge %	NA	5	10	15	20	25	50
Cellulose Into SSF Reactor (lb/hr)	40444	47511	47030	46558	46095	45641	43501
Cellulase In (lb/hr)	12133	5087	5370	5665	5972	6291	8025
Fresh (lb/hr)	12133	5087	5370	5665	5972	6291	8025
Recycled (lb/hr)	0	0	0	0	0	0	0
Cellu40 In (lb/hr)	0	4825	4831	4819	4787	4734	4073
Cellu30 In (lb/hr)	0	4577	4346	4099	3838	3563	2068
Cellu20 In (lb/hr)	0	4342	3909	3487	3076	2682	1050
Cellu10 In (lb/hr)	0	4119	3517	2966	2466	2018	533
Decellulase In (lb/hr)	0	76009	31512	16885	9963	6139	549
Tot Cellulase In (lb/hr)	12133	98958	53484	37920	30102	25427	16297
Cellulase Out (lb/hr)	10313	0	0	0	0	0	0
Cellu40 Out (lb/hr)	0	5087	5370	5665	5972	6291	8025
Cellu30 Out (lb/hr)	0	4825	4831	4819	4787	4734	4073
Cellu20 Out (lb/hr)	0	4577	4346	4099	3838	3563	2068
Cellu10 Out (lb/hr)	0	4342	3909	3487	3076	2682	1050
Decellulase Out (lb/hr)	1820	80128	35029	19850	12429	8158	1082
Total Cellulase Out (lb/hr)	12133	98958	53484	37920	30102	25427	16297
% Activity Added (Fresh /Total)	0.00	35.69	38.06	40.56	43.19	45.94	61.49
Activity Loss (Fresh Cellulase Weight)	0	4590	4394	4207	4028	3858	3150
Avg. % Activity Loss ((In-Out)/(In))	0	29.15					
Fresh DAP Added (lb/hr)	209	491	305	247	214	206	172
Fresh CSL Added (lb/hr)	1659	3872	2390	1933	1675	1609	1347
DAP Recycled (lb/hr)	0	416	239	181	147	133	81
CSL Recycled (lb/hr)	0	3347	1935	1470	1198	1085	670
Total DAP In (lb/hr)	209	907	4634	428	361	339	254
Total CSL In (lb/hr)	1659	7219	1966	3403	2873	2695	2017
Total DAP Out (lb/hr)	196	891	527	411	345	322	238
Total CSL Out (lb/hr)	1613	7159	4266	3344	2814	2637	1962
Fresh Zymo Added (lb/hr)	330	0	0	0	0	0	0
Ethanol wt% into SSF	0.0%	0.6%	0.9%	1.1%	1.2%	1.2%	1.1%
Ethanol Produced (lb/hr)	16973	20074	19862	19653	19449	19249	18305
Ethanol in product stream	16848	19474	18731	18043	17407	16810	14274
Ethanol in Purge (lb/hr)	0	526	1012	1462	1880	2270	3855
Ethanol Production (MM Gal. / Year)	22.50	25.54	25.27	25.01	24.75	24.49	23.29

Table F.2.2: Multi-Step Deactivation Scheme – Economics Summary

Multi Step Deactivation Scheme	No Recycle	0% Conversion					
Purge %	NA	5	10	15	20	25	50
Capital Costs (MM $/yr)							
Total Installed Equipment Cost (MM $/yr)	5.20	28.49	19.93	17.07	15.59	14.36	13.25
Added Costs (MM $/yr)	0.16	0.85	0.60	0.51	0.47	0.43	0.40
Total Project Investment (MM $/yr)	5.36	29.34	20.53	17.58	16.06	14.79	13.65
Operating Costs (cents/gal)							
CSL	5.0	8.6	5.3	4.4	4.0	3.7	3.8
Cellulase	24.5	1.3	2.5	3.8	5.0	6.3	12.5
DAP	0.6	1.0	0.6	0.5	0.5	0.4	0.4
Process Water	1.1	5.5	3.1	2.4	2.1	1.8	1.8
Zymo	1.8	-	-	-	-	-	-
Cooling Water	1.8	3.1	1.9	1.5	1.3	1.2	1.1
Chilled Water	-	6.9	4.3	3.5	3.2	2.9	2.8
Electricity	1.4	10.4	6.4	5.2	4.6	4.2	3.9
Capital Depreciation	1.2	5.8	4.1	3.5	3.2	3.0	2.9
Operating Costs (MM $/yr)							
CSL	1.12	2.20	1.34	1.10	0.99	0.90	0.87
Cellulase	5.52	0.33	0.64	0.95	1.25	1.54	2.92
DAP	0.12	0.25	0.15	0.12	0.11	0.10	0.10
Process Water	0.26	1.40	0.79	0.61	0.52	0.45	0.41
Zymo	1.85	0.00	0.00	0.00	0.00	0.00	0.00
Cooling Water	0.44	0.80	0.48	0.38	0.33	0.30	0.26
Chilled Water	0.00	1.76	1.09	0.89	0.79	0.71	0.64
Electricity	0.34	2.66	1.62	1.30	1.15	1.02	0.91
Capital Depreciation	0.26	1.47	1.03	0.88	0.80	0.74	0.68
MESP4 ($/gal)	0.41	0.57	0.38	0.33	0.32	0.31	0.36
Change in MESP Over No Recycle	0	0.00	0.19	0.24	0.25	0.26	0.21
Max Savings ($/gal)		0.26					
Purge %		25					

