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THE CONVERSION OF CARBOHYDRATES FROM MICROALGAE AND CORN STOVER INTO BUILDING BLOCK CHEMCIAL ACIDS

By

Andrew Joseph Kohler

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This document, submitted in partial fulfillment of the requirements for the degree from the University of North Dakota, has been read by the Faculty Advisory Committee under whom the work has been done and is hereby approved.

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Andrew Kohler

July 7, 2020

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ABSTRACT

Depletion of petroleum-based resources requires the development of high-valued chemicals from renewable sources. Previous work at the University of North Dakota demonstrated that four building block chemical acids, namely lactic, levulinic, acetic, and formic acids, can be generated by transforming cellulose into glucose followed the catalytic decomposition into the target acids. The most important of these chemicals is lactic acid, which can be used as a building block to produce a variety of biodegradable plastics. Traditional fermentation methods for lactic acid synthesis have proven costly, making it difficult for its subsequent products to compete economically. Alternatively, the use of a Lewis acidic heterogeneous catalyst, Sn-Beta, can be used to produce lactic acid in less time and more economically. Levulinic acid is emerging as an important chemical intermediate which can be transformed into a number of more valuable products, including biodegradable polyesters.

In the present study, Sn-Beta catalyst was used to convert the two most abundant sugars in lignocellulosic biomass, glucose and xylose, to lactic acid at yields of 13 wt% and 19 wt% of inlet carbon, respectively with yields of levulinic acid of 18 wt% and 0.8 wt%, respectively. However, the addition of $CaSO_4$ to the aqueous reaction solvent resulted in increased lactic acid yields of 68 wt% and 50 wt% of inlet carbon for glucose and xylose, respectively. Application of this reaction system to sugars extracted from corn stover had a comparable lactic acid yield of 60 wt% while microalgae extracted sugars decomposition appeared to be inhibited leading to a yield of only 8.2 wt% of inlet carbon. The addition of $CaSO_4$ to the aqueous reaction solvent neutralizes the Brønsted acid sites on the surface of Sn-Beta, decreasing the formation of unwanted dehydration products, and adds an additional Lewis acid, increasing the formation of lactic acid.

In a second study, the purification of lactic, levulinic, acetic, and formic acids produced from Sn-Beta was examined. The target acids were reacted with a basic tertiary amine, trioctylamine (TOA), to form an organic soluble ion-pair. Two active alcohol diluents, isoamyl alcohol and 1-octanol, were examined based on their ability to efficiently solvate the produced ion-pair. For each diluent, optimum acid composition, organic/aqueous phase ratio and TOA composition were determined using a central composite design. At optimal conditions isoamyl alcohol was observed to have 50% higher efficiency at extracting lactic acid and levulinic acid. Despite this advantage, similar boiling points between isoamyl and the short chain acids – acetic and formic – decreased the downstream separation efficiency, leading to the conclusion that 1-octanol is the more economical diluent.

Finally, a comparative scoping study was performed to determine the technological and economic feasibility of a world scale processing plant for the Sn-Beta conversion of corn stover-derived sugars. Two separate processes were developed, one based upon the preliminary Sn-Beta yields that produces higher levels of levulinic acid while the other was based upon the addition of calcium sulfate, yielding primarily lactic acid. AACE Class 4 cost estimates and economic analyses were conducted for each process. Net present values @ 20% were estimated as \$26 million ± 40% and \$250 million ± 40% for the base Sn-Beta and Sn-Beta + $CaSO_4$ alternatives, respectively.

Chapter I - Introduction

1. Introduction

One of the greatest global challenges of the next decade is to reduce the anthropogenic carbon footprint [1]. In order to best mitigate the looming effects of global climate change, the Intergovernmental Panel on Climate Change (IPCC) projects that global carbon dioxide emissions will need to be reduced by 45% through 2030 and reach net zero between 2045 and 2055 [2]. This in turn requires a transition away from fossil sources like coal, petroleum, and natural gas, which account for 93% of the CO_2 emissions in the United States [3].

Ripples of this transition extend beyond energy generation to the production of useful chemicals. For example, the global polymer industry is reliant on chemicals like polyethylene (PE), polypropylene (PP), nylon, polyester (PS), polytetrafluoroethylene (PTFE), and epoxy which are produced from petroleum-based hydrocarbons [4]. One of the most promising alternatives to petroleum-based polymers is the implementation of polymers derived from biomass [5]. These materials, often known as bioplastics, are favored due their ability to reduce carbon emissions. Additionally, some bioplastics like PHA and PLA can biodegrade, making them favorable for single-use applications.

A separate problem that petroleum-based plastics possess is disposability. Commonly, plastics are disposed through landfilling, recycling, or incineration [6]. Space for landfills are quickly decreasing while incineration contributes to air pollution. Plastics that are not properly disposed degrade slowly to micro-plastics, which can be toxic and persistent pollutants in most ecosystems [6].

The use of biodegradable plastics when possible provides numerous environmental advantages including soil enrichment, environmental stabilization, and elimination of environmental removal costs [6]. Currently, the most economically competitive plastics are polyhydroxyalkanoate (PHA) and polylactic acid (PLA) which are used for food packing and biomedical implants.

2. Lactic Acid

Currently, lactic acid is almost exclusively produced through fermentation. A wide range of microbial species have been shown to produce lactic acid from renewable sources. Those species include, but are not limited to, bacteria, fungi, yeast, microalgae, and cyanobacteria [7]. Though batch fermentation is the most widely used method to produce lactic acid, it has low productivity due to long batch times and low cell concentration. The process is also often bottlenecked through inhibitors in the substrate and the final product. Substrates for fermentation are obtained from carbohydrate sources like lignocellulosic and starchy biomass. Physiochemical and enzymatic treatments for carbohydrate saccharification also significantly contribute to the cost of lactic acid production [7]. To help combat these high processing costs, it may be possible to replace the fermentation step with a more efficient catalytic reaction [8-10].

The traditional applications of lactic acid are in the food industry. It is often used as a flavoring agent to provide a mild acidic taste. Process foods such as candy, beers, and dairy products utilize lactic acid to provide a pH buffer as well as an inhibitor for bacteria spoilage. Its conjugate base, lactate, can be used to form esters which act as emulsifying agents. These agents can then be applied to bakery products like cakes to improve texture and volume [11].

Lactic acid has also been used in pharmaceutical and cosmetic products. In the sodium salt form, lactic acid can be used for dialysis. When it is paired with a calcium ion, the salt can be administered for calcium deficiency therapy and as anti-caries agents [12]. Cosmetically, ethyl lactate provides an active ingredient for acne medication. Lactic acid can also be esterified with a phenol to give phenoxypropionic acid which is used in the production of herbicides [12].

The main application for lactic acid is for the production of poly-lactic acid (PLA), a biodegradable polymer. Originally the high cost, low availability, and low molecular weight of PLA limited its use to medical applications such as implants, tissue scaffolds, and internal sutures [13]. However, improvements in the production of high-density PLA has led to widespread use as a replacement for polyethylene terephthalate (PET).

PLA is attractive since it is compostable and made from renewable sources. This helps to address two societal problems: plastics building up in landfills and depletion of fossil fuels. Low toxicity of the polymer makes it ideal to be used in single-use food packages [14].

Lactic acid is present in one of two optical isomers, L- and D- lactic acid. As shown in Figure 1 it has a chiral center at the middle carbon atom. The ratio of these stereoisomers can greatly affect the material properties of PLA, changing its potential applications [13]. Typically, PLA allows for less carbon dioxide, water, and air to permeate through it than polystyrene [15]. When unoriented, PLA has good strength and stiffness, but has been shown to be brittle. When oriented, PLA has improved or comparable properties to polystyrene or polyethylene terephthalate [16].

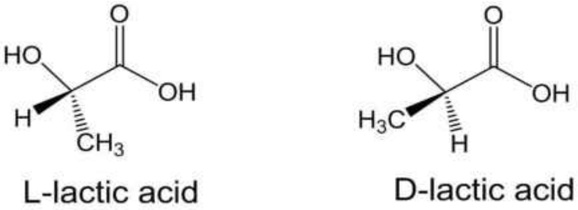

Figure 1. Stereoisomers of lactic acid

For commercial applications, PLA is a copolymer of poly(L-lactic acid) and poly(D,L-lactic acid), which is an altering sequence of the two lactic acid isomers. These polymers are produced from intermediates L-lactide and D,L-lactide, respectively, which are formed through the dual dehydration of two lactic acid molecules, forming a six-member ring. Current fermentation processes produce almost exclusively L-lactic acid, making it the more prominent isomer. Synthetic pathways produce a racemic mixture of lactic acid, which have previously been uneconomical [17]. When L-lactic acid is utilized at a purity greater than 90%, PLA will be crystalline. If the isomer concentration is any less, the polymer will be amorphous. Similarly, as the L-lactic acid concentration decreases so does the glass transition temperature of PLA [18].

Despite the advantages of PLA only 450,000 metric tons are produced per year, which is an almost insignificant amount compared to the over 200 million metric tons of plastic produced annually [19]. Lag in PLA production is mostly due to the high cost of producing lactic acid. It is anticipated that in order for PLA to become more competitive with fossil-fuel based plastics lactic acid prices will have to drop to $0.8 per kilogram [19].

Potentially more efficient synthetic pathways, which are explored in this thesis, form a racemic mixture of lactic acid, limiting the applications of PLA until such time that an efficient separation method for the L- and D- forms has been developed. Alternately, new polymerization strategies that can use the racemic mixture, but still produce a crystalline polymer may be investigated.

An alternative renewable polymer pathway is to convert lactic acid into acrylic acid or propylene glycol [20, 21]. Acrylic acid has wide polymer applications, with current uses in surface coatings, textiles, adhesives, paper treatment, baby diapers, and detergents [22]. Current acrylic acid production uses petroleum-sourced propylene as the primary raw material [22]. Similarly, propylene glycol production requires petroleum-derived propylene oxide and is desired for polyester resins, de-icing fluids, and antifreeze [20]. The major hurdle for both pathways is the high cost of lactic acid production through fermentation [21].

3. Levulinic Acid

Levulinic acid has been a chemical of interest for many years due to its potential to be converted into valuable products. An industrial scale process was developed by Biofine Renewables to produce levulinic acid using the cellulose and hemicellulose in agricultural residues [23]. The process hydrolyzes the carbohydrates to monomeric sugars and then dehydrates the sugars to form 5-hydroxymethylfurfural (5-HMF). 5-HMF is then recovered and sent to a second reactor yielding levulinic acid.

One of the main applications for levulinic acid is to form γ-valerolactone (GVL) which can be used as a liquid solvent or as a transition chemical to other products. Using a 5 wt% Ru/C catalyst at 150° C, 97% of levulinic acid can by hydrogenated giving GVL [24]. The hydrogenation reaction showed improved results using a Ru/SiO_2 catalyst in supercritical carbon dioxide, increasing the yield to 99% [25]. GVL has the potential to be used as a fuel additive as it has shown similar properties to ethanol when mixed with octane [26]. Its chemical and physical properties gives it applications as a green solvent since it is renewable, non-toxic, and biodegradable [27].

In the gas phase, GVL can react with formaldehyde over an alkaline hydroxide catalyst to form α-methylene-γ-valerolactone (MGVL). This reaction achieves a high selectivity of over 95%. MGVL is of interest since it has a structure similar to that of methyl methacrylate. The lactone structure available in MGVL can give its homopolymer a glass transition temperature over 100° C greater than poly-methyl methacrylate [24].

It has also been proposed to convert GVL to a mixture of pentenoic acids by way of a solid acid catalyst, like SiO_2/Al_2O_3 [28]. The same acid catalyst can then be used to decarboxylate the acids, forming n-butenes. The butenes can then be subsequently oligomerized to form C_8^+ alkenes which can be used for fuel applications.

Future research at the University of North Dakota will focus on hydrogenating GVL into 1,4-pentanediol (1,4-PDO) [29]. 1,4-PDO can then be used as a monomer for high strength polyester (polyester polyols) production [30]. This class of biodegradable polymers has application potential in textiles, packing materials, and engineering plastics.

A biofuel application of levulinic acid is to convert it to a kerosene type transportation fuel. This process begins by reducing levulinic acid to valeric (pentanoic) acid. Through the use of a catalyst at 350-400°C, two valeric acid molecules can be fused together through a ketonic decarboxylation reaction, eliminating a carbon dioxide and a water molecule in the process [31]. This reaction results in a nine-carbon ketone, 5-nonanone. The ketone can then be hydrogenated to form an alcohol, 5-nonanol. The hydroxyl group is then removed through a dehydration reaction, forming 4-nonene which can be hydrogenated to form nonane. The pathway from valeric acid to nonane can be seen on Figure 2.

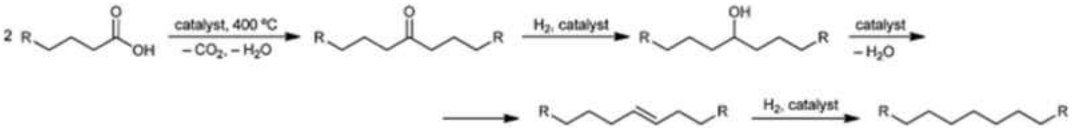

Figure 2. Reaction Pathway from Valeric Acid to Nonane

Zirconium oxide has shown high selectivity and stability in the ketonic decarboxylation reaction of valeric acid, forming the intermediate 5-nonanone. Oxygen can be then removed through the simultaneous hydrogenation/dehydration/hydrogenation reaction through a platinum/alumina catalyst. Sequential use of these catalysts have shown complete conversion of valeric acid, with 80% selectivity towards n-nonane and 10% towards C_{10}-C_{15} hydrocarbons [31].

An acid, like acetic anhydride and phosphoric acid, can be used to dehydrate levulinic acid to form α-angelica lactone, a monomer used to make biodegradable polyesters. Similar to polylactic acid, it is desired as a medical implant due to its biocompatibility and its fast degradability [32]. Levulinic acid is also used in the production diphenolic acid.

When mixed with two equivalents of phenol over an acidic catalyst, levulinic acid reacts to form diphenolic acid. This can be used as a substitute to bisphenol A (BPA), a petroleum-based chemical or as an intermediate in the production of polycarbonates [33]. BPA is currently used to add strength and resilience in polycarbonate plastics and as a part of the epoxy resin used to line metal food and beverage cans [34]. Current demand for polycarbonates is on the rise due to their stability and unique properties, making the production of a renewable diphenolic acid an attractive pathway.

A more recent application for levulinic acid is oxidation to form succinic acid. V_2O_5 has often been used as an oxidization catalyst to achieve this reaction. Dunlop, Smith, and Shelbert have patented a process converting levulinic acid over V_2O_5 at temperatures around 360-400˚ C [35]. A similar patent by Van Es achieves the reaction at milder conditions of 40-60˚ C while adding nitric acid as an additional oxidant [36]. Parvulescu was able to obtain higher yields using ruthenium-based magnetic nanoparticles as a solid catalyst at 150-180˚ C [37].

Succinic acid is considered a top 12 value-added chemical with an annual market of $400 million in 2016 [38]. Its applications span a wide range of industries including solvents, foods, pharmaceuticals, and polymers. It is primarily used as a pH adjustor, anti-microbial agent, and a flavoring agent in the food industry. Pharmaceutically, it is used as an additive to vitamins and can be used to help produce antibiotics. Other minor applications include use as a foaming agent, surfactant, and corrosion inhibitor to prevent pitting [38].

4. Scope of the Research in this Thesis

Previous research at the University of North Dakota has focused on lactic and levulinic acid production from glucose, a monomeric sugar which can be derived from biomass, using a Sn^{2+} doped beta zeolite (Sn-Beta) [39]. Lewis acid sites on Sn-Beta can be used to catalyze a retro-aldol reaction, which splits the sugar into two C_3 trioses that can be isomerized to lactic acid. Conversely, Brønsted acid sites on the catalyst surface can dehydrate the sugar to 5-hydroxymethylfurfural (5-HMF) which can be partially rehydrated to form levulinic acid. Preliminary results revealed modest yields of 7.6 wt% lactic acid and 19 wt% of levulinic acid from glucose on an inlet carbon basis [39].

The following chapters document work that builds upon the previous Sn-Beta work to advance its industrial process potential. Chapter II, “Catalytic Formation of Lactic Acid from Biomass Derived Monosaccarides through Sn-Beta Formed by Impregnation,” presents a study on the optimization of the reaction conditions. In addition to glucose, it evaluates the reaction potential of other biomass-derived sugars like xylose and galactose. Two separate biomass feedstocks, corn stover and microalgae, are used to evaluate their conversion potential.

Chapter III, “Reactive Extraction Optimization for Aqueous-Phase Carboxylic Acids Produced from Biomass,” documents experiments performed to generate data required to design the processing steps needed to purify the target carboxylic acids immediately following the Sn-Beta reaction. Initial screening and optimization studies were conducted to determine the most suitable conditions for liquid-liquid extraction of the product carboxylic acids. This was followed by an economic comparison of the two most suitable process steps for purification.

Chapter IV, “Techno-Economic Analysis of a Process for the Aqueous Conversion of Corn Stover into Lactic and Levulinic Acid through Sn-Beta + CaSO4 Catalysis” presents two process alternatives for the conversion of corn stover. One is optimized for high purity lactic production and one optimized for a balanced production of high purity lactic and levulinic acids. This section takes the reaction and separation data found experimentally in Chapters II and III, respectively, and applies it to a potential industrial process. Process profitability is then determined based on the capital investment, operating costs, and generated revenues.

Chapter II – Catalytic Formation of Lactic and Levulinic Acids from Biomass Derived Monosaccharides through Sn-Beta Formed by Impregnation

1. Introduction

The effects of global climate change have sparked an effort to transition from fossil-fuel based energy to renewable, lower emitting sources. This shift away from a petroleum-based economy trickles down to commodity and specialty chemicals that are produced from non-renewable sources. Lactic acid is one potential chemical that can be produced renewably and applied to a wide-range of industries including cosmetics, food, pharmaceuticals, and plastics [12-14, 21]. Its most common application is to produce poly[lactic acid] (PLA), a biodegradable plastic. Though this process currently requires pure L-lactic acid, although racemic mixtures can also produce biodegradable polymers through intermediate conversion to acrylic acid or propylene glycol [21].

Fermentation is currently the primary methods of producing lactic acid. However, this method has proven to be costly, hindering the ability of its subsequent products to compete with non-degradable fossil fuel analogs [19]. A potential cost saving measure is to produce lactic acid chemically instead of biologically. Previous work at the University of North Dakota has focused on catalytically transforming glucose into lactic acid and its derivatives through the use of a tin-doped beta zeolite (Sn-Beta) [39] with a subsequent preliminary study of the comparable decomposition of xylose (Foerster, 2020).

Levulinic acid has been a chemical of interest for many years due to its potential platform chemical from biomass [40]. An industrial scale process was developed by Biofine Renewables to produce levulinic acid using the cellulose and hemicellulose in agricultural residues [23]. The process hydrolyzes the carbohydrates to monomeric sugars and then dehydrates the sugars to form 5-hydroxymethylfurfural (5-HMF). 5-HMF is then recovered and sent to a second reactor yielding levulinic acid.

Starting from this previous work, the present study had the following objectives:

1 Advance the conversion efficiency of the reaction for both glucose and xylose into lactic and levulinic acids through a careful study of the effect of residence time, reaction temperature, and substrate sources on catalyst selectivity and product yields.

2 Evaluate the optimum reaction conditions for monomeric sugar mixtures that simulate the typical composition of sugars extracted from: 1) microalgae and 2) corn stover.

3 Apply the optimum reaction conditions to actual sugar mixtures derived from microalgae and corn stover to evaluate effectiveness.

Sn-Beta is a catalyst that combines tin with a Si/Al catalyst. This catalyst facilitates multiple reaction pathways in converting monosaccharides into lactic acid or levulinic acid. At temperatures as low as 110˚C, Lewis acid sites on Sn-Beta catalyzes the isomerization of sugars between their aldo- (e.g. glucose) and keto-form (e.g. fructose) [8, 41-43]. This isomerization can be critical since at temperatures above 150° C the monosaccharides undergo retro-aldol condensation [9]. As seen in Figure 1, if a hexose is in the keto-form it will split to two trioses, 1,3-dihydroxyacetone (DHA) and glyceraldehyde (GLA), which can then be isomerized into lactic acid [44, 45]. Conversely, retro-aldol condensation of aldohexoses forms a C_2 and C_4 fragment which cannot produce lactic acid (not shown).

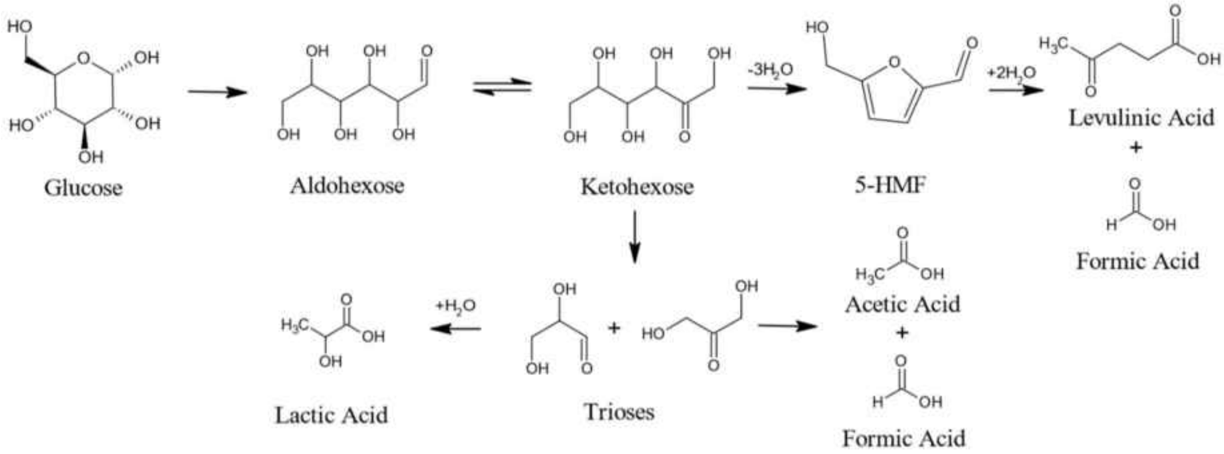

Figure 3. Reaction pathway for hexoses using Sn-Beta

Lactic acid yields can be hindered by the formation of multiple side-products. In the presence of Brønsted acid sites, ketohexoses can undergo a triple dehydration reaction to form 5-hydroxymethylfurfural (5-HMF) [46]. Double rehydration of 5-HMF in water then yields levulinic and formic acid, as seen in Figure 3. At hydrothermal conditions (>170° C), saccharides undergo intra- and intermolecular dehydration reactions, forming aromatized clusters [47]. Once these clusters reach the super-saturation point, burst nucleation occurs creating solid hydrochar microspheres [47].

The Sn-Beta facilitated reaction of xylose, the primary component of hemicellulose, has not been studied to the same degree as glucose. Holm found that in a methanol reaction solvent pentoses like xylose yielded about 10% less methyl lactate on an inlet carbon basis [44] than the comparable reactions of hexoses. This is due to the formation of C_2 and C_3 fragments produced through retro-aldol condensation (Figure 4), leaving only 60% of the carbon available to transform into lactic acid. Like hexoses, Brønsted acid sites competitively dehydrate the pentoses, forming furfural [48, 49].

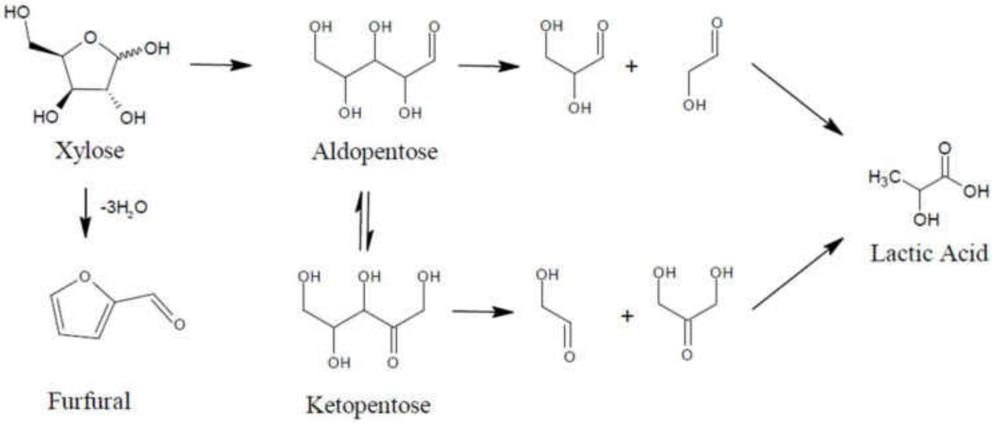

Figure 4. Reaction pathway for pentoses using Sn-Beta

Compared to metal dopants like titanium and zirconium, tin produces the most efficient Lewis acid sites for the transformation of glucose to lactic acid when incorporated into the framework of a beta zeolite [8]. This Lewis acidity is essential to increasing the selectivity toward the desired retro-aldol reaction [8]. Additionally, a tetravalent Lewis acid, like tin, can be used to replace the trivalent aluminum within the beta-zeolite. This removes a source of Brønsted acid sites and, subsequently, decreases the selectivity of undesired dehydration reactions.

Typically, Sn-Beta can be prepared by one of two methods: hydrothermal synthesis and post-synthesis. Hydrothermal synthesis can be considered the bottom-up approach where tin ions are incorporated into the zeolite framework during nucleation. The size of the ions can inhibit the zeolite nucleation, requiring catalyst synthesis to take up to 40 days [10] and limit the tin content on the catalyst [50]. Hydrothermal synthesis can also produce undesired metal oxide particles, providing transport limitations [46].

Conversely, post-synthesis preparation is considered the top-down approach and takes already formed beta zeolites and adds tin to the framework. This can be accomplished by dealuminating the zeolite with 65 wt% HNO_3 and then adding tin to the newly formed defect sites through solid-state ion-exchange [10, 50]. However, this significant nitric acid requirement makes large scale production very costly. A similar post-synthesis approach to solid-state ion-exchange is impregnation. In impregnation, the beta zeolite is mixed with a tin salt in 1 M HCl. This allows for the zeolite to simultaneously be dealuminated and have tin added to the defect sites. This results in a significantly lower acid requirement than the traditional solid-state ion-exchange.

Various modifications can be made to Sn-Beta to improve lactic acid yield. Both adding zinc [51] and grafting an amino group [52] have been attempted to increase the basicity of the catalyst and combat the Brønsted acid sites that produce dehydration products. These attempts have yielded results of significantly increasing lactic acid yield while decreasing the production of 5-HMF. Additional Lewis acids, like lead, have shown superior glucose isomerization abilities compared to tin. When combined as a dual-function catalyst lactic acid yields increase from 22% to 52 wt% of inlet carbon [53]. Similarly, when indium was used for the retro-aldol reaction of fructose and tin for the subsequent triose isomerization in methanol, a pseudo methyl lactate yield of 90% was observed [54]. Yang added a metal oxide, WO_3, to act as a promoter [41]. Doing so decreased silanol defects in the zeolite which catalyze the dehydration reaction, increasing methyl lactate yield from 25% to 52 wt% of inlet carbon.

The work reported here focused on the performance of a base Sn-Beta catalyst synthesized through impregnation. To the best of our knowledge, this catalyst preparation procedure has not previously been attempted for Sn-Beta. This study specifically looks at Sn-Beta reaction selectivity of glucose and xylose, the most naturally abundant hexose and pentose, respectively. Optimum reactions times and temperatures are initially studied using model sugar solutions.

Information from the model reaction studies were then used to compare catalyst performance for actual biomass sources, allowing for potential discrepancies, such as pH and catalyst inhibitors, to be identified. Corn stover is the most abundant biomass residue in the United States, consisting of the leftover cobs, husks, leaves, and stalks after harvest [55]. Corn stover is a lignocellulosic biomass consisting of cellulose, hemicellulose, and lignin. Both carbohydrates, cellulose and hemicellulose, can be extracted by hydrolysis yielding two primary sugars, glucose and xylose, respectively. Lignocellulosic biomass typically requires more intensive treatment to release these sugars than starch-based biomass sources, like corn kernels and microalgae (e.g. *Chlorella vulgaris)*, making it more difficult to economically process these materials in conventional processes like fermentation [56].

Microalgae are an attractive source for carbohydrates due to their short growth cycles, high carbohydrate concentrations, and relatively easy accessibility through acid extraction [57]. In previous work, Hammann [2019] found that almost all of the carbohydrates that can be extract from microalgae hydrolyze to glucose and galactose [58]. Thus, microalgae may be an attractive feedstock for the production of lactic acid.

2. Methods

2.1 Reactants and Catalysts Materials

Glucose (99.5% purity), galactose (99% purity), xylose (99% purity), fructose (99% purity), mannose (99% purity), tin (II) chloride (98% purity), sodium hydroxide (98% purity), calcium chloride (97% purity), and magnesium sulfate (97% purity) were obtained from Sigma Aldrich (St. Louis, USA). Hydrochloric acid (34-37% purity), sulfuric acid (96% purity), calcium carbonate, hemihydrate calcium sulfate was obtained from Fisher Scientific (Waltham, USA). Hydrogen form, beta zeolite with a SiO_2/Al_2O_3 ratio of 300 (CP811C-300) was purchased from Zeolyst International (Kansas City, USA). Accellerase enzymes were obtained from Genencor. Ultrapure water was obtained from an in-house ultra milli-Q filter system. Compressed nitrogen (99.99 % purity) was purchased from Praxair (Grand Forks, USA).

2.2 Catalyst Preparation

Sn-Beta was produced using a modified post-synthesis approach. The base beta zeolite was initially calcined at 600 ˚ C for 6 hours to remove any impurities. Beta zeolite was then impregnated with $SnCl_2$ in a 1.0 M HCl solution in a ratio of 10 g zeolite: 1 g $SnCl_2$: 100 mL of solution. The impregnation solution was stirred at 500 rpm overnight and then dried in an oven at 120˚ C for 48 hours. A second calcination step was then performed at 400˚ C for 8 hours.

2.3 Batch Reactor

All batch reaction experiments were conducted in a bench scale 500 mL Parr 5500 compacted series reactor (Moline, USA) configured as shown in Figure 5. The head of the reactor was mounted on a stand. The full reactor vessel was lifted up to the head and sealed with a clamp. A gas inlet on the top of the reactor was used to inject nitrogen into the reactor to displace any residual oxygen in the vapor space prior to the start of each reaction. This was done to prevent side reactions as a result of oxidation. The gas was vented through an outlet valve to a fume hood. A rate adjustable motor driven stirring shaft was used to insure complete mixing of the reactants and catalyst during each reaction. A thermocouple inside the reactor was used to monitor and provide feedback on the reaction temperature. A 4848 Parr series controller was used to control the reaction temperature at 200° C using the readout from the thermocouple. The reactor was placed within a heating jacket with an electrical resistance heater that was altered by the controller. Internal reactor pressure was monitored with an analog gauge attached to the head of the reactor.

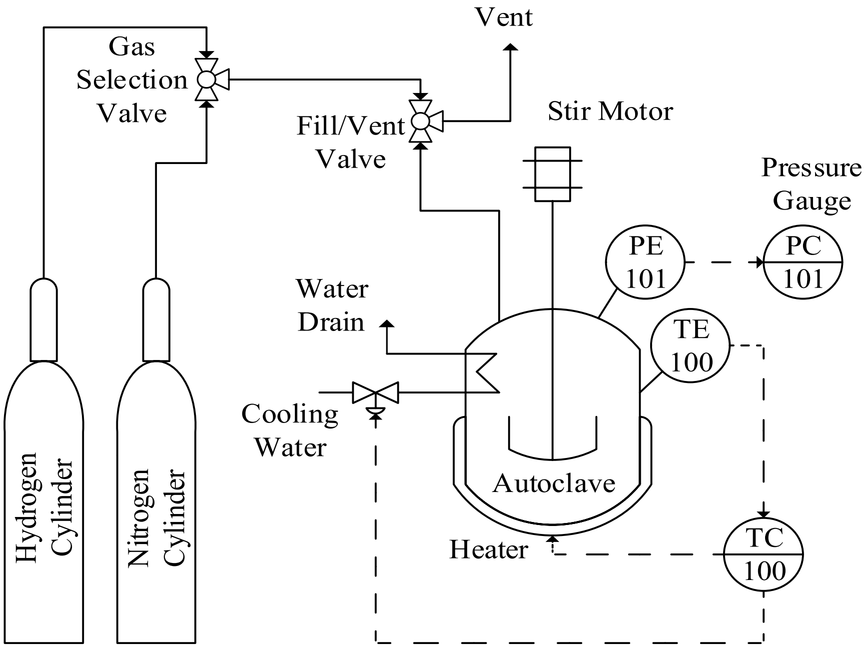

Figure 5. Simplified schematic of autoclave reactor used for all experiments.

2.4 Reaction

Each Sn-Beta catalytic reaction utilized ten grams of catalyst and five grams of sugars mixed in 300 mL of water. Water volume did not exceed this level to maintain safe reaction pressures as the 500 mL reactor reached the operating temperature. Once the reactor was filled and sealed, it was purged of oxygen by five pressure/vent cycles from 140 kPa to atmosphere using lab-grade nitrogen. The reactor was then brought up to an initial pressure of 2200 kPa using nitrogen and a five-minute pressure check was conducted.

Once it was confirmed that there were no leaks, the reactor was placed in the heating jacket. When the desired operating temperature was reached (100-200° C) it was held constant for the specified reaction time. At the end of the specified time, the reaction vessel was removed from the heating jacket and the vessel was allowed to cool.

When the reaction was complete, the products were removed from the reactor, weighed, and centrifuged. The liquid product was decanted and analysed while the remaining solids were dried overnight at 60°C. Dried solids were then weighed to estimate the accumulated mass of hydrochar.

2.5 Biomass Hydrolysis

In addition to testing pure sugars, trials were also conducted with sugars extracted from microalgae and corn stover. Sugars from microalgae were extracted using a process optimized by Hammann [58]. This process utilized 4 wt% sulfuric acid (H_2SO_4) in water solvent at a loading of 20 mL/g_{algae}. Hydrolysis occurred using a Consolidated Sterilall electrically heated double wall sterilizer type autoclave (Boston, USA) at 120° C for 20 minutes. Resulting hydrolysate was neutralized with excess calcium carbonate ($CaCO_3$)

and filtered. The solution was then concentrated using a RE500 rotary vacuum evaporator (Yamato Scientific, Santa Clara, USA).

Sugars from corn stover were extracted following slightly modified conditions based on those outlined by the National Renewable Energy Laboratory (NREL) [59]. Corn stover was initially treated with a 0.4 wt% sodium hydroxide solution at a 10 mL/g_{solids} to remove acetate from the hemicellulose. The mixture was then heated at 80° C for 1 hour in a 500 mL Parr reactor and then mixed with 0.4 wt% sulfuric acid in water at a loading of 11 mL/g_{solids} for 20 minutes at 160° C to hydrolyse the hemicellulose. Resulting hydrolysate was neutralized with excess calcium carbonate ($CaCO_3$) and filtered. Collected solids after acid hydrolysis were treated with accellerase enzymes to liberate glucose from the remaining cellulose. Enzymes were loaded at a 10 $mg_{enzymes}/g_{cellulose}$ ratio with corn stover in a 20 solids wt% solution. The mixture was then placed on a MAXQ 4000 shaker table (Fisher Scientific, Waltham, USA) for 72 hours at 48° C and 250 RPM. The resulting solution was filtered and combined with acid hydrolysate.

2.6 Analysis

Organic acids were identified using GC-MS (6890NGC, 5975 MS: Agilent Technologies, Santa Clara, USA). Sample injections were conducted in splitless mode at 250° C for 0.5 minutes. Separation occurred on a 30-m long DB-5 column, with 0.25 mm internal diameter and 0.25 μL film thickness (J&W Scientific, Folsom, USA). Temperature was programmed to hold at 35° C for 1 min., with subsequent gradients of 20° C/min to 85° C and 5° C/min to 320° C. Acids were derivatized to butyl esters using BF_3/1-Butanol by the method optimized by St'ávová [60]. The MS was used in full scan mode (50-600 m/z) at a dwell time of 50 ms. Samples were ionized by electron ionization with an energy of 70 eV.

Liquid reaction products were quantified using High Performance Liquid Chromatography (Agilent Technologies 1200 series, Santa Clara USA). Separation occurred in an Agilent Hi-Plex H organic acid column with a mobile phase of 5 mM H_2SO_4 at 0.6 mL/min and 40°C. Each sample was measured with a refractive index detector (G1362A: Agilent Technologies, Santa Clara, USA) for 45 minutes.

3. Results and Discussion

3.1 Catalyst Synthesis and Evaluation of Glucose Yields

Preliminary experiments were conducted to screen potential variables and find optimum values for the catalyst synthesis through impregnation method and for the reaction conditions associated with the conversion of glucose into lactic acid. It was determined that the most influential factors in lactic acid yield were the ratio of glucose to catalyst, the ratio of silica to alumina in the base beta-zeolite (SiO_2/Al_2O_3), and the temperature of the reaction. The highest yields of lactic acid were obtained at a catalyst to sugar ratio of 2.0 (10 g Sn-Beta/5 g glucose in 300 mL water), optimizing the concentration of Lewis Acid sites for retro-aldol condensation. Similarly, using a base beta-zeolite with a high SiO_2/Al_2O_3 ratio (300 compared to 25) diluted the presence of Brønsted Acid sites, slowing down the rate of dehydration reactions which also improved reaction selectivity towards lactic acid.

To further increase overall reaction selectivity, reaction temperature was varied to maximize the rate of formation of lactic acid over side-products. At a run time of 20 hours, it was found that both lactic acid yield and selectivity from glucose were maximized at 200° C [39]. Decreasing the temperature of the reaction by decrements of 10˚C cut lactic acid yields in half while yields of levulinic acid, the major side reaction product, remained relatively constant. When the reaction temperature exceeded 200˚C, lactic acid yield decreased likely through an increased rate of decomposition reactions. As seen in Table 1, the maximum lactic acid yield from the preliminary experiments was 7.6 wt% of the initial carbon from glucose, significantly less than Sn-Beta produced through similar post-synthesis methods [52, 53]. Conversely, the yield of levulinic acid exceeds that found by Xia. This suggests that the impregnation synthesis method is less effective at removing alumina from the zeolite framework, creating more sites for dehydration reactions.

Table 1. Preliminary yield of major acid products formed from glucose using impregnated Sn-Beta compared to post-synthesis Sn-Beta [NR = Not Reported][52, 53]

Product	*Preliminary*	*Kong 2018*	*Xia 2018*
Lactic Acid	7.6%	22%	22.4%
Levulinic Acid	19.0%	NR	14.4%

Preliminary trials were conducted at a reaction time of 20 hours to ensure complete conversion of the initial monosaccharide. This high residence time, coupled with a dilute inlet sugars solution (1.6 wt%), would require significant reactor volume on an industrial scale. Additionally, comparable studies utilized significantly shorter reaction times of 2-4 hours. Further optimization of the reaction time was thus indicated to allow for a better assessment of the reaction feasibility and comparison to similar catalyst synthesis methods.