Table F.2.2 (Continued)

Multi Step Deactivation Scheme	No Recycle	10% Conversion					
Purge %	NA	5	10	15	20	25	50
Capital Costs (MM $/yr)							
Total Installed Equipment Cost (MM $/yr)	5.20	27.70	20.16	17.15	15.78	15.02	12.43
Added Costs (MM $/yr)	0.16	0.83	0.60	0.52	0.48	0.45	0.38
Total Project Investment (MM $/yr)	5.36	28.53	20.76	17.67	16.26	15.47	12.81
Operating Costs (cents/gal)	0						
CSL	5.0	8.3	5.3	4.4	4.1	3.9	3.4
Cellulase	24.5	1.9	3.1	4.3	5.5	6.7	12.8
DAP	0.6	0.9	0.6	0.5	0.5	0.4	0.4
Process Water	1.1	5.2	3.2	2.5	2.2	2.0	1.5
Zymo	1.8	-	-	-	-	-	-
Cooling Water	1.8	3.0	1.9	1.5	1.4	1.3	1.0
Chilled Water	-	6.6	4.4	3.6	3.6	3.1	2.5
Electricity	1.4	10.0	6.5	5.2	4.7	4.4	3.6
Capital Depreciation	1.2	5.6	4.1	3.5	3.3	3.1	2.7
Operating Costs (MM $/yr)							
CSL	1.12	2.12	1.35	1.10	1.01	0.94	0.79
Cellulase	5.52	0.50	0.78	1.07	1.36	1.64	2.98
DAP	0.12	0.24	0.15	0.13	0.11	0.11	0.09
Process Water	0.26	1.33	0.80	0.61	0.53	0.49	0.35
Zymo	1.85	0.00	0.00	0.00	0.00	0.00	0.00
Cooling Water	0.44	0.77	0.49	0.38	0.34	0.32	0.24
Chilled Water	0.00	1.69	1.10	0.89	0.89	0.75	0.59
Electricity	0.34	2.56	1.64	1.31	1.17	1.09	0.84
Capital Depreciation	0.26	1.43	1.04	0.88	0.81	0.77	0.64
MESP4 ($/gal)	0.41	0.55	0.39	0.34	0.33	0.33	0.35
Change in MESP Over No Recycle	0	0.02	0.18	0.23	0.24	0.24	0.22
Max Savings ($/gal)		0.24					
Purge %		25					

Table F.2.2 (Continued)

Multi Step Deactivation Scheme	No Recycle	25% Conversion					
Purge %	NA	5	10	15	20	25	50
Capital Costs (MM $/yr)							
Total Installed Equipment Cost (MM $/yr)	5.20	28.76	20.15	17.48	15.75	15.29	13.32
Added Costs (MM $/yr)	0.16	0.87	0.60	0.53	0.47	0.45	0.40
Total Project Investment (MM $/yr)	5.36	29.63	20.75	18.01	16.22	15.74	13.72
Operating Costs (cents/gal)	0						
CSL	5.0	8.6	5.3	4.5	4.1	4.0	3.8
Cellulase	24.5	3.1	4.1	5.2	6.3	7.4	13.2
DAP	0.6	1.0	0.6	0.5	0.5	0.5	0.4
Process Water	1.1	5.5	3.1	2.5	2.1	2.1	1.8
Zymo	1.8	-	-	-	-	-	-
Cooling Water	1.8	3.2	1.9	1.6	1.4	1.3	1.1
Chilled Water	-	7.0	4.4	3.7	3.2	3.1	2.8
Electricity	1.4	10.6	6.5	5.4	4.7	4.6	4.0
Capital Depreciation	1.2	5.8	4.1	3.6	3.3	3.2	3.0
Operating Costs (MM $/yr)							
CSL	1.12	2.20	1.34	1.13	1.00	0.97	0.88
Cellulase	5.52	0.79	1.03	1.29	1.55	1.81	3.08
DAP	0.12	0.25	0.15	0.13	0.11	0.11	0.10
Process Water	0.26	1.40	0.79	0.63	0.52	0.50	0.41
Zymo	1.85	0.00	0.00	0.00	0.00	0.00	0.00
Cooling Water	0.44	0.82	0.49	0.40	0.34	0.32	0.26
Chilled Water	0.00	1.78	1.10	0.91	0.80	0.77	0.65
Electricity	0.34	2.70	1.64	1.35	1.16	1.12	0.92
Capital Depreciation	0.26	1.48	1.04	0.90	0.81	0.79	0.69
MESP4 ($/gal)	0.41	0.59	0.40	0.36	0.34	0.34	0.37
Change in MESP Over No Recycle	0	-0.02	0.17	0.21	0.24	0.23	0.20
Max Savings ($/gal)		0.24					
Purge %		20					

Table F.2.2 (Continued)

Multi Step Deactivation Scheme	No Recycle	50% Conversion					
Purge %	NA	5	10	15	20	25	50
Capital Costs (MM $/yr)							
Total Installed Equipment Cost (MM $/yr)	5.20	29.05	21.44	17.89	16.07	15.45	12.56
Added Costs (MM $/yr)	0.16	0.88	0.65	0.53	0.49	0.46	0.38
Total Project Investment (MM $/yr)	5.36	29.93	22.09	18.42	16.56	15.91	12.94
Operating Costs (cents/gal)	0						
CSL	5.0	8.6	5.9	4.6	4.2	4.0	3.4
Cellulase	24.5	5.1	5.9	6.8	7.8	8.7	14.0
DAP	0.6	1.0	0.7	0.5	0.5	0.5	0.4
Process Water	1.1	5.5	3.4	2.6	2.2	2.1	1.5
Zymo	1.8	-	-	-	-	-	-
Cooling Water	1.8	3.3	2.1	1.6	1.4	1.3	1.0
Chilled Water	-	7.0	4.7	3.8	3.3	3.2	2.6
Electricity	1.4	10.7	7.1	5.6	4.8	4.6	3.7
Capital Depreciation	1.2	5.9	4.4	3.7	3.4	3.3	2.8
Operating Costs (MM $/yr)							
CSL	1.12	2.20	1.49	1.16	1.03	0.97	0.80
Cellulase	5.52	1.29	1.49	1.70	1.92	2.14	3.26
DAP	0.12	0.25	0.17	0.13	0.12	0.11	0.09
Process Water	0.26	1.40	0.87	0.64	0.53	0.50	0.35
Zymo	1.85	0.00	0.00	0.00	0.00	0.00	0.00
Cooling Water	0.44	0.83	0.53	0.41	0.35	0.33	0.24
Chilled Water	0.00	1.79	1.20	0.94	0.82	0.78	0.60
Electricity	0.34	2.74	1.79	1.39	1.20	1.13	0.85
Capital Depreciation	0.26	1.50	1.10	0.92	0.83	0.80	0.65
MESP4 ($/gal)	0.41	0.61	0.45	0.38	0.36	0.36	0.36
Change in MESP Over No Recycle	0	-0.04	0.12	0.19	0.21	0.21	0.21
Max Savings ($/gal)		0.21					
Purge %		25					

Table F.2.2 (Continued)

Multi Step Deactivation Scheme	No Recycle	85% Conversion					
Purge %	NA	5	10	15	20	25	50
Capital Costs (MM $/yr)							
Total Installed Equipment Cost (MM $/yr)	5.20	31.79	22.47	18.56	17.08	15.86	12.68
Added Costs (MM $/yr)	0.16	0.95	0.67	0.56	0.51	0.48	0.38
Total Project Investment (MM $/yr)	5.36	32.74	23.14	19.12	17.59	16.34	13.06
Operating Costs (cents/gal)							
CSL	5.0	9.9	6.2	5.0	4.5	4.2	3.5
Cellulase	24.5	7.9	8.5	9.2	10.0	10.8	15.1
DAP	0.6	1.1	0.7	0.6	0.5	0.5	0.4
Process Water	1.1	6.2	3.6	2.7	2.4	2.1	1.5
Zymo	1.8	-	-	-	-	-	-
Cooling Water	1.8	3.7	2.3	1.7	1.5	2.8	1.1
Chilled Water	-	7.9	5.0	3.9	3.6	3.3	2.6
Electricity	1.4	12.2	7.6	5.9	5.3	4.8	3.7
Capital Depreciation	1.2	6.4	4.6	3.8	3.6	3.3	2.8
Operating Costs (MM $/yr)							
CSL	1.12	2.52	1.57	1.25	1.11	1.04	0.81
Cellulase	5.52	2.01	2.16	2.31	2.47	2.64	3.53
DAP	0.12	0.28	0.18	0.14	0.13	0.12	0.09
Process Water	0.26	1.60	0.92	0.67	0.58	0.52	0.35
Zymo	1.85	0.00	0.00	0.00	0.00	0.00	0.00
Cooling Water	0.44	0.94	2.26	1.73	0.38	0.68	0.24
Chilled Water	0.00	2.03	5.03	3.95	0.89	0.81	0.61
Electricity	0.34	3.11	7.56	5.85	1.30	1.17	0.86
Capital Depreciation	0.26	1.64	1.16	0.96	0.88	0.82	0.65
MESP4 ($/gal)	0.41	0.71	0.50	0.42	0.40	0.40	0.38
Change in MESP Over No Recycle	0	-0.14	0.07	0.15	0.17	0.17	0.19
Max Savings ($/gal)		0.19					
Purge %		50					