Initial time optimization trials looked at glucose in one-hour reaction increments. As shown in Figure 6, after just one hour lactic acid experienced a yield of about 11 wt% of the inlet carbon, compared to 18% for levulinic acid. Increasing the reaction time to 3 hours resulted in a slight increase in lactic acid yield with little impact on levulinic yield. From hours 3-5, a lactic acid decrease is coupled with an increase in levulinic acid yield. This change in yield is most likely associated with modest noise within the data, which is also depicted by larger sample deviation at early reaction times. Noise can be attributed to a different catalyst batch used for each time set (Hours 1-5), where slight variations in the degree of dealumination and tin addition to the zeolite can slightly alter the reactivity. Accounting for this noise and high initial acid yields, it can be concluded that minor differences in catalyst preparation have a larger effect on product yield than reaction time (>1 hour). Also of note is that shorter reaction times (<5 hours) increased the yield of lactic acid to almost 13 wt% of inlet carbon, indicating that extended time at 200° C may lead to thermal degradation.

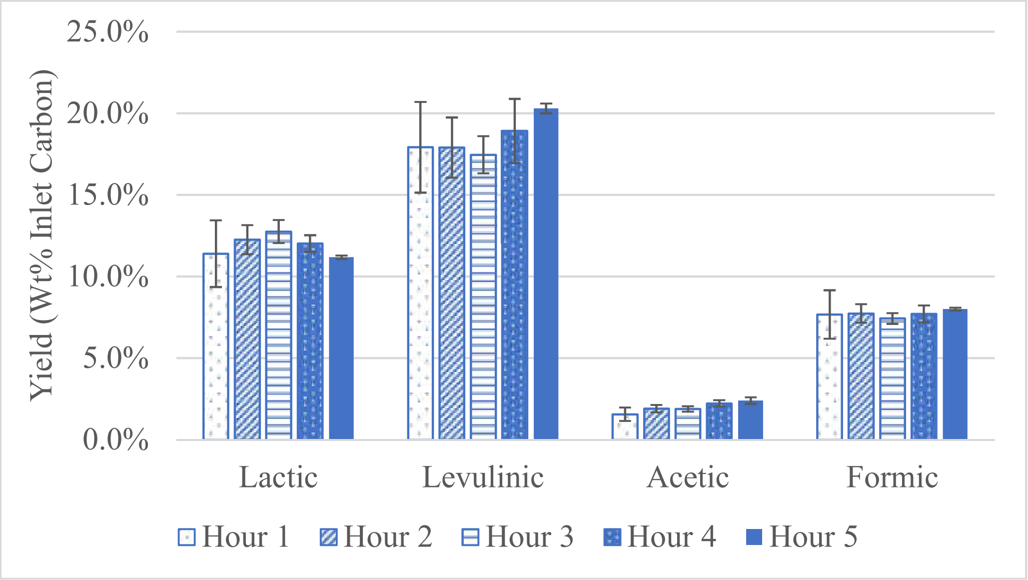

Figure 6. Effect of time on organic acid yields from Sn-Beta reaction of glucose (200˚ C)

3.2 Reaction Yields of Xylose

With the goal of utilizing carbohydrates from lignocellulosic biomass, the Sn-Beta reaction of xylose was investigated. Figure 7 illustrates how the acid yields for xylose vary as a function of reaction temperature in the range of 100-200˚ C. Lactic acid selectivity for xylose was maximized at 180˚ C compared to a selectivity maximum of 200º C for glucose. The drop in yield at temperatures above 180˚ C is likely due to an increased rate of xylose degradation reactions, which begin to out-pace retro-aldol condensation reactions for lactic acid formation. At temperatures below 120° C, high levels of unreacted sugars remain, which can be attributed to the low rate of the retro-aldol condensation reaction. It is also of note that Lewis acid sites on Sn-Beta isomerize xylose. At reaction conditions of 100° C and 20 hr, xylose is most favorably isomerized to lyxose [44]. Like xylose, lyxose is an aldopentose forming the same lactic acid precursor, glyceraldehyde. The ketose in the isomerization trio, xylulose, was not observed at 20-hour reaction times, while it was previously observed at 20% after xylose isomerization [44]. This would indicate that xylulose more readily undergoes retro-aldol condensation than its aldose isomers.

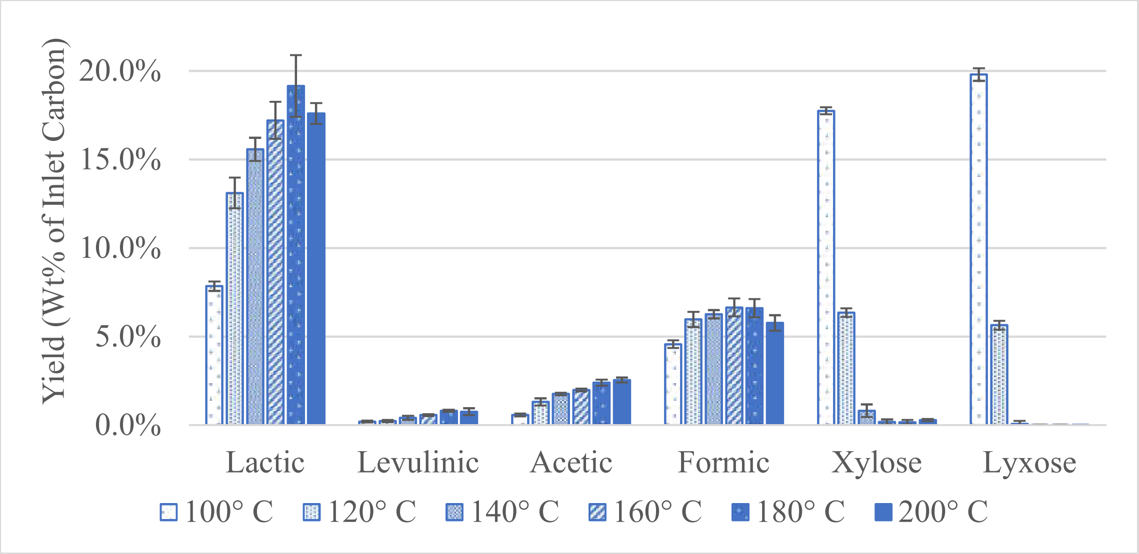

Figure 7. Temperature's effect on organic acid yields from Sn-Beta reaction of xylose (20 hr batch reaction time basis).

Theoretically, the lactic acid yield from xylose should not exceed glucose as pentoses are stoichiometrically limited to one lactic acid per molecule. This should result in half the lactic acid yield for xylose compared to glucose. Previous studies in methanol by Holm indicated a general increased yield for hexoses compared to pentoses, though not at the stoichiometric carbon ratio [44]. Conversely, the study presented herein suggests that aqueous reactions of pentoses have more favorable lactic acid yields. This could be a result of less effective glucose-fructose isomerization in water compared to methanol. However, when compared to glucose, aqueous fructose reactions with Sn-Beta had an 3% increase in yield of lactic acid. This is on par with results in methanol from Holm [44], suggesting that isomerization in water is comparable to methanol. Another potential reason is that pentose dehydration to furfural (Figure 3) occurs at a slower rate than that of the hexose dehydration to 5-HMF (Figure 4). This could result from the low solubility of furfural in water and the tendency of 5-HMF to rehydrate to levulinic and formic acids in acidic conditions, continually shifting the equilibrium point and increasing the hexose dehydration.

Industrial utilization of lignocellulosic biomass will likely require the conversion of a mixture of glucose and xylose. The lack of continuity between the optimum reaction temperatures of the two sugars needs to be reconciled. Examining corn stover specifically, a model solution of sugar extract (65 wt% glucose, 35 wt% xylose) was reacted at each optimum temperature (180°, 200° C) and the midpoint (190° C). As seen on Figure 8, the highest lactic acid yield occurs at 200° C, indicating that the prevalence of glucose outweighs the higher yields from xylose.

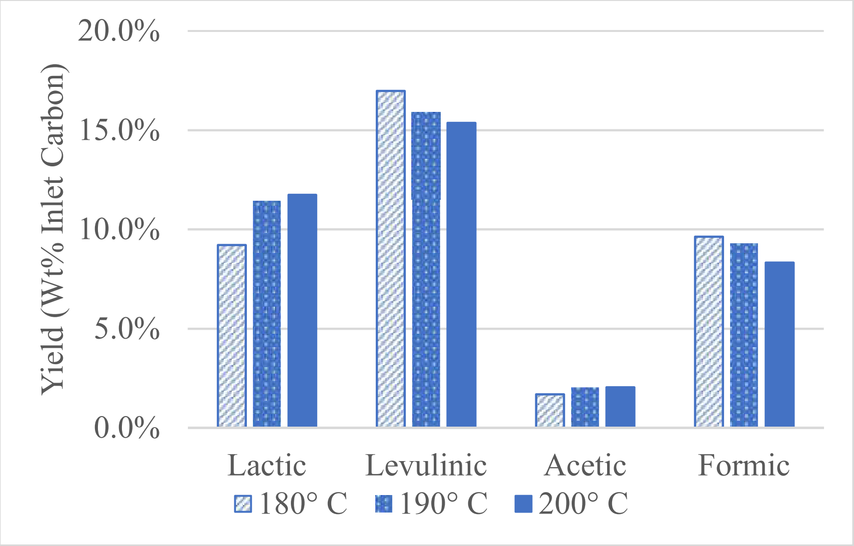

Figure 8. Temperature effect on the conversion of a corn stover model solution: 65 wt% glucose, 35 wt% xylose (5 hr reaction time basis).

With the lactic acid yield of a model corn stover solution optimized at 200°C, a time study of the xylose reaction was conducted. As seen in Figure 9, after only one hour a lactic acid yield of 18.2 wt% of initial carbon was reached. Increasing the reaction time by an hour yielded a minimal 0.4% increase by carbon weight in lactic acid production. Trials beyond the second hour experienced large variability and no discernable change in the lactic acid yield. These results agree with previous reaction time studies of glucose and the microalgae extract model solutions where variance in the catalyst batch had a larger effect than the time of reaction.

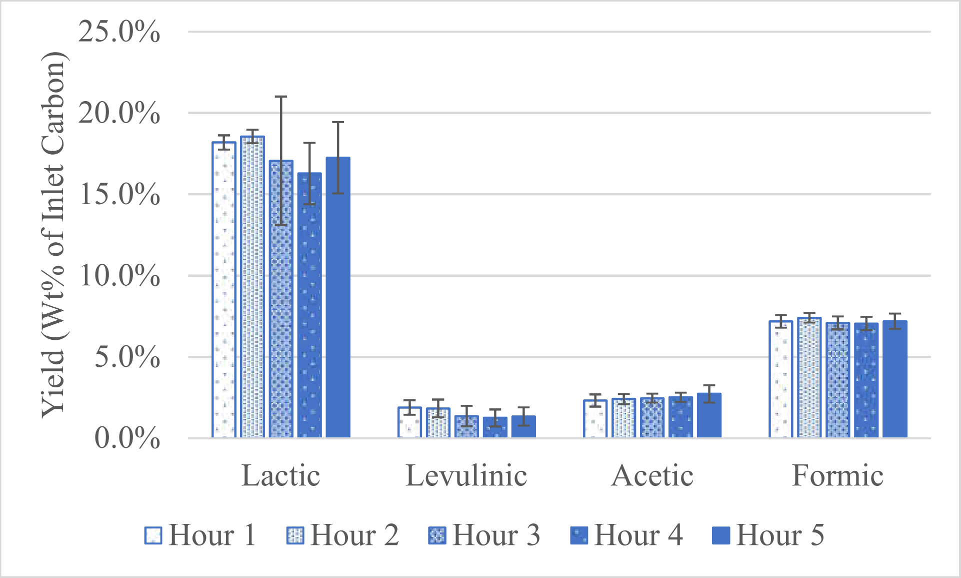

Figure 9. Effect of time on organic acid yields from Sn-Beta xylose reaction (200˚ C).

3.3 Biomass Application

A goal of this research was to optimize the synthesis of lactic acid from various biomass sources. Two such sources are microalgae, specifically the strain *Chlorella vulgaris,* and corn stover. *Chlorella vulgaris* is of interest as it has a high content of lipids, which can be transformed into biodiesel or other fuels/chemicals. Previous studies have concluded that the utilization of lipids alone is not currently economical. However, recovery and conversion of carbohydrates, which comprise about 40 wt% of the

Chlorella vulgaris biomass [57], to saleable products could be used to supplement lipid production and increase commercial viability.

Carbohydrates from *Chlorella vulgaris* can be hydrolyzed through treatment with dilute sulfuric acid to primarily two monosaccharides: glucose and galactose. To ensure that the Sn-Beta catalyzed reaction was run at optimum conditions, time trials were conducted using a sugar solution representative of the extract. The composition of this solution can be seen in Table 2 and is based on experimental results found by Hammann [58]. Figure 10 shows the effect of time on acid yield from the model solution. The longer reaction times (5-9 hours) tested for the model microalgae solutions gave similar results to the shorter times (1-5 hours) applied to pure glucose, where small changes in product yields over time were more likely a result of variability in the catalyst batch.

Table 2. Sugar composition of Chlorella vulgaris model solutions.

Sugar	*Mass %*
Glucose	62 %
Galactose	27 %
Arabinose	6 %
Fructose	3 %
Mannose	2 %

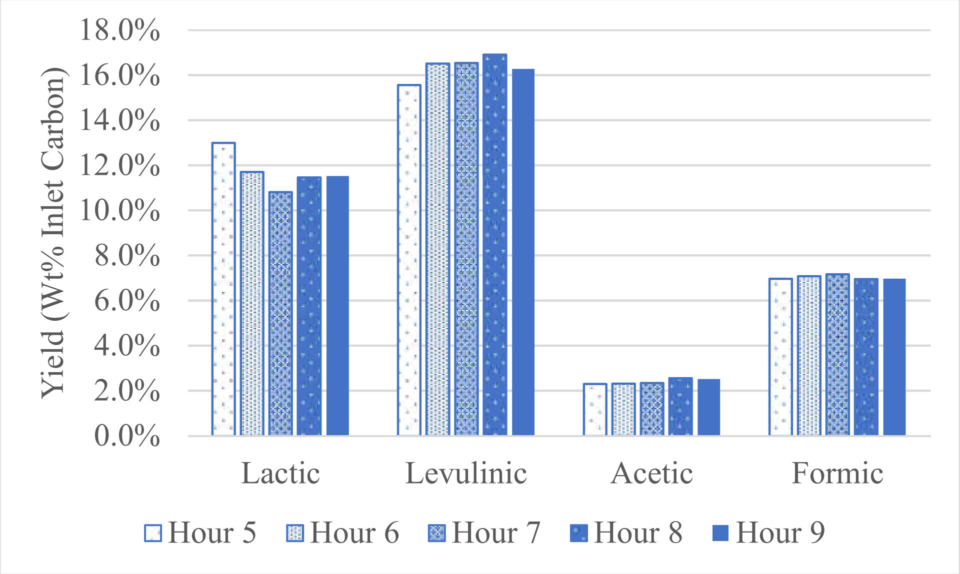

Figure 10. Effect of time on acid yield for Chlorella model solutions (200° C)

Sugar solutions extracted from an actual *Chlorella* colony, seen in Table 3, contained a larger concentration of glucose than previously expected. When reacted with Sn-Beta, *Chlorella sugar* extract yielded 5% less lactic acid by carbon weight than expected by the model and essentially none of the dehydration products, levulinic and formic acid, as shown in Table 4. Instead, two unknown products appeared on the chromatograph (Figure 11), which account for an estimated 35% of the inlet carbon weight. It is likely that these peaks were the result of a chemical component being co-extracted with the sugars from the algae. This component either catalyzed an alternative reaction pathway for the sugars or degraded into the unknown products.

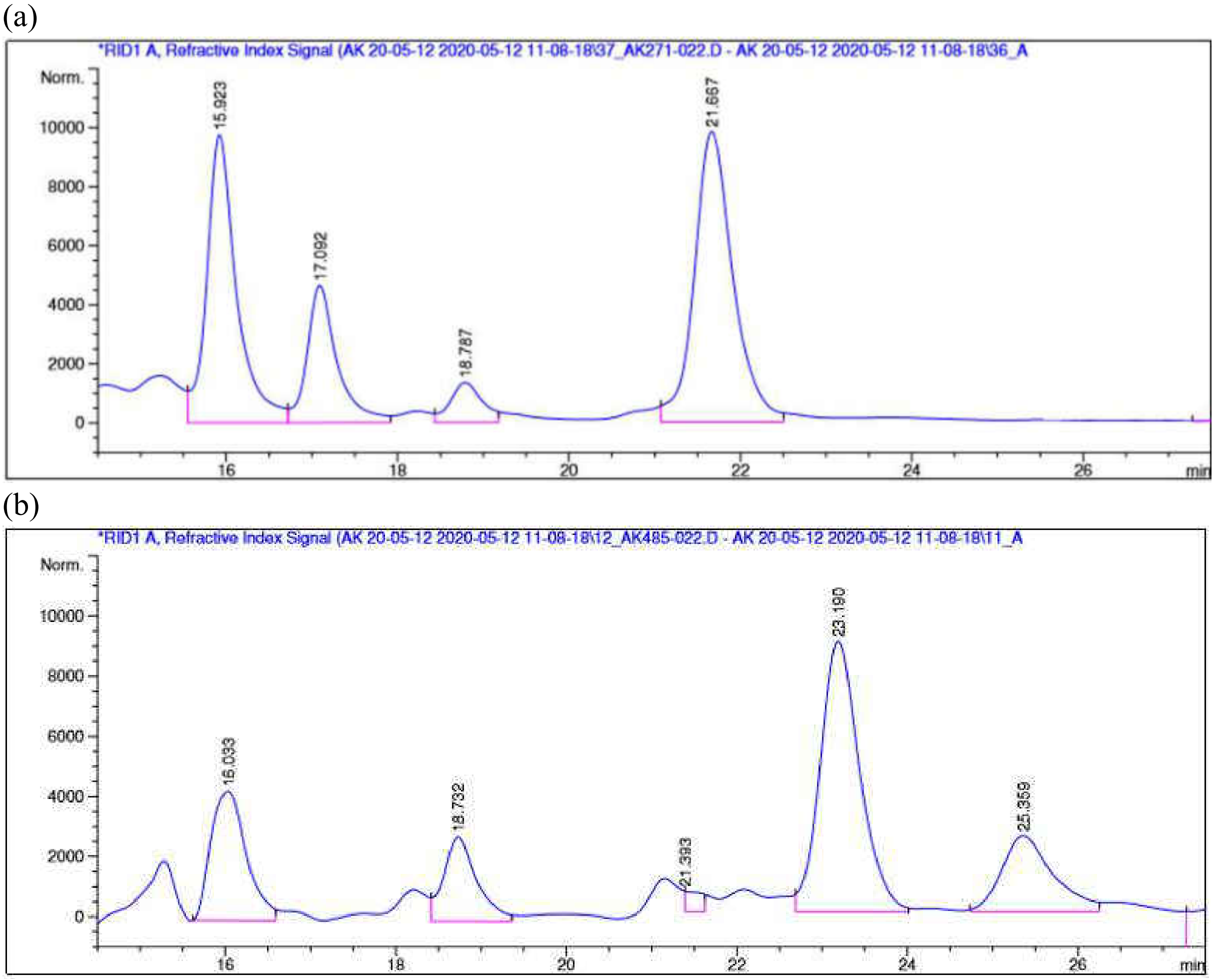

Figure 11. HPLC chromatographs of model (a) and actual (b) algae extract sugar solutions after Sn-Beta reaction. Elution order of products are as follows: lactic acid (15.9 minutes), formic acid (17.1), acetic acid (18.7), levulinic acid (21.7), and unidentified products (23.2 and 25.4). Separation occurred with an Agilent Hi-Plex H column with 5 mM H_2SO_4 mobile phase at 0.6 mL/min and 40° C.