Table F.2.2 (Continued)

Multi Step Deactivation Scheme	No Recycle	95% Conversion					
Purge %	NA	5	10	15	20	25	50
Capital Costs (MM $/yr)							
Total Installed Equipment Cost (MM $/yr)	5.20	32.27	22.75	18.75	17.22	15.98	12.72
Added Costs (MM $/yr)	0.16	0.97	0.68	0.57	0.52	0.48	0.38
Total Project Investment (MM $/yr)	5.36	33.24	23.43	19.32	17.74	16.46	13.10
Operating Costs (cents/gal)	0						
CSL	5.0	10.0	6.3	5.1	4.5	4.3	3.5
Cellulase	24.5	8.7	9.3	10.0	10.6	11.4	15.5
DAP	0.6	1.1	0.7	0.6	0.5	0.5	0.4
Process Water	1.1	6.4	3.7	2.7	2.4	2.1	1.5
Zymo	1.8	-	-	-	-	-	-
Cooling Water	1.8	3.8	2.3	1.8	1.6	1.4	1.1
Chilled Water	-	8.1	5.1	4.0	3.6	3.3	2.6
Electricity	1.4	12.4	7.7	5.9	5.3	4.8	3.7
Capital Depreciation	1.2	6.5	4.6	3.9	3.6	3.3	2.8
Operating Costs (MM $/yr)							
CSL	1.12	2.56	1.59	1.27	1.12	1.05	0.81
Cellulase	5.52	2.21	2.35	2.49	2.64	2.79	3.61
DAP	0.12	0.29	0.18	0.14	0.13	0.12	0.09
Process Water	0.26	1.63	0.94	0.67	0.59	0.52	0.35
Zymo	1.85	0.00	0.00	0.00	0.00	0.00	0.00
Cooling Water	0.44	0.97	0.58	0.44	0.39	0.35	0.25
Chilled Water	0.00	2.07	1.29	1.00	0.90	0.81	0.61
Electricity	0.34	3.17	1.95	1.49	1.32	1.19	0.87
Capital Depreciation	0.26	1.66	1.17	0.97	0.89	0.82	0.66
MESP4 ($/gal)	0.41	0.73	0.51	0.43	0.41	0.40	0.38
Change in MESP Over No Recycle	0	-0.16	0.06	0.14	0.16	0.18	0.19
Max Savings ($/gal)		0.19					
Purge %		50					

Table F.2.2 (Continued)

Multi Step Deactivation Scheme	No Recycle	100% Conversion					
Purge %	NA	5	10	15	20	25	50
Capital Costs (MM $/yr)							
Total Installed Equipment Cost (MM $/yr)	5.20	32.51	22.93	18.95	17.29	16.21	13.58
Added Costs (MM $/yr)	0.16	0.97	0.69	0.57	0.52	0.48	0.41
Total Project Investment (MM $/yr)	5.36	33.48	23.62	19.52	17.81	16.69	13.99
Operating Costs (cents/gal)							
CSL	5.0	10.1	0.1	5.2	4.5	4.4	3.9
Cellulase	24.5	9.1	9.7	10.3	11.0	11.7	15.7
DAP	0.6	1.1	10.3	0.6	0.5	0.5	0.4
Process Water	1.1	6.4	3.7	2.7	2.4	2.2	1.8
Zymo	1.8	-	-	-	-	-	-
Cooling Water	1.8	3.8	2.3	1.8	1.6	1.4	1.2
Chilled Water	-	8.2	5.2	4.1	3.6	3.4	2.8
Electricity	1.4	12.6	7.8	6.0	5.4	4.9	4.1
Capital Depreciation	1.2	6.5	4.7	3.9	3.6	3.4	3.0
Operating Costs (MM $/yr)							
CSL	1.12	2.58	0.02	1.29	1.12	1.07	0.90
Cellulase	5.52	2.31	2.44	2.58	2.72	2.86	3.65
DAP	0.12	0.29	2.60	0.15	0.13	0.12	0.10
Process Water	0.26	1.64	0.95	0.69	0.59	0.53	0.41
Zymo	1.85	0.00	0.00	0.00	0.00	0.00	0.00
Cooling Water	0.44	0.98	0.59	0.45	0.39	0.35	0.27
Chilled Water	0.00	2.09	1.31	1.01	0.90	0.83	0.66
Electricity	0.34	3.21	1.97	1.51	1.33	1.21	0.95
Capital Depreciation	0.26	1.67	1.18	0.98	0.89	0.83	0.70
MESP4 ($/gal)	0.41	0.74	0.55	0.44	0.41	0.40	0.40
Change in MESP Over No Recycle	0	-0.17	0.02	0.13	0.16	0.17	0.17
Max Savings ($/gal)		0.17					
Purge %		50					

Table F.3.1: Parallel Reactor Scheme – Raw Material Usage

Parallel Reactor Scheme	Activity Loss	0		
Number of Parallel Trains	No Recycle	3	4	5
Cellulose Into SSF Reactor (lb/hr)	40,444	40,930	45,952	46,383
Cellulase In (lb/hr)	12,133	12,066	12,613	13,886
Fresh (lb/hr)	12,133	4,489	3,519	2,791
Recycled (lb/hr)	-	7,577	9,094	11,096
Total Cellulase Out (lb/hr)	12,133	3,098	3,098	2,764
Fresh DAP Added (lb/hr)	209	161	169	176
Fresh CSL Added (lb/hr)	1,659	1,260	1,322	1,373
DAP Recycled (lb/hr)	-	64	83	95
CSL Recycled (lb/hr)	-	524	688	782
Total DAP In (lb/hr)	209	224	253	271
Total CSL In (lb/hr)	1,659	1,783	2,010	2,155
Total DAP Out (lb/hr)	196	211	238	256
Total CSL Out (lb/hr)	1,613	1,737	1,958	2,102
Fresh Zymo (lb/hr)	330			
Ethanol wt% into SSF	0.0%	1.0%	1.2%	6.3%
Ethanol Produced (lb/hr)	16,973	17,275	19,440	19,637
Ethanol in product stream	16,848	16,642	19,245	19,475
Ethanol Production (MM Gal. / Yr)	22	22	25	25

Table F.3.1 (Continued)

Parallel Reactor Scheme	Activity Loss	5		
Number of Parallel Trains	No Recycle	3	4	5
Cellulose Into SSF Reactor (lb/hr)	40,444	42,909	46,043	46,503
Cellulase In (lb/hr)	12,133	13,024	13,457	13,661
Fresh (lb/hr)	12,133	5,096	3,640	3,033
Recycled (lb/hr)	-	7,928	9,817	10,628
Total Cellulase Out (lb/hr)	12,133	3,098	3,098	2,447
Fresh DAP Added (lb/hr)	209	170	172	176
Fresh CSL Added (lb/hr)	1,659	1,335	1,342	1,376
DAP Recycled (lb/hr)	-	69	85	97
CSL Recycled (lb/hr)	-	566	702	800
Total DAP In (lb/hr)	209	239	257	273
Total CSL In (lb/hr)	1,659	1,901	2,043	2,176
Total DAP Out (lb/hr)	196	225	242	258
Total CSL Out (lb/hr)	1,613	1,852	1,991	2,123
Fresh Zymo (lb/hr)	330			
Ethanol wt% into SSF	0.0%	1.0%	1.2%	6.5%
Ethanol Produced (lb/hr)	16,973	18,084	19,487	19,694
Ethanol in product stream	16,848	17,262	19,339	19,529
Ethanol Production (MM Gal. / Yr)	22	25	25	25

Table F.3.1 (Continued)

Parallel Reactor Scheme	Activity Loss	10		
Number of Parallel Trains	No Recycle	3	4	5
Cellulose Into SSF Reactor (lb/hr)	40,444	41,959	47,018	46,714
Cellulase In (lb/hr)	12,133	12,309	13,597	13,853
Fresh (lb/hr)	12,133	4,853	3,883	3,397
Recycled (lb/hr)	-	7,456	9,714	10,455
Total Cellulase Out (lb/hr)	12,133	3,098	3,098	2,208
Fresh DAP Added (lb/hr)	209	165	175	173
Fresh CSL Added (lb/hr)	1,659	1,296	1,366	1,346
DAP Recycled (lb/hr)	-	67	88	101
CSL Recycled (lb/hr)	-	549	724	830
Total DAP In (lb/hr)	209	232	263	273
Total CSL In (lb/hr)	1,659	1,845	2,090	2,176
Total DAP Out (lb/hr)	196	218	248	258
Total CSL Out (lb/hr)	1,613	1,797	2,036	2,123
Fresh Zymo (lb/hr)	330			
Ethanol wt% into SSF	0.0%	1.0%	1.2%	7.1%
Ethanol Produced (lb/hr)	16,973	17,696	19,894	19,794
Ethanol in product stream	16,848	16,980	19,681	19,622
Ethanol Production (MM Gal. / Yr)	22	23	25	25