Table 3. Sugar composition of actual Chlorella vulgaris extract.

Sugar	*Mass %*
Glucose	73 %
Galactose	26 %
Arabinose	1 %

It was hypothesized that this change in yield is related to an increased pH of 4.9 in the extract (compared to 2.8 in the model solutions), due to the neutralization of sulfuric acid (H_2SO_4) required for acid hydrolysis. To test this, a model sugar solution was placed in 4 wt% sulfuric and neutralized with calcium carbonate ($CaCO_3$), forming calcium sulfate ($CaSO_4$). Doing so nearly eliminated the production of levulinic and formic acids, emulating the observed result from the actual algae solution. In turn, the reduced competition of dehydration products resulted in a 500% increase in lactic acid yield to 62.5 wt% of the inlet carbon (Table 4). This sharp increase in lactic acid yield is not consistent with the results obtained from the actual algae sugar extract, indicating the presence of a catalytic inhibitor from the algae extract suppressing the improved lactic acid selectivity that occurs when $CaSO_4$ acts as a catalytic facilitator of the retro-aldol reaction. It is also possible that lactic acid reacts with an extraction product into one of the unknown products. These hypotheses were tested and are discussed in section 3.4.

Table 4. Comparison of yields between model sugar solutions and actual sugar solutions extracted from Chlorella vulgaris. [Model Algae Solution = expected extract sugars, Actual Algae Solution = *Chlorella vulgaris* extract, Neutralization Model = expected sugars in H_2SO_4 with $CaCO_3$ neutralization]

Product	*Model Algae Solution*	*Actual Algae Solution*	*Neutralization Model*
pH	2.8	4.9	5.7
Lactic Acid	13.0%	8.2%	62.5%
Levulinic Acid	15.6%	0.8%	2.7%
Acetic Acid	2.3%	4.4%	3.4%
Formic Acid	7.0%	0.4%	3.1%

Extraction of sugars from corn stover required a two-step process. Like sugar extraction from microalgae, corn stover was initially treated with sulfuric acid to depolymerize the hemicellulose, primarily releasing xylose. Similarly, sulfuric acid used for the hydrolysis was neutralized to calcium sulfate using calcium carbonate. The remaining carbohydrates, in the form of cellulose, are hydrolyzed using cellulase enzymes. Laboratory extraction experiments yielded a sugar extract solution of 49 wt% glucose and 51 wt% xylose. When reacted with Sn-Beta at 200° C, a lactic acid yield of 60% was observed. This differed significantly from the model solution (65% glucose/35% xylose) examined in section 3.2 where a lactic acid yield of 12% was observed. Additionally, the production of levulinic acid was almost entirely suppressed. These yields most closely match the microalgae neutralization model, indicating that it is related to the post-acid hydrolysis neutralization. Further, high lactic acid yields directly from the biomass extract reveals that corn stover hydrolysis does not release catalytic inhibitors as is apparently the case for microalgae.

Table 5. Comparison of yields between model sugar solutions and actual sugar solutions extracted from corn stover. [Model Corn Stover Solution = expected extract sugars, Actual Corn Stover Solution = Corn Stover extract]

Product	*Model Corn Stover Solution*	*Actual Corn Stover Solution*
pH	2.7	4.5
Lactic Acid	12%	60%
Levulinic Acid	15%	1.6%
Acetic Acid	2.1%	8.7%
Formic Acid	8.3%	6.7%

3.4 Effect of Acid Hydrolysis Neutralization

Conversion of microalgae and corn stover derived sugars revealed that the presence of a salt like calcium sulfate ($CaSO_4$) can lead to a significant increase in lactic acid yield with a concomitant decrease in dehydration products (i.e. levulinic and formic acid). This finding is consistent with previous results reported by Tolberg, where the combination of K_2CO_3 with Sn-Beta resulted in a lactic acid yield of 75% by carbon weight in methanol [9].

One possible explanation of this phenomena is that the salt is catalyzing the retro-aldol reaction. To test this hypothesis, a series of experiments were performed with various combinations of $CaSO_4$ and catalyst. The results are shown in Table 5. Solely using calcium sulfate for the catalysis of glucose yielded only a marginal increase in lactic acid yield (2.8 wt% of inlet carbon) compared to no catalyst at all. Similarly, a mixture of calcium sulfate and beta-zeolite resulted in a 2.9% lactic acid yield. This indicates that an interaction between $CaSO_4$ and the Brønsted Acid active sites on the zeolite are not catalyzing the retro-aldol reaction ipso facto. Further, it reconfirms that lactic acid formation is improved by the presence of tin as the Lewis Acid.

Table 6. Organic acid yields from CaSO4 catalyzed reaction of glucose (200°C, 5 hr batch reaction time basis.)

Catalyst	*Salt*	*Lactic Acid*	*Levulinic Acid*	*Acetic Acid*	*Formic Acid*
--	--	0.5%	8.4%	1.2%	3.6%
--	$CaSO_4$	2.8%	2.1%	4.4%	2.7%
Beta	$CaSO_4$	2.9%	3.5%	2.9%	2.5%
Sn-Beta	--	11%	20%	2.4%	8.0%
Sn-Beta	$CaSO_4$	68%	2.0%	3.6%	3.2%

With the elimination of individual retro-aldol catalysis through calcium sulfate as an explanation, two possible reaction mechanisms persist. The first is that the disassociated calcium ion acts in tandem with tin as a second Lewis Acid. It is known that tin is highly efficient in the final isomerization of trioses into lactic acid, but not as productive in the glucose-fructose isomerization and retro-aldol condensation steps [51, 54]. It is possible that the Lewis acidity of calcium can more efficiently catalyze those two steps, resulting in an increased yield of lactic acid. The second possible mechanism results from the mild Brønsted basicity of sulfate. This basicity neutralizes the Brønsted acid sites that result from the presence of aluminum in the catalyst framework. Dehydration products are then inhibited, increasing the reaction selectivity towards lactic acid.

To test these mechanisms two separate trials were conducted with $CaCl_2$ and NaOH to isolate the Lewis acidity and Brønsted basicity, respectively. Addition of calcium nearly doubled the yield of lactic acid from 11% to 21% by inlet carbon weight. This increase coincided with a slight decrease in the yield of dehydration products, which is likely due to the increased rate of formation of lactic acid. Neutralization of the Brønsted acid sites with NaOH resulted in a more significant increase in lactic acid yield to 35%, due to the decreased competition with dehydration product formation. Despite these increases, neither mechanism reached the efficiency of calcium sulfate, indicating that the two might work in tandem. Testing this hypothesis, calcium chloride and sodium hydroxide were both added to the reaction mixture, resulting in a lactic acid yield of 56%. This result indicates that the addition of calcium sulfate modifies Sn-Beta in situ, blocking the Brønsted acid sites with sulfate and enhancing the Lewis acid sites with calcium.

Based on the described mechanism, similar salts were examined with slightly differing properties. Increasing the Brønsted basicity by replacing sulfate with carbonate

(CO_3^{2-}) decreased the yield of lactic acid from 68 wt% to 58 wt% on inlet carbon. This resulting yield more closely resembled that from the combination of calcium and hydroxide, indicating that a weaker base more efficiently blocks Brønsted acid sites on the catalyst. Similarly, directly replacing calcium with magnesium as the Lewis acid decreases the lactic acid yield from 68 wt% to 64 wt% inlet carbon. A slightly lower yield from magnesium indicates that it is not as complimentary to tin as calcium.

Table 7. Evaluation of Lewis acid and Brønsted base effects on Sn-Beta catalyzed reaction of glucose (200°C, 5 hr batch reaction time basis)

Catalyst	*Salt*	*Lactic Acid*	*Levulinic Acid*	*Acetic Acid*	*Formic Acid*
Sn-Beta	--	11%	20%	2.4%	8.0%
Sn-Beta	$CaCl_2$	21%	16%	2.4%	6.4%
Sn-Beta	NaOH	35%	7.9%	2.4%	4.7%
Sn-Beta	$CaCl_2$ + NaOH	56%	3.4%	2.9%	0.9%
Sn-Beta	$CaCO_3$	58%	2.6%	5.1%	2.2%
Sn-Beta	$MgSO_4$	64%	3.5%	3.6%	3.1%
Sn-Beta	$CaSO_4$	68%	2.0%	3.6%	3.2%

Each salt was present at 3 g/L, which is greater than the solubility of calcium carbonate (0.013g/L) and calcium sulfate (2.1 g/L) in water. This left the possibility that the salts were catalyzing the reaction heterogeneously. Obtaining a similar yield from highly water soluble magnesium sulfate (350 g/L) eliminates this possibility, indicating that the salts alter the catalyst in situ to favor lactic acid production.

The combined Sn-Beta and calcium sulfate reaction system was then applied to different sugars, as seen in Table 8. Fructose, glucose, and mannose constitute ketose and aldose epimers capable of transformation to each other through aldo-keto isomerization. With only Sn-Beta as the catalyst, fructose had 120% of the lactic acid yield of glucose since isomerization was not necessary for fructose to undergo retro-aldol condensation. With the addition of calcium sulfate both aldose epimers (glucose and mannose) matched the lactic acid yield of their respective ketose (fructose). This increased favorability of the aldoses indicates that the addition of calcium increases the selectivity of the initial isomerization step.

Galactose, an aldohexose belonging to a separate ketose and aldose epimer trio, exhibits a lower lactic acid yield at 55 wt% of inlet carbon than glucose. This is likely due to a steric effect based on the flipped orientation of the hydroxyl group on its 4th carbon compared to glucose. It is possible this steric effect would increase the difficulty of breaking the bond between C_3 and C_4, decreasing the rate of retro-aldol condensation.

Xylose had the lowest lactic acid yield of all tested monosaccharides with Sn-Beta and $CaSO_4$ at 50 carbon wt%. This is a result of xylose being a pentose, stoichiometrically limiting its yield to 60%. Actual corn stover extract, analyzed in section 3.3, experienced a lactic acid yield of 60 carbon wt% from a 49 wt% glucose and 51 wt% xylose inlet solution. Based on Sn-Beta and $CaSO_4$ conditions, a solution with these inlet compositions is expected to have a lactic acid yield of 59.6%, almost exactly the observed corn stover yield. This further indicates that the corn stover hydrolysis does not add any catalytic inhibitors to the reaction.

Table 8. Reaction yields of various sugars using Sn-Beta and CaSO4 (200°C, 5 hr batch reaction basis.)

Sugar	*Lactic Acid*	*Levulinic Acid*	*Acetic Acid*	*Formic Acid*
Fructose	66%	2.2%	4.0%	2.7%
Galactose	55%	1.6%	3.6%	3.0%
Glucose	68%	2.0%	3.6%	3.2%
Mannose	65%	2.3%	3.9%	2.5%
Xylose	50%	2.5%	5.2%	5.4%

4. Conclusions

Sn-Beta formed through impregnation produced 9% lower lactic acid yields on a mass fraction of inlet carbon basis than those formed through traditional post-synthesis methods. This discrepancy is likely due to the presence of Brønsted Acid sites from alumina in the impregnated Sn-Beta. Maximum lactic acid yields from glucose and xylose were produced at 200° C and 180° C, respectively. Using only Sn-Beta as the catalyst, xylose unexpectedly produced higher yields of lactic acid compared to glucose. This is likely due to less selectivity to the pentose dehydration products in water. Yields of both sugars appear to reach a maximum after around one hour of reaction time with fluctuations in yields in the subsequent hours likely due to variance in Sn-Beta synthesis.

Yields of lactic and levulinic acids from Sn-Beta facilitated reaction of sugars extracted from the microalgae strain *Chlorella vulgaris* were significantly lower than model solutions. Conversely, sugars extracted from corn stover had significantly higher lactic acid yield with much lower levulinic acid yields. These results are most likely due to the combination of calcium sulfate inhibiting the production of levulinic acid and an unknown algae extract inhibiting the formation of lactic acid. Further investigation into the inclusion of calcium sulfate in the reaction revealed that sulfate modifies the surface of the catalyst in situ, blocking the Brønsted Acid sites. Additionally, calcium provided a second Lewis Acid improving the efficiency of the initial sugar isomerization step.

Applying this catalytic system to monosaccharides that are closely related to glucose (fructose and mannose) yielded nearly identical lactic acid yields around 66 wt%. Conversely, galactose, a stereoisomer that cannot be transformed to glucose through ketose-aldose isomerization, yielded only 55% lactic acid. This was likely due to steric effects related to the positioning of the hydroxyl group on its 4th carbon. Similarly, lactic acid yield from five-carbon xylose only reached 50% due to stoichiometric limitations.

Chapter III - Reactive Extraction Optimization for Aqueous-Phase Carboxylic Acids Produced from Biomass

1. Introduction

Lactic acid is a widely used chemical with applications in food, cosmetic, and pharmaceutical industries [12]. More recent applications have focused on the production of sustainable products like biodegradable plastics, such as polylactic acid (PLA), or use as a precursor to acrylic acid [21]. To meet these demands, recent work has been conducted to determine if lactic acid can be efficiently produced through chemical synthesis as opposed to traditional fermentation [8-10]. One such pathway is to utilize the Lewis acidity of a Sn-Beta zeolite catalyst to facilitate a retro-aldol condensation of monosaccharides to produce lactic acid. The reactor effluent is not pure as Brønsted acid sites on Sn-Beta catalyze dehydration reactions producing acetic, formic, and levulinic acids as side products. This diversity of short-chain carboxylic acids makes efficient separation steps a requirement for commercial scale applications.

Water or methanol are typically used as reaction solvents and participate as reactants in the final step of product formation. Typical reaction solutions have dilute levels of sugar (0.09-0.13 M) content, resulting in low product concentrations (0.13-0.18 M lactic acid) [44]. Sn-Beta reactions in methanol have typically produced higher product yields than water, likely due to increased degradation of sugars at hydrothermal conditions [9, 47]. However, the use of methanol as a solvent results in various processing hurdles. The first is that the sugars required for the reaction feed are typically obtained by hydrolyzing biomass, which occurs in aqueous solutions [59, 61]. This would then require costly steps to crystallize and then transfer the sugars to methanol. Additionally, the use of a methanol as a solvent would result in the methylation of lactic acid to methyl lactate. This product would require alternative uses or additional processing to convert back into lactic acid. For these reasons, water solvents are believed to be more scalable for industrial applications.

In conventional fermentation processes, lactic acid is purified by precipitation where calcium carbonate or calcium hydroxide are added to the fermentation broth to form calcium lactate and maintain a pH between 5 and 6 [62]. Sulfuric acid is then used to transform the resulting calcium lactate back to lactic acid, producing gypsum as a by-product. This leads to low lactic acid purities (22-44%), high sulfuric acid consumption and unwanted by-product production. More importantly, these methods are not effective in purifying lactic acid generated chemically, as the reaction product mixture is very different than the fermentation broth produced biologically.

After sugars are chemically converted into lactic acid and by-product acids in an aqueous solution, the next step in recovery from the dilute reaction product solution is to reduce the water content of the solution from the primary constituent to a minor constituent. This can be accomplished by pervaporation techniques (membrane separation), thermal evaporation, or by extracting the acids into a non-polar solvent.

Recent advancements in membrane technology has made this separation technique more competitive in the recent years [63, 64]. The one potential membrane system would utilize electrodialysis to selectively remove water from the reaction product mixture, resulting in a more concentrated lactic acid solution. Electrodialysis applies a charge to either side of the membrane repelling or attracting charged species. An anion or cation exchange membrane can then be employed to allow for selective permeation of these analytes [64]. Despite high separation efficiency, fouling and the high cost of the membranes make industrial applications difficult to justify [62, 63]. Similarly, the high energy costs for traditional evaporation reduces the attractiveness of this technique.

The most common and efficient solvents used to extract carboxylic acids out of aqueous phases are long-chain aliphatic amines. In these systems the basicity of the amine is much greater than the strength of the acid [65]. This initiates the ion-pair mechanism where the acidic proton in the carboxylic acid is transferred to the nitrogen in the amine, as seen in Figure 12 [66]. Tertiary amines are preferred for this extraction as they are cheaper and more efficient than primary and secondary amines. Additionally, when heated, primary and secondary amines form amides in carboxylic acid solutions, limiting the ability for the solvent to be regenerated [67]. The most studied tertiary amines for these applications are trioctylamine, tridodecyl amine, and Alamine 336 which is a commercially available mixture of C_8-C_{10} chain amines [68]. A novel amine extractant, N,N-didodecylpyridin-4-amine, has shown increased extraction efficiency of lactic acid compared to these other amines due to the presence of a second nitrogen, increasing complex stability through hydrogen bonding. [69]

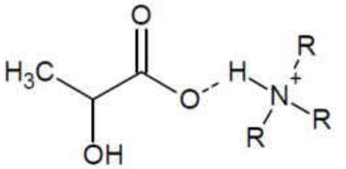

Figure 12. Ion-pair mechanism for lactic acid extraction from amines [66]

Alternatives to amines are phosphorous-based extractants including trioctylphosphine oxide (TOPO) and tributylphosphate (TBP) [70]. For these extractants, extraction equilibrium is governed by the hydropobicity of the carboxylic acids, resulting in an acid-extractant complex formed through hydrogen bonding [71]. This mechanism differs from amines where the extraction of carboxylic acids increases with decreasing pK_a [72]. However, the short-chain carboxlyic acids studied here, like lactic acid, have very low hydropobicity resulting in higher extraction efficiencies from amine extractants than from phosphorous-based extractants.

A diluent is often combined with the amine extractant to improve the physical properties of the solvent solution such as viscosity and surface tension [73]. The diluent can also increase the extraction efficiency by affecting the basicity of the amine and stabilizing the ion-pair [74]. Diluents can act as either dissociating or non-dissociating and as active or inactive. Dissociating diluents utilize high relative permittivity to dissociate the acid, resulting in reduced interactions from oppositely charged ions [70]. Diluents considered to be active utilize hydrogen bonding to participate within the complex formation. When inactive diluents are used, amines can become overloaded by forming complexes with multiple acids (see Figure 13), making it difficult for the complex to be dissolved in the organic phase, often forming a third phase [75]. A diluent with a strong solvation power will be able to reduce the frequency of overload, resulting in more efficient extraction. Alcohols like 1-octanol and isoamyl alcohol have the strongest solvation power followed by halogenated proton donors (chloroform, dichloromethane) and ketone (methyl isobutyl ketone) [67].

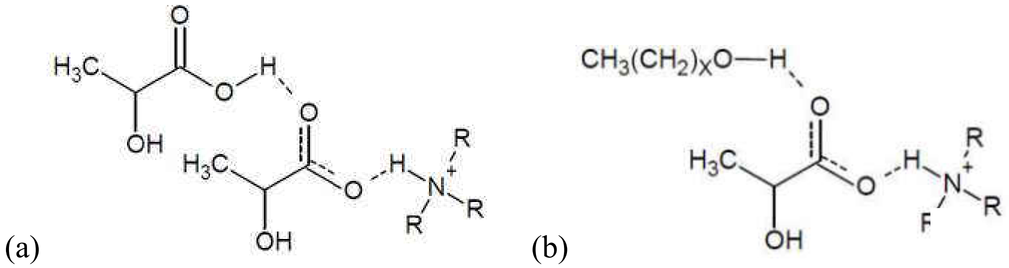

Figure 13. Mechanisms for amine reactive extraction of lactic acid: (a) overloaded amine in a (2,1)-ion-pair complex and (b) ion-pair complex with active alcohol diluent.