Table F.3.1 (Continued)

Parallel Reactor Scheme	Activity Loss	15		
Number of Parallel Trains	No Recycle	3	4	5
Cellulose Into SSF Reactor (lb/hr)	40,444	42,909	48,083	46,915
Cellulase In (lb/hr)	12,133	12,515	13,999	13,445
Fresh (lb/hr)	12,133	5,096	4,247	3,640
Recycled (lb/hr)	-	7,419	9,753	9,805
Total Cellulase Out (lb/hr)	12,133	3,098	3,098	1,882
Fresh DAP Added (lb/hr)	209	169	180	173
Fresh CSL Added (lb/hr)	1,659	1,325	1,405	1,346
DAP Recycled (lb/hr)	-	69	92	102
CSL Recycled (lb/hr)	-	566	758	842
Total DAP In (lb/hr)	209	238	272	275
Total CSL In (lb/hr)	1,659	1,891	2,163	2,189
Total DAP Out (lb/hr)	196	224	256	260
Total CSL Out (lb/hr)	1,613	1,842	2,108	2,135
Fresh Zymo (lb/hr)	330			
Ethanol wt% into SSF	0.0%	1.0%	1.2%	7.4%
Ethanol Produced (lb/hr)	16,973	18,084	20,348	19,889
Ethanol in product stream	16,848	17,261	20,120	19,710
Ethanol Production (MM Gal. / Yr)	22	23	26	25

Table F.3.1 (Continued)

Parallel Reactor Scheme	Activity Loss	25		
Number of Parallel Trains	No Recycle	3	4	5
Cellulose Into SSF Reactor (lb/hr)	40,444	40,452	50,283	47,364
Cellulase In (lb/hr)	12,133	11,795	16,549	13,648
Fresh (lb/hr)	12,133	4,975	4,975	4,489
Recycled (lb/hr)	-	6,820	11,575	9,159
Total Cellulase Out (lb/hr)	12,133	3,098	3,098	1,407
Fresh DAP Added (lb/hr)	209	156	193	184
Fresh CSL Added (lb/hr)	1,659	1,226	1,510	1,434
DAP Recycled (lb/hr)	-	68	102	115
CSL Recycled (lb/hr)	-	561	838	947
Total DAP In (lb/hr)	209	225	295	299
Total CSL In (lb/hr)	1,659	1,787	2,348	2,381
Total DAP Out (lb/hr)	196	212	279	284
Total CSL Out (lb/hr)	1,613	1,741	2,291	2,327
Fresh Zymo (lb/hr)	330			
Ethanol wt% into SSF	0.0%	1.0%	1.2%	8.0%
Ethanol Produced (lb/hr)	16,973	17,085	21,293	20,102
Ethanol in product stream	16,848	16,703	21,058	19,920
Ethanol Production (MM Gal. / Yr)	22	24	27	26

Table F.3.1 (Continued)

Parallel Reactor Scheme	Activity Loss	50		
Number of Parallel Trains	No Recycle	3	4	5
Cellulose Into SSF Reactor (lb/hr)	40,444	49,394	54,133	48,355
Cellulase In (lb/hr)	12,133	14,737	23,111	14,780
Fresh (lb/hr)	12,133	7,765	7,401	7,644
Recycled (lb/hr)	-	6,972	15,710	7,136
Total Cellulase Out (lb/hr)	12,133	3,098	3,098	473
Fresh DAP Added (lb/hr)	209	184	209	185
Fresh CSL Added (lb/hr)	1,659	1,440	1,634	1,439
DAP Recycled (lb/hr)	-	84	120	131
CSL Recycled (lb/hr)	-	695	987	1,072
Total DAP In (lb/hr)	209	268	329	316
Total CSL In (lb/hr)	1,659	2,135	2,621	2,511
Total DAP Out (lb/hr)	196	253	312	300
Total CSL Out (lb/hr)	1,613	2,079	2,559	2,455
Fresh Zymo (lb/hr)	330			
Ethanol wt% into SSF	0.0%	1.1%	1.3%	11.5%
Ethanol Produced (lb/hr)	16,973	20,742	22,992	20,586
Ethanol in product stream	16,848	19,514	23,091	20,381
Ethanol Production (MM Gal. / Yr)	22	27	30	26

Table F.3.2: Parallel Reactor Scheme – Economics Summary

Parallel Reactor Scheme	Activity Loss	0		
Number of Parallel Trains	No Recycle	3	4	5
MESP4 ($/gal)	0.41	0.37	0.35	0.36
Change in MESP Over No Recycle		0.20	0.22	0.21
Max Savings ($/gal)		0.22		
No. of Trains			4	
Capital Costs ($/yr)				
Total Installed Equipment Cost ($/yr)	5,200,397	9,548,420	11,042,704	12,870,671
Indirect Costs ($/yr)	160,000	280,000	330,000	390,000
Total Project Investment ($/yr)	5,360,000	9,830,000	11,370,000	13,260,000
Operating Costs (cents/gal)				
CSL	5	4	4	4
Cellulase	25	9	6	5
DAP	1	0	0	0
Process Water	1	2	2	2
Capital Depreciation	2	1	1	0
Cooling Water	2	1	0	0
Chilled Water	-	2	2	3
Electricity	1	11	12	13
Capital Depreciation	1	2	2	3
Operating Costs ($/yr)				
CSL	1,120,000	840,000	910,000	920,000
Cellulase	5,520,000	2,040,000	1,600,000	1,270,000
DAP	120,000	100,000	100,000	100,000
Process Water	260,000	410,000	450,000	470,000
Zymo	1,850,000	170,000	130,000	410,000
Cooling Water	440,000	140,000	110,000	450,000
Chilled Water	-	410,000	490,000	2,710,000
Electricity	340,000	2,470,000	2,930,000	12,770,000
Capital Depreciation	260,000	480,000	550,000	640,000

Table F.3.2 (Continued)

Parallel Reactor Scheme	Activity Loss	5		
Number of Parallel Trains	No Recycle	3	4	5
MESP4 ($/gal)	0.41	0.38	0.35	0.37
Change in MESP Over No Recycle		0.19	0.22	0.20
Max Savings ($/gal)			0.22	
No. of Trains			4	
Capital Costs ($/yr)				
Total Installed Equipment Cost ($/yr)	5,200,397	10,154,618	11,156,571	12,823,396
Indirect Costs ($/yr)	160,000	310,000	330,000	390,000
Total Project Investment ($/yr)	5,360,000	10,460,000	11,490,000	13,210,000
Operating Costs (cents/gal)				
CSL	5	4	4	4
Cellulase	25	9	7	6
DAP	1	0	0	0
Process Water	1	2	2	2
Capital Depreciation	2	1	1	0
Cooling Water	2	1	0	0
Chilled Water	-	2	2	3
Electricity	1	11	11	13
Capital Depreciation	1	2	2	3
Operating Costs ($/yr)				
CSL	1,120,000	950,000	900,000	920,000
Cellulase	5,520,000	2,150,000	1,710,000	1,380,000
DAP	120,000	110,000	100,000	100,000
Process Water	260,000	470,000	430,000	460,000
Zymo	1,850,000	170,000	140,000	110,000
Cooling Water	440,000	150,000	110,000	110,000
Chilled Water	-	450,000	480,000	680,000
Electricity	340,000	2,730,000	2,880,000	3,250,000
Capital Depreciation	260,000	510,000	560,000	640,000

Table F.3.2 (Continued)

Parallel Reactor Scheme	Activity Loss	10		
Number of Parallel Trains	No Recycle	3	4	5
MESP4 ($/gal)	0.41	0.38	0.35	0.38
Change in MESP Over No Recycle		0.19	0.22	0.19
Max Savings ($/gal)			0.22	
No. of Trains			4	
Capital Costs ($/yr)				
Total Installed Equipment Cost ($/yr)	5,200,397	9,660,765	11,170,018	12,892,236
Indirect Costs ($/yr)	160,000	290,000	340,000	390,000
Total Project Investment ($/yr)	5,360,000	9,950,000	11,510,000	13,280,000
Operating Costs (cents/gal)				
CSL	5	4	4	4
Cellulase	25	10	7	6
DAP	1	0	0	0
Process Water	1	2	2	2
Capital Depreciation	2	1	1	1
Cooling Water	2	1	0	0
Chilled Water	-	2	2	3
Electricity	1	11	12	13
Capital Depreciation	1	2	2	3
Operating Costs ($/yr)				
CSL	1,120,000	860,000	910,000	900,000
Cellulase	5,520,000	2,210,000	1,770,000	1,550,000
DAP	120,000	100,000	100,000	100,000
Process Water	260,000	410,000	440,000	440,000
Zymo	1,850,000	180,000	150,000	130,000
Cooling Water	440,000	140,000	110,000	110,000
Chilled Water	-	410,000	480,000	680,000
Electricity	340,000	2,520,000	2,930,000	3,310,000
Capital Depreciation	260,000	480,000	560,000	640,000

Table F.3.2 (Continued)