Though many studies have looked at the reactive extraction of lactic, levulinic, acetic, and formic acid [74, 76-78], none have developed conditions to maximize their co-extraction from the aqueous phase. The work presented here documents a study to optimize the reactive extraction of lactic, levulinic, acetic and formic acids at relevant concentrations to those formed through the Sn-Beta reaction of biomass-derived monosaccharides. Optimization focused on the most common and efficient amine extractant, trioctylamine (TOA). The goal of this work was to determine the optimum conditions for these carboxylic acids on an industrial scale as it relates to the amount of organic phase required, extractant concentration, extraction time, extraction temperature, and reaction product composition. Additionally, two active alcohol diluents, 1-octanol and isoamyl alcohol, were examined to determine their impact on the overall separation system efficiency.

2. Methods

2.1 Materials

Organic solvents required for extraction were TOA, isoamyl alcohol, and 1-octanol. Trioctylamine (98% purity) was used as the primary extractant and was obtained from EMD Millipore (Billerica, USA). Isoamyl alcohol (98% purity) and 1-octanol (99% purity) were used as active diluents and obtained from TCI America (Portland, USA) and Fisher Scientific (Waltham, USA), respectively. Carboxylic acid solutions were obtained by Sn-Beta reaction of glucose using the procedure defined in the previous chapter. Separate samples were used to match each of the desired carboxylic acid compositions.

2.2 Extraction

Each extraction experiment utilized 5.00 g of carboxylic acid solution (containing lactic, levulinic, acetic, and formic acids in water) in a test tube. Acid concentrations were altered by mixing or diluting available solutions to meet the intended concentration. Trioctylamine was then added with either isoamyl alcohol or 1-octanol as the diluent to meet the intended extractant concentration and organic-to-aqueous phase ratio. Test tubes were then placed on an Innova model 40 shaker table (New Brunswick, Edison, USA) for the intended extraction time. The table was enclosed and the temperature closely controlled. After extraction, the solution was allowed to settle and separate. If the phases took more than ten minutes to separate, the sample was centrifuged for five minutes. The organic layer was then removed using a syringe.

2.3 Analysis

Initial and final acid concentrations of the aqueous phase were quantified using High Performance Liquid Chromatography (Agilent Technologies 1200 series, Santa Clara USA). Separation occurred using an Agilent Hi-Plex H organic acid column with a mobile phase of 5 mM H_2SO_4. The HPLC operated at a mobile phase flow of 0.6 mL/min at 40° C. A refractive index detector (G1362A: Agilent Technologies, Santa Clara, USA) was used and measured each analyte for up to 75 minutes.

Extraction effectiveness was measured using the distribution coefficient (K_D) which is defined as:

$$K_D = [HA]_{org} / [HA]_{aqu}$$

> where $[HA]_{org}$ is the concentration of the target acid in the organic phase produced during the extraction and $[HA]_{aqu}$ is the target acid concentration in the aqueous phase produced during the extraction.

Organic phase concentrations were calculated by finding the mass loss of the target acid in the aqueous phase and dividing it by the organic phase volume. Experimental data was further analyzed statistically using Minitab software (NIST, v.18).

3. Results

3.1 Factor Screening

A six factor, half factorial design was created to screen the effect of possible factors on the distribution coefficients of acetic, formic, lactic, and levulinic acid. The design, seen in Table 9, has a resolution of VI so that the main effects do not confound with any two-, three-, or four-way interactions. Five of the six factors were chosen to understand the optimum conditions of an industrial scale extraction process. These included organic/aqueous ratio (to determine the optimal requirement), extractant concentration (to determine how much diluent is required to solvate the complex), extraction time, extraction temperature, and the mass ratio of lactic acid to levulinic acid ratio. The lactic acid/levulinic acid ratio was chosen to express the selectivity of the prior catalytic reaction of glucose. A high lactic acid concentration would correspond towards a more productive retro-aldol reaction, while a high levulinic acid concentration would result from dehydration reactions. Table 10 shows the concentration of the solutions representing these mixtures. The final parameter was chosen to establish the difference in performance achieve by two alcohol diluents: 1-octanol and isoamyl alcohol.

Table 9. Reactive extraction screening six-factor, half factorial design.

Run #	Run Order	Loading	Extract Conc.	Time (min)	Diluent	Lac/Lev Ratio	Temp
1	10	0.25	15	1	1-octanol	0.05	45
2	7	0.25	15	1	1-octanol	1.42	25
3	8	0.25	15	1	Isoamyl Alcohol	0.05	25
4	15	0.25	15	1	Isoamyl Alcohol	1.42	45
5	20	0.25	15	5	1-octanol	0.05	25
6	29	0.25	15	5	1-octanol	1.42	45
7	26	0.25	15	5	Isoamyl Alcohol	0.05	45
8	19	0.25	15	5	Isoamyl Alcohol	1.42	25
9	5	0.25	40	1	1-octanol	0.05	25
10	13	0.25	40	1	1-octanol	1.42	45
11	16	0.25	40	1	Isoamyl Alcohol	0.05	45
12	2	0.25	40	1	Isoamyl Alcohol	1.42	25
13	28	0.25	40	5	1-octanol	0.05	45
14	23	0.25	40	5	1-octanol	1.42	25
15	22	0.25	40	5	Isoamyl Alcohol	0.05	25
16	31	0.25	40	5	Isoamyl Alcohol	1.42	45
17	1	0.75	15	1	1-octanol	0.05	25
18	11	0.75	15	1	1-octanol	1.42	45

19	9	0.75	15	1	Isoamyl Alcohol	0.05	45
20	3	0.75	15	1	Isoamyl Alcohol	1.42	25
21	32	0.75	15	5	1-octanol	0.05	45
22	17	0.75	15	5	1-octanol	1.42	25
23	24	0.75	15	5	Isoamyl Alcohol	0.05	25
24	25	0.75	15	5	Isoamyl Alcohol	1.42	45
25	14	0.75	40	1	1-octanol	0.05	45
26	6	0.75	40	1	1-octanol	1.42	25
27	4	0.75	40	1	Isoamyl Alcohol	0.05	25
28	12	0.75	40	1	Isoamyl Alcohol	1.42	45
29	18	0.75	40	5	1-octanol	0.05	25
30	30	0.75	40	5	1-octanol	1.42	45
31	27	0.75	40	5	Isoamyl Alcohol	0.05	45
32	21	0.75	40	5	Isoamyl Alcohol	1.42	25

Table 10. Solution concentrations (g/L) used to represent high and low lactic/levulinic acid ratios in screening experiments.

	High (1.42)	Low (0.05)
Acetic Acid	0.7	0.8
Formic Acid	0.5	0.3
Lactic Acid	5.2	0.5
Levulinic Acid	2.7	5.0

Screening trials were split into 4 separate blocks depending on extraction time and temperature. Trials run in each block were then randomized to limit error. Results of these screening runs can be seen in Table 11.

Table 11. Distribution coefficient (KD) for each acid in screening trials.

Run #	Acetic	Formic	Lactic	Levulinic
1	0	3.4	0.6	0.8
2	8.2	28.9	13.4	6.2
3	4.6	8.5	1.3	4.8
4	21.8	76.0	16.2	5.6
5	2.9	9.0	0.9	2.0
6	6.7	30.2	13.9	5.2
7	7.6	30.1	3.3	6.6
8	24.3	56.8	36.8	17.6

9	4.8	8.6	1.5	3.5
10	7.7	29.2	14.7	5.8
11	9.7	21.3	2.7	8.7
12	21.9	41.4	30.5	18.1
13	5.7	16.1	2.0	4.7
14	12.3	35.3	20.6	8.9
15	13.3	27.4	2.8	9.7
16	17.4	54.0	33.2	14.2
17	6.7	9.1	1.1	5.1
18	3.9	13.0	6.0	2.9
19	5.5	20.4	1.6	10.1
20	29.0	32.6	23.1	19.0
21	2.2	10.3	0.9	3.6
22	7.0	30.7	10.3	6.2
23	15.7	14.0	1.1	10.7
24	15.5	29.5	30.2	13.6
25	2.5	12.4	0.5	2.4
26	6.5	20.1	7.7	5.3
27	11.4	26.3	2.1	9.8
28	13.0	16.9	19.2	17.5
29	7.9	8.5	0.4	4.5
30	11.2	18.7	11.3	6.8
31	6.8	14.8	2.0	7.6
32	18.8	20.7	20.5	15.2

Factor effects for each distribution coefficient were analyzed using the “Design of Experiments” feature for factorial designs in Minitab. Factors found to have a statistically significant effect can be seen in Table 12. A more detailed analysis of model reduction is provided in Appendix C.

Table 12. Factors with significant effects on the distribution coefficients of each acid (“+” indicates main effect increased KD, “-” indicates main effect decreased KD¬, “x” indicates two-way interaction was significant).

Factor	Acetic	Formic	Lactic	Levulinic
A: Org/Aqu. Ratio		-	-	
B: Extract. Conc.				+
C: Time			+	
D: Diluent	Isoamyl	Isoamyl	Isoamyl	Isoamyl
E: Lac/Lev Ratio	+	+	+	+

F: Temperature	-			-
AB				x
AC				
AD		x		
AE		x	x	
AF				
BC				
BD	x			
BE		x		
BF				
CD				
CE			x	
CF				
DE	x		x	x
DF				
EF				

The most prevalent factor for acid extraction is the type of diluent. For each acid, isoamyl alcohol resulted in a distribution coefficient nearly double that of 1-octanol. This agrees with previous results seen by Uslu, were the shorter-chained isoamyl alcohol has twice the solvation power of 1-octanol for levulinic acid extraction [68]. It is important to note that isoamyl alcohol has a solubility in water of 0.3 M, fifty times greater than that of 1-octanol [79, 80]. In the context of the overall catalytic conversion of sugars process, significant process water usage in the reaction stage requires it to be recycled. A high solubility of the extraction solvent in water could result in lower reaction efficiency or higher downstream separation costs. As seen in Table 13, when water saturated with isoamyl alcohol is used as the liquid media for the sugar conversion reaction, lactic acid yield decreases by 14 wt% of inlet carbon. This is likely due to the formation of isoamyl lactate as a side-product. The significantly lower solubility of 1-octanol in water would help to mitigate this side-product formation. Though less efficient, the lowest average distribution coefficient of 4.5 for the extraction of levulinic acid using 1-octanol is still highly efficient for an industrial scale separation. Therefore, it is still worth considering 1-octanol as the extraction diluent.

Table 13. Sn-Beta + CaSO4 reaction yields from glucose with and without an isoamyl alcohol saturated reaction solvent (wt % of Inlet Carbon).

Reaction Solvent	Lactic Acid	Levulinic Acid	Acetic Acid	Formic Acid
Water Saturated with Isoamyl	54%	6.3%	3.0%	4.2%
Pure Water	68%	2.0%	3.6%	3.2%

The second most influential factor was the ratio of lactic to levulinic acid concentrations. For each acid a higher K_D was observed when more lactic acid was present. The reason for this is that levulinic acid has a higher pKa than lactic acid (4.64 vs. 3.86) [81, 82]. Thus, when levulinic acid is in abundance, it provides less hydronium ions to the solution through disassociation than lactic acid. This in turn results in a more basic environment, forcing the other acids to deprotonate and reducing their ability to form complexes through proton transfer [71].

The other major factor was the ratio between the organic and aqueous volumes. For both formic and lactic acids, increasing the organic phase volume decreases the distribution coefficient. This would indicate that lower organic volumes sufficiently solvate the ion-pair and additional organic phase only serves to dilute the extracted product.

Similarly, temperature was significant for two acids. An increase in the temperature decreased the K_D of acetic and levulinic acid, while not having a significant effect on the other two acids. This decrease is due to exothermic nature of ion-pair formation, making extraction less favorable at higher temperatures [83]. Since this trend is consistent with previous work, temperature was not explored further.

Each of the three remaining main effects were only significant for one acid. Increasing the extractant concentration increases the K_D of levulinic acid, indicating that a 15 wt% concentration of TOA is too low for the maximum formation of levulinic acid ion-pairs. A five-minute mixing time slightly increased the K_D of lactic acid compared to one minute, while not influencing the other acids. This would indicate that mixing time is not a major factor for extraction.

3.2 Optimization

Based on the screening experiments, it was determined that extraction temperature and time could be neglected. Further experiments kept these variables constant at 25° C and 5 minutes. This left three quantitative parameters – organic/aqueous ratio, extractant concentration, and lactic acid/levulinic acid ratio - to still be optimized. Additionally, since the physical properties of isoamyl alcohol raise downstream separation concerns, there was a need to find optimum conditions to compare overall extraction efficiencies of both diluents. To optimize these parameters identical three-factor central composite designs were created for each diluent, seen in Tables 14 and 15.

Table 14. Reactive extraction three-factor central composite design and KD results for isoamyl alcohol as the diluent.

	Optimization Factors				K_D			
Run #	Diluent	Org./Aqu. Ratio	Extract. Conc.	Lac/Lev Ratio	Acetic	Formic	Lactic	Levulinic
1	Isoamyl	0.45	20	0.19	35	105	40	26
2	Isoamyl	0.3	30	0.53	29	121	49	27
3	Isoamyl	0.15	20	1.15	23	107	66	23
4	Isoamyl	0.3	30	0.53	29	94	46	26
5	Isoamyl	0.45	20	1.15	24	80	48	34
6	Isoamyl	0.3	30	0.53	27	120	46	26
7	Isoamyl	0.15	20	0.19	16	58	24	14
8	Isoamyl	0.15	40	1.15	19	113	54	18
9	Isoamyl	0.3	30	0.53	27	90	46	25
10	Isoamyl	0.45	40	1.15	20	67	37	25
11	Isoamyl	0.45	40	0.19	33	80	32	25
12	Isoamyl	0.3	13	0.53	23	140	45	21
13	Isoamyl	0.05	30	0.53	27	84	41	31
14	Isoamyl	0.15	40	0.19	31	92	38	26
15	Isoamyl	0.3	30	0.53	31	138	53	25
16	Isoamyl	0.3	47	0.53	30	81	43	25
17	Isoamyl	0.48	30	0.53	7	67	28	30
18	Isoamyl	0.3	30	0.53	41	130	54	35
19	Isoamyl	0.3	30	0.10	15	34	5	14
20	Isoamyl	0.3	30	1.46	37	114	59	11

Table 15. Reactive extraction three-factor central composite design and KD results for 1-octanol as the diluents.

	Optimization Factors				K_D			
Run #	Diluent	Org./Aqu. Ratio	Extract. Conc.	Lac/Lev Ratio	Acetic	Formic	Lactic	Levulinic
1	1-octanol	0.15	40	1.15	14	99	42	13
2	1-octanol	0.15	20	1.15	12	89	29	10
3	1-octanol	0.3	30	0.53	23	108	40	19
4	1-octanol	0.15	20	0.19	18	71	27	13
5	1-octanol	0.3	13	0.53	14	63	24	10
6	1-octanol	0.05	30	0.53	32	78	2	21
7	1-octanol	0.3	47	0.53	24	89	34	16
8	1-octanol	0.05	30	0.53	29	90	31	21
9	1-octanol	0.3	30	0.53	29	119	42	19
10	1-octanol	0.3	30	0.53	36	130	43	23
11	1-octanol	0.3	30	0.53	26	119	40	18
12	1-octanol	0.45	20	0.19	34	118	36	20
13	1-octanol	0.15	40	0.19	23	107	36	15
14	1-octanol	0.45	20	1.15	23	96	40	17
15	1-octanol	0.45	40	0.19	27	78	30	18
16	1-octanol	0.45	40	1.15	23	81	33	16
17	1-octanol	0.3	30	0.53	25	103	39	18
18	1-octanol	0.3	30	0.53	28	123	42	19
19	1-octanol	0.3	30	0.10	8	13	3	7
20	1-octanol	0.3	30	1.46	18	57	61	13

To balance the extraction efficiencies of the four target acids, Minitab's Response Optimizer function was utilized to find the optimum conditions. When considering co-extraction of these acids on an industrial scale, more weight is given to the harder to extract acids when developing optimum conditions. For amine extraction, acid-extractant complex equilibrium increases with decreasing pK_a of the carboxylic acid [72]. Based on the order of increasing pK_a, extraction efficiency is as follows: formic, lactic, acetic, and levulinic. Accounting for this, the response optimizer was given more weight to levulinic

(10 weighting factor) and acetic (5), and less weight to highly efficient formic (0.1). The results are summarized in Figure 14.

Extraction with isoamyl alcohol as a diluent was optimized at conditions of 0.38 organic/aqueous phase weight ratio, 15.6 wt% of TOA in the organic phase, and a lactic/levulinic acid ratio of 1.36. Initial 1-octanol optimization resulted in an unbounded maximum for organic/aqueous phase ratio. To bound this, two additional phase ratios beyond the high (0.70 and 0.85) were experimentally generated while keeping the other two variables at the center of their range (30 wt% TOA and 0.53 Lac/Lev ratio), seen on Table 16. Doing so resulted in a maximum organic/aqueous phase ratio of 0.63. Additional optimum conditions for extraction with 1-octanol include 25 wt% of TOA and 0.87 lactic/levulinic acid ratio.

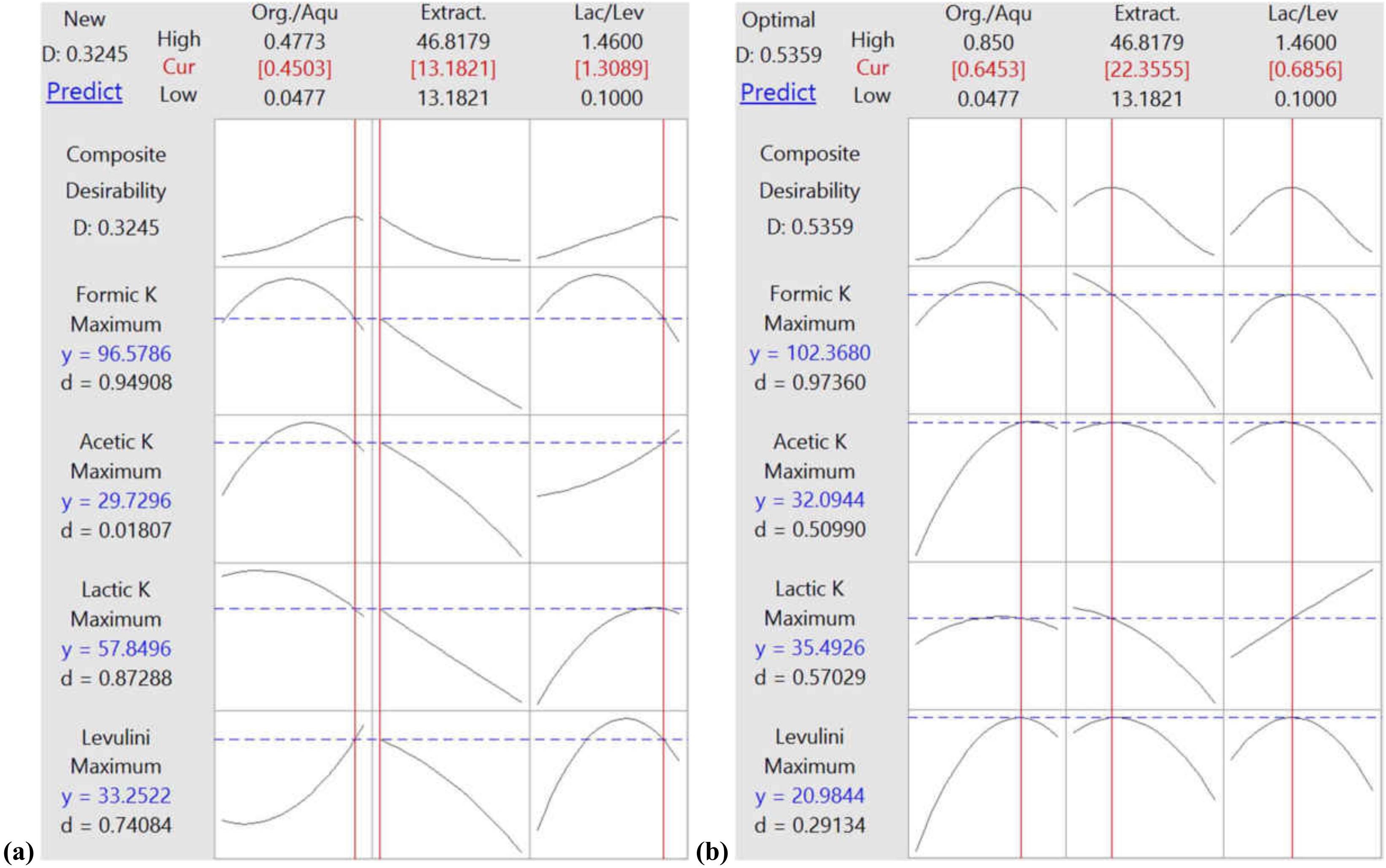

Figure 14. Optimum reactive extraction conditions using Isoamyl Alcohol (a) and 1-octanol (b) as the active diluent to remove carboxylic acids from the aqueous phases with TOA as extractant. Response optimizer results were obtained from Minitab with the following weightings: Levulinic Acid 10, Acetic Acid 5, Lactic Acid 1, and Formic Acid 0.1.