Parallel Reactor Scheme	Activity Loss	15		
Number of Parallel Trains	No Recycle	3	4	5
MESP4 ($/gal)	0.41	0.38	0.36	0.38
Change in MESP Over No Recycle		0.19	0.21	0.19
Max Savings ($/gal)			0.21	
No. of Trains			4	
Capital Costs ($/yr)				
Total Installed Equipment Cost ($/yr)	5,200,397	9,844,335	11,463,770	12,965,156
Indirect Costs ($/yr)	160,000	300,000	350,000	380,000
Total Project Investment ($/yr)	5,360,000	10,140,000	11,810,000	13,350,000
Operating Costs (cents/gal)				
CSL	5	4	4	4
Cellulase	25	10	7	7
DAP	1	0	0	0
Process Water	1	2	2	2
Capital Depreciation	2	1	1	1
Cooling Water	2	1	0	0
Chilled Water	-	2	2	3
Electricity	1	11	12	13
Capital Depreciation	1	2	2	3
Operating Costs ($/yr)				
CSL	1,120,000	880,000	940,000	900,000
Cellulase	5,520,000	2,320,000	1,930,000	1,660,000
DAP	120,000	100,000	110,000	100,000
Process Water	260,000	420,000	450,000	430,000
Zymo	1,850,000	190,000	160,000	130,000
Cooling Water	440,000	150,000	110,000	110,000
Chilled Water	-	420,000	500,000	680,000
Electricity	340,000	2,600,000	3,050,000	3,350,000
Capital Depreciation	260,000	490,000	570,000	650,000

Table F.3.2 (Continued)

Parallel Reactor Scheme	Activity Loss	25		
Number of Parallel Trains	No Recycle	3	4	5
MESP4 ($/gal)	0.41	0.41	0.38	0.42
Change in MESP Over No Recycle		0.17	0.19	0.15
Max Savings ($/gal)			0.19	
No. of Trains			4	
Capital Costs ($/yr)				
Total Installed Equipment Cost ($/yr)	5,200,397	10,506,208	12,185,149	13,711,443
Indirect Costs ($/yr)	160,000	310,000	360,000	410,000
Total Project Investment ($/yr)	5,360,000	10,820,000	12,550,000	14,120,000
Operating Costs (cents/gal)				
CSL	5	4	4	4
Cellulase	25	11	8	8
DAP	1	0	0	0
Process Water	1	2	2	2
Capital Depreciation	2	1	1	1
Cooling Water	2	1	0	0
Chilled Water	-	2	2	3
Electricity	1	12	12	14
Capital Depreciation	1	2	2	3
Operating Costs ($/yr)				
CSL	1,120,000	960,000	1,010,000	960,000
Cellulase	5,520,000	2,590,000	2,260,000	2,040,000
DAP	120,000	110,000	110,000	110,000
Process Water	260,000	470,000	480,000	450,000
Zymo	1,850,000	220,000	190,000	170,000
Cooling Water	440,000	160,000	120,000	120,000
Chilled Water	-	460,000	550,000	720,000
Electricity	340,000	2,880,000	3,360,000	3,680,000
Capital Depreciation	260,000	530,000	610,000	690,000

Table F.3.2 (Continued)

Parallel Reactor Scheme	Activity Loss	50		
Number of Parallel Trains	No Recycle	3	4	5
MESP4 ($/gal)	0.41	0.41	0.42	0.49
Change in MESP Over No Recycle		0.16	0.15	0.09
Max Savings ($/gal)			0.16	
No. of Trains			3	
Capital Costs ($/yr)				
Total Installed Equipment Cost ($/yr)	5,200,397	10,937,509	13,350,522	14,143,532
Indirect Costs ($/yr)	160,000	330,000	400,000	430,000
Total Project Investment ($/yr)	5,360,000	11,270,000	13,750,000	14,570,000
Operating Costs (cents/gal)				
CSL	5	4	4	4
Cellulase	25	14	12	13
DAP	1	0	0	0
Process Water	1	2	2	2
Capital Depreciation	2	1	1	1
Cooling Water	2	1	0	0
Chilled Water	-	2	2	3
Electricity	1	11	13	16
Capital Depreciation	1	2	2	3
Operating Costs ($/yr)				
CSL	1,120,000	990,000	1,120,000	960,000
Cellulase	5,520,000	3,700,000	3,530,000	3,480,000
DAP	120,000	110,000	130,000	110,000
Process Water	260,000	450,000	500,000	400,000
Zymo	1,850,000	280,000	270,000	290,000
Cooling Water	440,000	170,000	140,000	130,000
Chilled Water	-	480,000	620,000	730,000
Electricity	340,000	3,090,000	3,910,000	4,090,000
Capital Depreciation	260,000	550,000	670,000	710,000

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