Table 16. Conditions and Partitioning Results for Additional 1-Octanol Trials Conducted to Bound the Organic/Aqueous Phase Ratio.

	Optimization Factors				K_D			
Run #	Diluent	Org./Aqu. Ratio	Extract. Conc.	Lac/Lev Ratio	Acetic	Formic	Lactic	Levulinic
21	1-octanol	0.85	30	0.53	28	47	23	18
22	1-octanol	0.7	30	0.53	29	65	25	18
23	1-octanol	0.05	30	0.53	13	76	29	10
24	1-octanol	0.7	30	0.53	31	68	27	18
25	1-octanol	0.85	30	0.53	30	59	23	18
26	1-octanol	0.55	30	0.53	28	86	30	19

Maximum conditions for extraction with each diluent differed significantly. Extraction with isoamyl alcohol is optimized with significantly less total organic phase than 1-octanol (0.45 and 0.64 organic/aqueous phase ratio, respectively). Similarly, extraction with 1-octanol requires nearly double the concentration of TOA (22 wt% compared to 13 wt%). These requirements are both related to the much higher efficiency of isoamyl alcohol to solvate the carboxylic acid-TOA ion-pair. Additionally, the optimum lactic/levulinic ratio of isoamyl alcohol extraction more closely matches the ratio generated in the most efficient yields experiment from the prior Sn-Beta reaction (35 g lactic acid/ 1 g levulinic acid).

3.3 Process Design Comparison

Though isoamyl alcohol was shown to be more efficient than 1-octanol for carboxylic acid extraction, its significantly higher affinity to water and similar volatility to formic acid are significant drawbacks for downstream separation unit operations. To compare these diluents, simple process designs were developed to determine if the additional efficiency provided by isoamyl will out weight its additional purification requirements. Economic analysis based on the capital investment and operating expenses was the main criteria.

To ensure a consistent basis between the two processes, the following assumptions were employed:

1. The mixture of carboxylic acids was based on the most efficient Sn-Beta conversion of monosaccharides, primarily glucose and xylose, derived from corn stover.
2. Based on yields found in Chapter 2, Sn-Beta converts glucose to 68% lactic, 1.6% levulinic, 7% acetic, and 3.2% formic on a wt% of inlet carbon basis. Similarly, xylose has conversions of 50% lactic, 1.6% levulinic, 9.8% acetic, and 7.9% formic.
3. Scale was determined based on using 30 wt% of the average corn stover produced to generate the amount of feed corn needed to operate a typical North Dakota ethanol plant at its design operating (nameplate capacity) capacity of 94 million gallons per year [84]. This results in designing a facility that can process of 225,000 dry tons/year (27,000 kg/hr) of feed corn stover into the four acid products, plus by-products.
4. Based on assumptions 1-3, the carboxylic acid stream composition consists of 9,400 kg/hr of lactic acid, 390 kg/hr levulinic acid, 1,200 kg/hr acetic acid, and 1,200 kg/hr of formic acid.
5. The extraction process was designed to extract 99% of acetic acid in the isoamyl process and 99% of levulinic acid in the 1-octanol process. Since these acids

have the smallest extraction coefficient for their assigned process, all subsequent acids should exceed this efficiency.

The process to separate the acids from the TOA/1-octanol extraction solution, recover the extraction solution for reuse, and purify the acids to their product grades is shown in Figure 4. To account for the composition of the inlet carboxylic acid stream, optimal 1-octanol conditions needed to be adjusted to match the acid concentrations. This was done by holding the lactic/levulinic acid ratio constant at the maximum value (1.46) using the response optimizer function in Minitab. These updated conditions can be seen in Table 17. Extraction with isoamyl alcohol is at the optimum with the maximum lactic/levulinic acid ratio, so design conditions are as previously specified.

Table 17. Updated 1-octanol extraction conditions accounting for inlet stream composition used for process design (Lactic/Levulinic Acid ratio held at 1.46).

Condition	Value
Organic/Aqueous Ratio	0.64
Extractant Concentration (wt%)	23.4%
Acetic K_D	21.4
Formic K_D	43.5
Lactic K_D	48.8
Levulinic K_D	14.7

When designing the entire separation/purification process, it is of note that TOA forms an azeotrope with both lactic and levulinic acid, eliminating simple distillation as a viable separation method. Further, both lactic and levulinic acid have volatilities below either of the alcohol diluents. Straight distillation would result in a diluent-less TOA stream that would provide transport concerns due to its high viscosity [62]. To avoid this, the use of a second tertiary amine with a higher basicity than TOA has been proposed as a back extractant. The most common is trimethylamine (TMA), which possesses shorter hydrocarbon branches making its subsequent ion-pair soluble in water [77]. Since this study did not optimize back extraction with TMA and a water diluent, distribution coefficients were estimated based on previous work which can be seen in Table 18 [85, 86].

Table 18. Distribution coefficients Associated with the Back Extraction of Carboxylic Acids using a Trimethylamine (TMA) Extractant and a Water Diluent [85-87]

Target Acid	K_D
Acetic Acid	5
Formic Acid	4
Lactic Acid	7
Levulinic Acid	10

Since both acetic and formic acids form azeotropes with water (16.6 and 38.2 mol%, respectively), it is advantageous to remove these volatile acids before the back extraction step. When 1-octanol is the diluent for extraction, this is can be accomplished by distillation followed by adsorption. At atmospheric pressure, complete separation of 1-octanol from formic acid is inhibited by an azeotrope at 100.4° C (98.5 mol% formic acid) so distillation to the azeotrope is followed by size exclusion adsorption using a 13x molecular sieve (Figure 4). The ensuing acetic-formic acid stream can then be further purified to their saleable grades through distillation.

The lactic and levulinic acid that remain in the TOA/1-octanol solution after the distillation of the volatile acids are then back extracted into TMA/water. The high relative volatility of TMA and water compared to these heavier acids makes regeneration straightforward through distillation. Both solvent systems (TOA/1-octanol and

TMA/water) are fully recovered and recycled back for further extraction. After solvent regeneration, lactic and levulinic acids can be distilled to their desired purities.

The process to separate the acids from the TOA/isoamyl extraction solution, recover the extraction solution for reuse, and purify the acids to their product grades is shown in Figure 5. For the case of using isoamyl alcohol as the diluent, separation of extracted acids from the isoamyl alcohol solution is hindered since isoamyl: 1) forms a more significant azeotrope with formic acid at atmospheric conditions (89.2 % at 98.2° C) than 1-octanol and 2) has a higher volatility than acetic acid. To ensure that TOA remains in the diluent, the latter relationship requires acetic acid to be back extracted into TMA/water, where further purification is also hindered by an azeotrope. To simplify this separation both acetic and formic acid are back extracted with TMA/water. The ensuing azeotropes between these acids and water are then broken through extractive distillation (Figure 5).

Previous work has identified nonvolatile carboxylic acids, like heptatonic and azelaic, for the removal of formic acid from water [88]. Using ChemCAD (version 7.1.4.10142) process simulation, it was found that azelaic acid fed at twice the combined molar concentrations of acetic and formic was able to remove nearly all of the entrained acids through extractive distillation. For this step to occur, TMA needs to be removed first to ensure it does not form an ion-pair with the azelaic acid, resulting in that acid being reactively extracted.

In the isoamyl alcohol process, both lactic and levulinic acid were removed from the solvents and volatile acids immediately after back extraction with TMA/water. These two acids were then distilled to produce their saleable grades.

Economic comparisons of these two processes were limited in scope to the separation steps depicted in Figures 15 and 16, excluding the formic/acetic and lactic/levulinic distillation columns. These were omitted because there are only minor differences between the two processes that would not affect the comparison. Since both proposed processes utilize the same input carboxylic acid compositions and produce the same grade of saleable products, revenues generated were assumed to be equal. This left two primary measures of economic cost: annual operating expenses and total capital investment. Total capital investment was distributed across a 36-month period prior to commissioning while operating costs were applied to the subsequent 20 years of process operating life. Since these separation steps represent only part of a grass-roots plant, federal and state taxes were applied as normal to calculate the how much less tax is due based on the operating expenses and equipment depreciation. Total expenses were then summed at a 20% hurdle rate, yielding a net present value of each process, seen on Table 19. Complete economic analysis of these two separation processes can be seen in Appendix D.

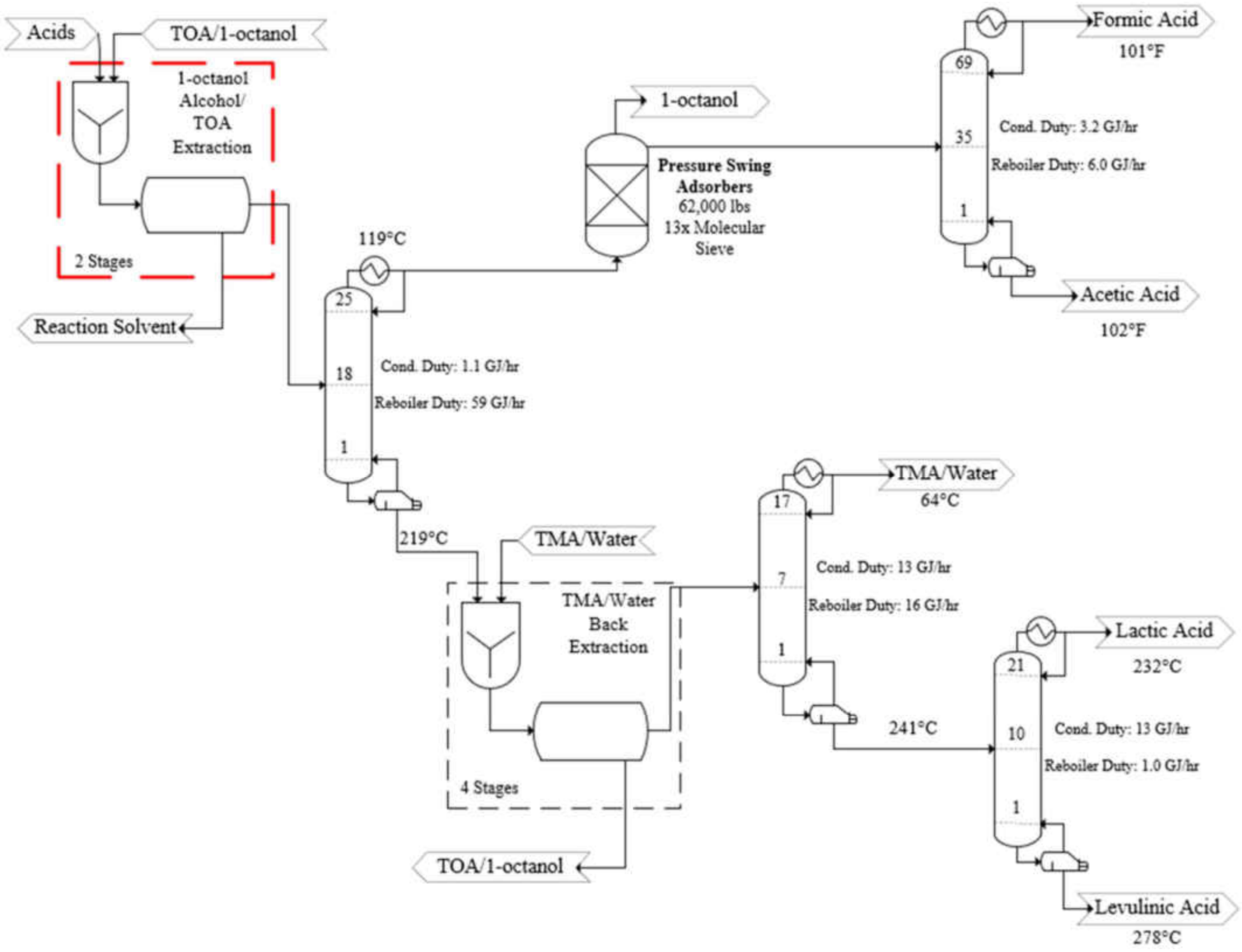

Figure 15. Remaining separation process steps for carboxylic acid extraction with 1-octanol

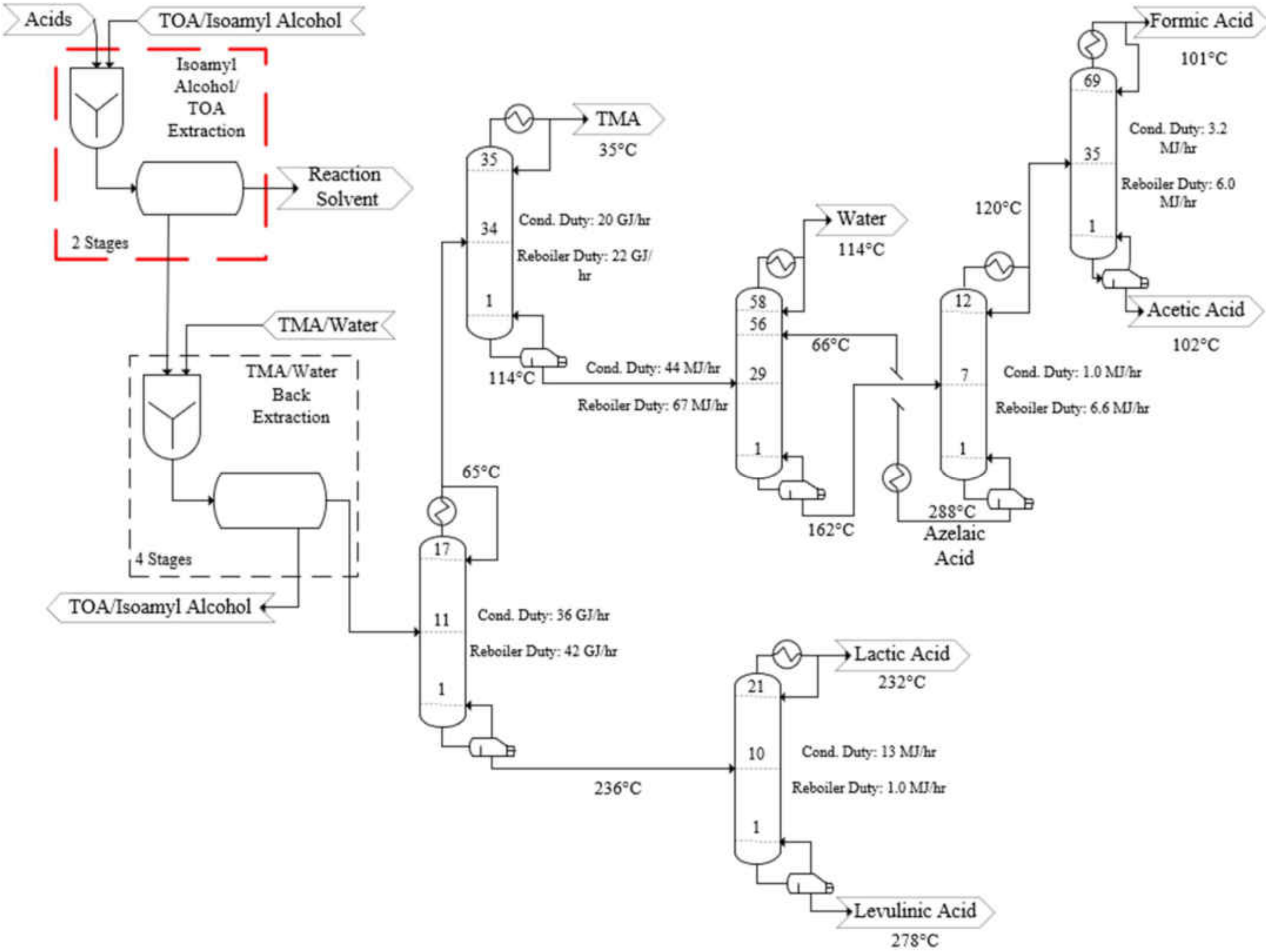

Figure 16. Remaining separation process steps for carboxylic acid extraction with isoamyl alcohol.

Table 19. Economic comparison of isoamyl and 1-octanol carboxylic acid separation processes. Estimate was conducted using a January 2019 basis and 20-year project life cycle.

Measure	1-octanol	Isoamyl
Average Annual Operating Costs	$10.1 million/yr	$12.2 million/yr
Total Capital Investment	$14.1 million	$19.9 million
Net Present Value @ 20%	**($52.5 million)**	**($66.8 million)**

The biggest source of economic variance between the two processes is the average annual operating costs. Operating expenses used in this assessment include chemicals, operating labor, maintenance, cooling water, boiler-feed water, and steam. The primary expenses for both processes were the cooling water and steam required for the distillation towers. These expenses are higher in the isoamyl alcohol process which requires 40% more heating and cooling duty than 1-octanol. The larger distillation train required to break the formic acid azeotrope in the isoamyl alcohol process also results in a slightly larger total capital investment.

When weighting these two costs across a 20-year project life cycle, the incremental NPV@20% is $14.3 million in favor of the 1-octanol process. Considering the $50 million scale of the expenses, this difference lies within the ±40% confidence interval (~$20 million) of the estimate. However, the large dependency on distillation for both processes makes it unlikely that the estimates diametrically differ from the final cost. This makes it highly likely that using 1-octanol as a diluent will be more cost effective. Further, the isoamyl alcohol process design does not take into account isoamyl alcohol's high solubility in water (~30 g/L). Another costly separation step would be required to ensure total recovery of isoamyl alcohol so that the extraction solvent could be recycled back into the process without adverse effects to the catalytic reaction.

4. Conclusions

Reactive extraction with tertiary amines provides an efficient mean of separation for a dilute mixture of carboxylic acids. Solvation of the subsequent acid-amine complex is aided by the presence of an alcohol to act as an active diluent. Screening and optimization studies were conducted to maximize the separation efficiency of acetic, formic, lactic, and levulinic acids from aqueous solutions using trioctylamine (TOA) as a reactive extractant and isoamyl alcohol or 1-octanol as active diluents. Giving weight to harder-to-extract acids, the optimum conditions using isoamyl alcohol were found to be a 0.38 organic/aqueous loading ratio, 13.8 TOA wt% in organic phase, and a 1.41 lactic/levulinic acid ratio in the initial solution. Conversely, 1-octanol extraction was optimized at a 0.63 organic/aqueous phase ratio, 25.4% TOA concentration, and a 0.87 lactic/levulinic ratio.

For the separation of these short-chain carboxylic acids, it was found that isoamyl alcohol is nearly twice as effective as 1-octanol. However, the high solubility of isoamyl alcohol in water as well as its similar volatility to two of the more volatile by-product acids, acetic and formic, offset the effectiveness advantage for a commercially relevant design. To weigh these effects, a process design and economic analysis was performed to better understand the economics associated with both diluents. Doing so revealed that additional separation steps for the isoamyl extraction, namely extractive distillation for acetic acid and formic acid purification from water is more costly when isoamyl alcohol is used rather than 1-octanol, which has a simpler separation process.

Chapter IV - Techno-Economic Analysis of a Process for the Aqueous Conversion of Corn Stover into Lactic and Levulinic Acid through Sn-Beta + CaSO4 Catalysis[*]

1. Introduction

Changes to the global environment have resulted in a concerted effort to transition away from petroleum as a feedstock for fuel and chemical production. Lignocellulosic biomass is a desirable alternative carbon source as the world's ability to produce lignocellulosic biomass is substantial. One of the most readily available sources in the United State is corn stover, which consists of the leftover stalks, leaves, husks, and cobs from the corn plant and makes up nearly half of the yield of a corn crop. These components are rich in structural carbohydrates, comprising about 60% of the dry weight (35.1% glucan and 19.5% xylan) [56]. The structural nature of corn stover makes it difficult to breakdown into sugars which can be used to produce renewable fuels and chemicals. Common practice for lignocellulosic materials like corn stover is to pretreat it with sulfuric acid [61]. The acid breaks down the hemicellulose releasing xylose and makes the rest of the carbohydrates more accessible. Enzymes are then utilized to break apart the cellulose, releasing glucose [56].

Using a Sn-Beta catalyst, it is possible to convert the monomeric sugars (glucose and xylose) to more lucrative products such as lactic and levulinic acid as well as by-products such as formic and acetic acid [8, 10, 41, 89]. The reaction of glucose follows one of two reaction pathways. The first is retro-aldol condensation which splits the six carbon sugar in the presence of a Lewis Acid [45]. This produces 2 three-carbon trioses, which can

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[*] An initial version of this work was developed with the assistance of the following UND Chemical Engineering students as part of CHE412, Plant Design 2 during the Spring 2019 semester: Cara Bjerke, Jacob Dahl, and Cassandra Shaffer

both be isomerized to lactic acid [45]. As a side reaction, these trioses can decompose to acetic and formic acids [53]. Glucose can also triple dehydrate in the presence of a Brønsted Acid into 5-hydoxymethylfurfural (5-HMF)[46]. In the presence of water, a 5-HMF molecule can rehydrate and form a five carbon levulinic acid molecule and a one carbon formic acid molecule [46]. Xylose, a pentose, is stoichiometrically limited to one lactic acid through retro-aldol condensation. Additionally, when dehydrated, xylose transforms into five-carbon furfural.

Lactic acid is a renewable chemical that can serve as an alternative to fossil fuel derivatives. It serves as a platform chemical in applications across a wide-range of industries including cosmetics, food, pharmaceuticals, and, most prominently, biodegradable plastics [12-14, 21]. Traditional synthesis methods through fermentation have proven costly, making it difficult for its subsequent products to economically compete. More economical production can be achieved by chemical pathways, instead of biological. This is not optimal for its main application, polylactic acid (PLA), as it produces a racemic mixture of lactic acid. However, this mixture can still be used in biodegradable polymer production through intermediate conversion to acrylic acid or propylene glycol [21].

Initial studies found that optimum lactic acid yield was achieved using methanol as the reaction solvent [9, 44]. This results in the formation of ester derivatives of the target products (i.e. methyl lactate instead of lactic acid). Production of these derivatives would require alternative uses for the end-products or subsequent conversion of the esters to carboxylic acids. An additional drawback to this pathway is that sugars are hydrolyzed from biomass in aqueous solutions. Thus, energy-intensive steps would be required to extract the sugars out of their water solution so they can be transferred to methanol as well as to extract the product methyl ester forms of the product acids out of the methanol solvent for use as products.

Our recent work has found that the using Sn-Beta for catalytic sugar reactions yielded 13% lactic acid and 18% levulinic from glucose on a weight percent of inlet carbon basis. Similarly, using xylose as a substrate Sn-Beta yielded lactic acid at 19 carbon wt% and levulinic acid at 1.8 carbon wt%. The addition of $CaSO_4$ to aqueous reactions mixtures can result in lactic acid yields comparable to those achieved in methanol. This is due to sulfate ions present in the solution that help to neutralize Brønsted Acid sites present on the surface of Sn-Beta. The loss of these sites suppresses the production of dehydration side-products. Additionally, calcium ions act as a second Lewis Acid, acting in tandem with tin to improve the yield of the retro-aldol pathway. This results in a lactic acid yield of 68 wt% of the inlet carbon, significantly closer to the 75% yield obtained in methanol [9].

This chapter presents the results of a techno-economic analysis (TEA) of the commercial feasibility to utilize Sn-Beta facilitated reactions to produce biorenewable chemicals from corn stover. Two process designs are evaluated which represent two separate reaction yields from Sn-Beta. The first utilizes our original base Sn-Beta conditions to primarily produce a balanced yield of both lactic and levulinic acid. The second uses the co-catalysis of Sn-Beta and $CaSO_4$ to almost exclusively produce lactic acid. The reaction data that forms the basis of this TEA are provided in Chapter II.

Both process designs are scaled to utilize 30 wt% of the corn stover produced from the quantity of corn fed to a typical traditional ethanol plant located in the Upper

Midwest of the U.S. The remaining 70 wt% of the harvested corn stover is typically recycled back into the soil for quality preservation [56]. This equates to about 230,000 metric tons of corn stover per year as feed to the process for both alternatives.

2. Experimental methods and materials

2.1 Design

A preliminary design for each process was created by identifying and approximating using preliminary design techniques, the necessary process steps to transform corn stover into the target organic acids and the subsequent separation steps to obtain saleable purities. Temperature and pressure profiles were developed to estimate heat transfer and motive force requirements. Equipment sizing was estimated using Ulrich [90] and ChemCad™. Corn stover hydrolysis design was based off of current and target yields specified by the National Renewable Laboratory (NREL)[56, 59]. In addition to equipment sizing, utilities and chemical requirements were also estimated. The process design was summarized and depicted in process flow diagrams (Appendix J).

2.2 Economic Analysis

A broad capital cost estimate was developed using an American Associate of Cost Engineers (AACS) Class 4 factored estimating method [Cost Estimate Classification System – as Applied in Engineering, Procurement, and Construction for the Process Industries, AACE International Recommended Practice No. 18R-97, 2005] at an accuracy of ±40%. Vendor estimates were obtained for compressors and rotary vacuum filters. All other equipment costs plus all of the estimating factors were estimated using cost charts in Ulrich [91] and used to calculate a total bare module cost. Contingency and fees of 18% of the total bare module cost were added to produce the total module cost. To account for building costs, site preparation, and offsite areas, an auxiliary factor of 30% of the total module cost was included to yield the fixed capital investment (FCI). The costs for the initial charge of solvents and catalysts and working capital (15% of FCI) were added to the FCI to yield a total capital investment (TCI).

In addition, revenues and operating costs were estimated at an equivalent level of accuracy. The unit prices of all products, raw materials, chemicals, and catalysts were calculated based on their historical price trend at the basis date of January 2019. When price history was unavailable, spot prices were used. The following additional operating expenses were estimated: operating labor, maintenance, utilities, wastewater treatment, and general allocated expenses. Utility costs were calculated for steam, cooling and boiler feed water from prices estimated with correlations based capacity and fuel cost found in Ulrich [91].

Profitability was calculated using a cash flow sheet method with an assumed facility operating life of 20 years. The fixed capital investment was spread across the preliminary project development schedule (Years -2 to 0). Initial solvent and catalyst expenses and working capital were applied immediately before operation (Year 0) and the working capital was reimbursed in the final year of operation (Year 20). Income tax requirements were estimated by calculating the gross profit (revenue less the operating costs). Depreciation based on a 17 year MACRS (full year convention) of the fixed capital investment was subtracted from the gross profit for each year of operation, yielding the federal taxable profit. Federal taxes were applied at the 2019 rate of 21%. State taxable profit was calculated as the federal taxable profit less the federal income tax. State tax was applied at the 2019 North Dakota rate of 4.31% [92]. Net profit after process commissioning (Years 1-20) was calculated as gross profit minus federal and state taxes. Annual net profits were discounted to year 0 using a hurdle rate (discount factor) of 20%. The NPV@20% was calculated as the sum of the discounted net profit of each year and the DCFROR was calculated by finding the hurdle rate at which the NPV@DCFROR=$0.

2.3 Process Design Assumptions

1. Processes were designed as grassroots plant with a 20 year lifespan.
2. Designed processes have an operating factor of 95%.
3. Corn Stover is available pre-milled at a mean size 5 mm and a density of 560 kg/m^3.
4. Corn stover has an initial composition (dry wt%) of 35.1% glucan, 19.5% xylan, 15.8% lignin, 4.9% ash, 3.1% protein, 2.4% arabinan, 1.8% acetate, 1.4% galactan, 0.8% sucrose, 0.6% mannan, and 14.7% extractives [59]. Extractives component is a general term used to sum the initial composition of corn stover to 100% [59].
5. Deacetylation requires a sodium hydroxide loading of 17 mg/g dry biomass [59]. Sodium hydroxide is available in a 0.4 wt% solution with water [59].
6. Deacetylation removes 100% of water-soluble extractives, 88% of acetate, 75% of ash, 20% of lignin, and 2% of xylan from the initial corn stover [59].
7. Solids leaving the deacetylation reactor are assumed to be 20 wt% liquid and all are removed from the screw and recycled back to the reactor [59].

8. Acid hydrolysis requires a sulfuric acid loading of 9 mg/g of dry biomass. Sulfuric acid is diluted to 0.3 wt% in water [59].
9. At 160° C and 560 kPa for 5 minutes, the acid hydrolysis reactor has a xylan conversion of 90 wt% to xylose, 5 wt% to furfural, and 2.4 wt% to xylose oligomers. Glucan conversion are 9.9 wt% glucose, 3 wt% glucose oligomers, and 3 wt% HMF. 100 wt% of the remaining acetate is converted to acetic acid.
10. Corn stover has a heat capacity of 1300 J/(kg-K) [93].
11. Heats of formation used are -1271 kJ/mol for glucose, -973 kJ/mol xylose, -484.5 kJ/mol for acetic acid, -425 kJ/mol for formic acid, -590 kJ/mol for lactic acid, -581 kJ/mol for levulinic acid, and -286 kJ/mol for water[94].
12. Enzymatic hydrolysis requires an enzyme loading of 10 mg/g of cellulose and operates at a residence time of 84 hr [59].
13. Enzymatic hydrolysis has glucan conversions of 91.2 wt% to glucose and 4 wt% to glucose oligomers [59].
14. Since HMF is an intermediate of levulinic acid formation, it is assumed that all HMF produced from acid hydrolysis is converted to levulinic acid and formic acid with Sn-Beta.
15. Glucose and xylose are assumed to have 100% conversion in the Sn-Beta reactor. Sugars that are not converted into organic acid products through Sn-Beta coke assumed to be hydrochar [47].
16. Hydrochar can be calcined off the catalyst using compressed air at 260° C and 1800 kPa.

17. Desired efficiency of reactive extraction for both TOA and TMA extractors were 99% for all product acids.

2.4 Equipment Design Assumptions

1. Conveyors anticipated to stay within one process area are 15 m long. Conveyors anticipated to go between process areas sized at 150 m in length. All conveyors are 0.61 m wide.
2. Pumps operate at 65% efficiency.
3. Compressors have a polytrophic efficiency of 75%
4. Hold up time in a reflux drum is 450 s.
5. The height to diameter ratio used for reflux and hold up drums was 2.
6. The liquid residence time for hold up drums was 5 minutes and the total drum volume was double the liquid space.
7. The liquid residence time for mixers was 5 minutes and the height to diameter ratio of 2.
8. Distillation columns were sized with sieve trays at a spacing of 0.61 m.
9. A 20 kPa pressure drop was assumed across all units with an additional 14 kPa for an anticipated control valve and 20 kPa for a mist eliminator.
10. A 70 kPa pressure drop was assumed for every 4 meters of catalyst packing.
11. Trayed columns had pressure changes of 0.7 kPa per tray, with pressure decreasing with increased height. Other pressure drops for distillation columns include 100 kPa for feed above 9 m and 1 kPa for thermosyphon reboilers.

12. Interstage cooling systems for multistage compressors have a 14 kPa pressure drop across a heat exchanger and 7 kPa across a knockout drum.

2.5 Utility Assumptions

1. Steam grades produced in the auxiliary facilities include: low pressure steam at 148° C and 450 kPa, medium-low pressure steam at 175° C and 900 kPa, moderate pressure steam at 204° C and 1700 kPa, medium-high pressure steam at 3300 kPa, and high pressure steam at 400° C and 4500 kPa. All grades of steam are rated at their saturation pressures except for high pressure steam which is superheated.
2. Cooling and boiler feed water are produced in the auxiliary facilities at 30° C and 310 kPa.

2.6 Economic Assumptions

1. Economic estimates were brought to a basis date of January of 2019 using CEPCI values of 400 and 616 for January 2004 and January 2019, respectively.
2. Annual maintenance costs were estimated as 4% of the fixed capital investment.
3. The fixed capital investment was depreciated over 17 years using the Modified Accelerated Cost Recovery System (MACRS).
4. Income tax rates were applied as 21% for Federal and 4.31% for State (North Dakota)[92].

5. A rough project completion schedule was estimated using a 30% design, 40% procurement, and 30% implementation rule of thumb. Procurement was based on the longest equipment lead time (14 months) with no overlap of project phases, yielding a total schedule of 36 months.
6. No royalties or patent fees are required for the process
7. Recycled solvents are replaced at a rate of 15% per year while solid catalysts and sorbents are replaced at a rate of 5% per year.
8. Solid catalysts and sorbents are completely changed-out every 4 years.
9. The loaded cost of an operator in a chemical plant in North Dakota is $30/hr [95].
10. A natural gas price is $4.4/GJ was used in utility cost correlations [96].
11. The price of electricity was assumed to be the 2019 industrial rate in North Dakota, $0.0619/kWh [97]
12. The project Hurdle Rate (minimum acceptable rate of return) is 20%.

3. Results and discussion

3.1 Process Design – Sugar Recovery and Conversion

Both processes were scaled to utilize the corn stover generated by farmers who sell their corn to a typical bioethanol plant in the Midwest Corn Belt. This would allow for the utilization of existing distribution networks, ensuring proper supply of the raw material. A single bushel of corn (about 25 kg) produces about 10.6 liters of ethanol[98] and every kg of harvested corn produces about 1.0 kg of corn stover [56]. Not all corn stover is usable, however, since 70% of it is required to remain in the soil for erosion control[56]. In

North Dakota, the average nameplate capacity of bio-refinery is about 360 million liters per year [99], making a process feed of about 230,000 metric tons of corn stover per year.

An overview of both process alternatives can be seen in Figure 17. Both alternatives were split into five process areas. Process area 01 is the corn stover pretreatment area. Process area 02 is the glucose recovery area. Process area 03 is the Sn-Beta catalytic reaction area. Process area 04 is the acetic and formic acid purification area and Process area 05, seen on is the lactic and levulinic acid purification area. Complete Process Flow Diagrams (PFDs) with more detailed information on both designs are provided in Appendix J.

Equipment requirements for both process alternatives were similar, with the only functional difference originating in the addition of calcium sulfate before the catalytic reactor in the Sn-Beta + $CaSO_4$ alternative. This addition significantly shifts the selectivity of the reaction to almost exclusively produce lactic acid. Reaction yields applied to both processes can be seen on Table 20 and were based on the experimental work described in Chapter II. Subsequent separation units in both process alternatives have identical functionality and are scaled to fit the differing reaction yields.

Table 20. Reaction yields for each evaluated process alternative (wt% of Inlet Carbon).

Alternative	*Sugar*	*Lactic Acid*	*Levulinic Acid*	*Acetic Acid*	*Formic Acid*
Sn-Beta	Glucose	11%	20%	2.4%	8.0%
Sn-Beta	Xylose	19%	1.8%	2.4%	7.4%
Sn-Beta + $CaSO_4$	Glucose	68%	2.0%	3.6%	3.2%
Sn-Beta + $CaSO_4$	Xylose	50%	2.5%	5.2%	5.4%

Process areas 01 and 02 are identical for both processes and are focused on the extraction and recovery of glucose and xylose from corn stover. In process area 01, corn stover is fed into the process and initially treated with a 0.4 wt% sodium hydroxide solution. This treatment acts primarily to remove acetate groups that are mostly included in the hemicellulose portion of the biomass. Removal of the acetate groups allows for easier access to the xylan in the subsequent acid hydrolysis step [61]. The xylan in the hemicellulose is then depolymerized with 0.3 wt% sulfuric acid in the acid hydrolysis reactor to produce xylose. The remaining corn stover solids are then removed and rinsed with process water in a rotary vacuum filter. The filtered acid hydrolysate is then neutralized with calcium carbonate, producing calcium sulfate.

The corn stover solids remaining after acid hydrolysis are routed to process area 02. Here the corn stover solids are mixed with cellulase enzymes, which react with the cellulose portion of the solids, liberating glucose. This enzymatic hydrolysis occurs at a solids loading of 20 wt%, which is too high for the solution to be pumped. Therefore, the reaction is performed in a vertical continuous reactor allowing mixture transport by gravity. At a residence time of 24 hours, the enzymes break down the solids enough for them to be pumped to a batch reactor. The remainder of the enzymatic reaction occurs in this reactor for an additional 60 hours. This results in an overall 95 wt% conversion of inlet cellulose to glucose. The design for the first two process areas are consistent with target sugars yields from corn stover outlined by NREL for the catalytic conversion of sugars into hydrocarbons [59].

The sugars extracted by acid and enzymatic hydrolysis in process areas 1 and 2 are then combined in process area 03. For the Sn-Beta + $CaSO_4$ alternative, calcium sulfate is added at this point to produce a saturated solution. The sugar solutions are then fed to multiple parallel Sn-Beta reactors where competing retro-aldol condensation and dehydration reactions produce lactic, levulinic, acetic, and formic acids. The reactors are designed for a reaction temperature of 200° C and a residence time of 2 hr. The balance of the carbon not specified in the reaction yields on Table 20 is assumed to be converted into hydrocar as a result of intra- and intermolecular dehydration of the sugars [47].

It is assumed that the hydrochar can be calcined off the catalyst with compressed air at 260° C. To accomplish this, 2 of the 18 available reactors are offline at any time allowing for continual catalyst regeneration. For a reactor to go offline, its outlet lactic acid concentration would have to drop below an acceptable standard, indicating subpar catalyst activity. Once this occurs, the reactor is then drained and fed with compressed air. The hydrochar on the catalyst is then calcined to carbon dioxide until completely removed. The reactor will then be reactivated once a separate reactor reached the minimum reaction conversion.

Also in process area 03, furfural generated as a side product through acid hydrolysis is removed from the process stream. This is accomplished by extracting the furfural using toluene, which does not have any affinity for the produced carboxylic acids.

3.2 Process Design – Acid recovery and purification

In both alternatives, process areas 04 and 05 were designed to recover and purify the target acid products. The initial hurdle in product purification is to remove the water from the dilute acid stream (acid concentration is ~4 wt%). This can be done by reactive extraction where a basic tertiary amine can react with the carboxylic acids yielding a non-polar ion-pair [70, 73, 78]. The high viscosity and tension of these amines create transport limitations when used on an industrial scale [62]. To mitigate these a diluent can be added to improve its physical properties and aid in separation [62]. Though there is a general consensus on which amine extractant (trioctylamine, TOA) and diluents (C_5-C_{10} alcohols) are most effective for extraction of these acids, consistent conditions for their co-extraction are not available [74, 76-78].

For both alternatives, reactive extraction conditions were determined by the optimization experiments conducted in Chapter III. However, screening and optimization experiments occurred in carboxylic acid solutions that did not contain any calcium sulfate. The presence of calcium sulfate would decrease the acidity of the solution, and in turn, decrease the equilibrium constant for ion-pair formation. To account for these conditions, a cursory experiment was conducted to extract carboxylic acids produced using the Sn-Beta in a $CaSO_4$ solution (Table 21). Doing so significantly lowered the distribution coefficients all of the target acids. The most significant drop occurred with lactic acid, which is likely due to a significant increase in its formation from the more favorable reaction conditions. The higher concentration coupled with lower solution acidity worked to reduce the equilibrium of ion-pair formation with TOA.

Table 21. Distribution coefficients at optimum 1-octanol/TOA reactive extraction conditions for carboxylic acid solutions produced using Sn-Beta and Sn-Beta + $CaSO_4$.

	Sn-Beta	Sn-Beta + $CaSO_4$
Lactic Acid KD	35	1.9
Levulinic Acid KD	21	6.4
Acetic Acid KD	32	2.3
Formic Acid KD	102	3.4

Applying this to the design of the Sn-Beta and Sn-Beta + $CaSO_4$ processes, reactive extraction with TOA/1-octanol was utilized in process area 04 with the distribution coefficients specified in Table 4. Within that area, both acetic and formic acids are separated from the extract through distillation. The resulting acetic/formic acid distillate is than further distilled to reach their respective product grades. The remaining TOA/1-octanol extract is sent to process area 05 where lactic and levulinic acids are back extracted into TMA/water. Both solvents are regenerated and lactic and levulinic purified through distillation in the final process area. An overview for both processes can be seen on Figure 17.

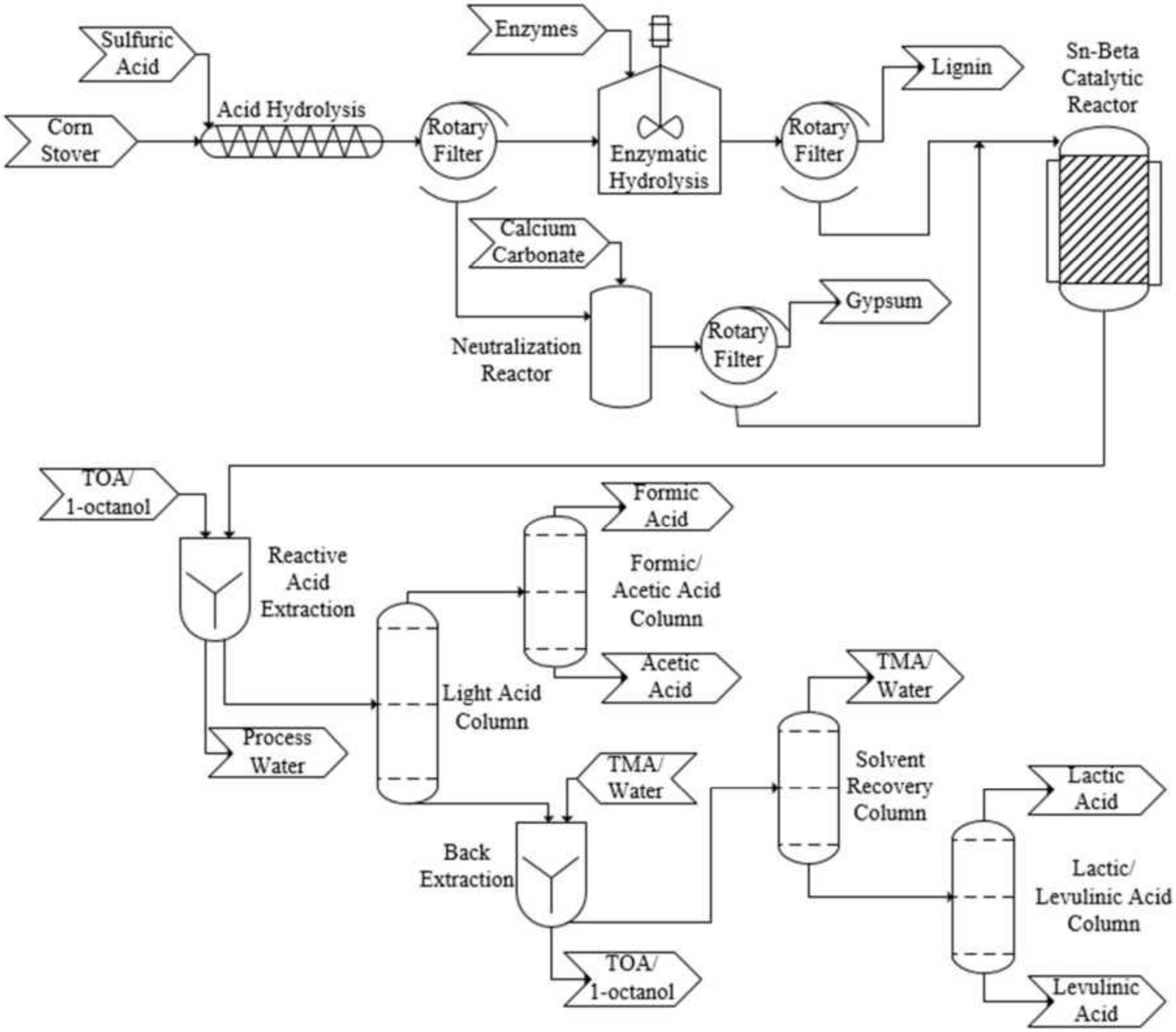

Figure 17. Major Unit Operation Requirements for Sn-Beta and Sn-Beta + CaSO4 Process Designs

3.3 Capital Cost Estimates

A category 4 estimate of capital costs was conducted based on equipment required for the process of pump size or larger. The condensed estimated capital costs can be seen in Tables 5 and 6. These tables are condensed based on unit classification. The broadest of these classifications is process vessels which includes distillation columns and trays, drums, and mixing vessels. Tables 22 and 23 also include the additional investments required as a part of the total capital investment at a basis date of January 2019. More detailed capital cost information can be seen in Appendixes H and I.

Vendor estimates were obtained for compressors and rotary vacuum filters. Costs for the remaining capital equipment were estimated utilizing cost charts published by Ulrich and Vasudevan [100]. These cost charts provided estimates at a basis date of 2004. To bring these estimates to a January 2019 basis, CEPCI values of 400 and 616 were used for 2004 and 2019, respectively.

Equipment requirements for the two process alternatives were nearly identical. Any differences were due to differences in product capacity due to the catalytic reaction yield. A major sizing difference that occurred between the two alternatives was for the catalyst calcination compressor due to increased hydrochar formation in the base Sn-Beta processes. Additional mixers and settlers were required for the Sn-Beta + $CaSO_4$ process since lower efficiency in the reactive extraction step requires three additional extraction stages. The fixed capital investment was estimated at $130 million ± 40% for both alternatives.

Additional capital costs included working capital and the initial inventory of chemicals and catalysts. As a rule of thumb, the working capital was estimated to be 15% of the FCI. Costs were included for the initial inventories of 1-octanol, TOA, TMA, toluene, Sn-Beta, 13x molecular sieves, and water. Vendor quotes were obtained for TOA, TMA and 1-octanol, being valued at \$8.49/kg, \$2.64/kg, and \$1.54/kg, respectively. Toluene was estimated to be priced at \$1.17/kg [101]. Water was priced as demineralized water through Ulrich [91]. Based on the plant capacity for demineralized water, that price was estimated to be \$0.55 per 100 kilogram [91]. Sn-Beta and 13x molecular sieves were priced as Silica/Alumina doped catalysts and zeolites in Ulrich, respectively[100]. Sn-Beta were then priced as \$12.30/kg and 13x molecular sieves as \$10.80/kg. The capital costs for the initial inventory of all chemicals and catalysts was valued at \$2.1 million and \$2.2 million for the base Sn-Beta and Sn-Beta + $CaSO_4$ processes, respectively. This slight difference is due to a higher acid concentration in the Sn-Beta + $CaSO_4$ process, resulting in a larger solvent requirement. Adding in working capital and the initial chemical and catalyst costs to the FCI, yielded a total capital investment (TCI) of \$150 million for both process alternatives.

Table 22. Capital Investment for Base Sn-Beta Reaction Process Alternative

Equipment Type	**# of Units**	**PEC (2019)^{A}**	**Total BMC^{B}**
Process Vessels	221	$ 3,400,000	$ 20,000,000
Heat Exchangers	51	$ 3,100,000	$ 12,000,000
Tanks	2	$ 74,000	$ 260,000
Compressors	7	$ 2,500,000	$ 3,800,000
Filters	11	$ 2,600,000	$ 11,000,000
Conveyors	36	$ 1,200,000	$ 2,600,000
Pumps	74	$ 1,100,000	$ 4,200,000
Mixers	23	$ 5,200,000	$ 11,000,000
Reactors	36	$ 3,300,000	$ 19,000,000

Total Bare Modular Cost		**C_{TBM}^{C} »**		$ 83,000,000
	Contingency and Fees		**C_{TBM}^{A} * 0.18 =**	$ 15,000,000
Total Module Cost		**C_{TM}^{D} »**		$ 97,000,000
	Auxiliary Facilities	**C_{AUX}^{E}**	**C_{TM}^{B} * 0.30 =**	$ 29,000,000
Fixed Capital Investment		**FCI^{F} »**		$ 130,000,000
	Working Capital	**C_{WC}^{G} »**	**FCI^{D} * 0.15 =**	$ 19,000,000
	Initial Charge of Chemicals and Catalysts			$ 2,100,000
Total Capital Investment		**TCI^{H} »**		$ 150,000,000

A. PEC – Purchased Equipment Cost at basis date
B. BMC – Bare Module Cost
C. C_{TBM} – Total Bare Module Cost
D. C_{TM} – Total Module Cost
E. C_{AUX} – Auxiliary Facilities Cost
F. FCI – Fixed Capital Investment
G. C_{WC} – Working Capital
H. TCI – Total Capital Investment

Table 23. Capital Investment for Sn-Beta + CaSO4 Reaction Process Alternative[A]

Equipment Type	**# of Units**	**PEC (2019)**	**Total BMC**	
Process Vessels	227	$ 3,800,000	$ 22,000,000	
Heat Exchangers	50	$ 3,300,000	$ 12,000,000	
Tanks	2	$ 74,000	$ 260,000	
Compressors	7	$ 1,600,000	$ 2,400,000	
Filters	11	$ 2,600,000	$ 11,000,000	
Conveyors	36	$ 1,200,000	$ 2,600,000	
Pumps	74	$ 1,100,000	$ 4,200,000	
Mixers	23	$ 5,200,000	$ 11,000,000	
Reactors	36	$ 3,300,000	$ 19,000,000	
Total Bare Modular Cost		**C_{TBM} »**		$ 85,000,000
	Contingency and Fees		**C_{TBM}^{A} * 0.18 =**	$ 15,000,000
Total Module Cost		**C_{TM} »**		$ 100,000,000
	Auxiliary Facilities	**C_{AUX}**	**C_{TM}^{B} * 0.30 =**	$ 30,000,000
Fixed Capital Investment		**FCI »**		$ 130,000,000
	Working Capital	**C_{WC} »**	**FCI^{D} * 0.15 =**	$ 20,000,000
	Initial Charge of Chemicals and Catalysts			$ 2,300,000
Total Capital Investment		**TCI »**		$ 150,000,000

A. Terms in this table are defined in Table 5

3.4 Operating Cost Estimates

Annual operating costs can be seen in Table 25. This table provides a breakdown of the primary expenses that would be incurred for both process alternatives. Those expenses include: raw material costs, makeup chemical and catalyst costs, operating labor costs, maintenance costs, the costs for utilities and waste disposal, and other factors (R&D charges plus marketing & distribution charges). The average annual operating costs were estimated at $79 million and $80 million for the base Sn-Beta and Sn-Beta + $CaSO_4$ processes, respectively, on a January 2019 basis.

This process utilizes one raw material: corn stover. Corn stover prices are currently around $50 per dry ton. However, if corn stover were to be adopted as a commodity chemical for agricultural processing, the National Renewable Energy Laboratory expects the price to increase to $80 per dry ton[59], which was the price used in this TOA. This results in a corn stover cost about $0.09/kg and an overall raw material cost of $21 million per year for both process alternatives.

Consumable chemicals and catalysts used for by both processes include sulfuric acid, calcium carbonate, cellulase enzymes, sodium hydroxide, and make-up amounts for the reusable chemicals and catalysts purchased as part of the TCI. Prices of sulfuric acid and calcium carbonate were estimated at spot prices of $0.33/kg and $0.11/kg, respectively. The cellulase enzymes used for enzymatic hydrolysis were estimated to cost about $3.52/kg. Sodium hydroxide was valued at about $400/metric ton, which is equivalent to $0.40/kg [102]. As a rule of thumb, all recycled chemicals were assumed to experience a yearly depletion of 15% via losses and/or degradation. These included 1-octanol, TOA, TMA, toluene, and water, which were priced at $1.54/kg, $8.49/kg,

\$2.64/kg, \$1.17/kg and \$0.25/m^3, respectively. Similar to the chemicals being recycled, catalysts and solid sorbents were assumed to have yearly losses of 5%. This loss is applicable to both Sn-Beta and 13x molecular sieves at \$12.30/kg and 10.80/kg, respectively. Since the base material of both of these are zeolites, a total change out was forecast to be required every four years. For the base Sn-Beta process alternative, the total price for consumable chemicals and catalysts in a typical year was \$4.5 million with \$660,000 of that being spent on make-up materials each year. The Sn-Beta + $CaSO_4$ process alternative required \$4.6 million in consumable chemicals a year with \$700,000 required for make-up. This difference is due to the increased solvent requirement for product extraction in the Sn-Beta + $CaSO_4$ process.

Operating labor requirements were based on the number and type of unit operations required for each process. Due to similar unit operation requirements, both processes were estimated to require 90 operators split between 4.5 shifts to provide for 24/7 coverage. The loaded pay rate for a plant operator in North Dakota of \$33.30 per hour was used [103]. Using a 40-hour work week basis, the labor cost of this project was estimated at \$7.2 million per year. This includes an extra 15% to account for supervisory wages.

Maintenance costs for both alternatives were estimated to be 4% of the FCI. This results in a yearly cost of \$5.1 million for the base Sn-Beta process alternative and \$5.2 million for the Sn-Beta + $CaSO_4$ process alternative.

Utility requirements were calculated based on the usage of cooling water, boiler-feed water, process steam, and electricity throughout each process alternative. Unit costs of cooling water, boiler-feed water, and process steam were calculated using equations from Ulrich[91]. Each utility equation is based on a combination of the price of energy and capital requirements for utility production. Utility costs were estimated using the Wahpeton, North Dakota industrial rates of electricity, \$0.062/kWh, and natural gas, \$4.4/GJ[96, 97]. Though used throughout both process alternatives, low (148°C, 450 kPa) and medium low (175° C, 900 kPa) pressure steam is not included in this table. This is because an excess of the two grades are produced through the vaporization of boiler feed-water used for cooling applications in many of the heat exchangers.

Between the two process alternatives, the biggest difference in utility costs results from differences in electricity requirements. This is due to higher hydrochar production in the base Sn-Beta alternative. As a result more compressed air is required for catalyst calcination, leading to a larger compressor duty and subsequent higher use of compressor motor power. An additional difference is that the Sn-Beta + $CaSO_4$ alternative requires more than double the high pressure steam than the base Sn-Beta alternative. This is due to an increase of lactic acid, resulting in a higher reboiler duty for the TMA/water recovery column. Since the reboiler operates higher than the condensation point of high pressure steam, the cooled utility is then used as medium-high pressure steam, lowering its production requirement. Overall, the annual utility requirements were estimated to be \$29 million for both process alternatives.

Black liquor wastewater produced through deacetylation needs to be treated before it is released. To do so the stream will be treated through filtration and activated sludge. The sludge will be used to remove the carbonaceous matter (acetate and lignin) removed from the corn stover. Based on the estimation equation from Ulrich[91] the processing of 120,000 kg/hr of wastewater will be charged at a rate of \$0.26/m^3, totaling

$280,000 per year. Due to the calcium sulfate dissolved in the reaction solvent, this solvent is also treated after the acid extraction step. Since both processes have calcium and sulfate concentrations exceeding emission standards and these salts are not biodegradable, tertiary treatment is required[91]. For the base Sn-Beta process 150,000 kg/hr of water is estimated to require treatment at a cost of around $990,000 per year. For the Sn-Beta + $CaSO_4$ process, part of this solvent is recycled to the catalytic reactor, resulting in only 130,000 kg/hr being treated at $880,000 per year.

Other operating expenses include research and development, marketing and distribution, and general administration. Research and development expenses were estimated at 5% of the total operating expenses (including raw materials, chemicals and catalysts, operating labor, maintenance, and utilities. This resulted in a cost of $3.4 million for both process alternatives. Marketing and distribution fees were similarly calculated as 10% of the total operating expenses, resulting in $6.7 million and $6.8 million for base Sn-Beta and Sn-Beta + $CaSO_4$ alternatives, respectively. General administration costs were estimated to be 25% of the overhead. Overhead is defined as 60% of the operating labor and maintenance expenses. General administration was estimated as $1.8 million for the base Sn-Beta alternative and $1.9 million for the Sn-Beta + $CaSO_4$ alternative.

Table 24. Utility requirements for each process.

	Base Sn-Beta Alternative		*Sn-Beta +$CaSO_4$ Alternative*	
Utility Description	Annual Cost ($/yr)	Requirement	Annual Cost ($/yr)	Requirement
Electricity	$5,600,000	Amount: 90 million kWh/yr	$3,600,000	Amount: 59 million kWh/yr
Cooling Water	$660,000	Amount: 3.4 billion kg/yr Temperature: 30°C Pressure: 300 kPa	$710,000	Amount: 4.4 billion kg/yr Temperature: 30°C Pressure: 300 kPa
Boiler Feed Water	$6,700,000	Amount: 1.8 billion kg/yr Temperature: 30°C Pressure: 450 kPa	$7,600,000	Amount: 2.0 billion kg/yr Temperature: 30°C Pressure: 450 kPa
Medium-High Pressure Steam	$9,100,000	Amount: 480 million kg/yr Temperature: 239°C Pressure: 3,300 kPa	$3,400,000	Amount: 160 million kg/yr Temperature: 239°C Pressure: 3,300 kPa
High Pressure Steam	$6,400,000	Amount: 320 million kg/yr Temperature: 400°C Pressure: 4,500 kPa	$14,000,000	Amount: 730 million kg/yr Temperature: 400°C Pressure: 4,500 kPa
Total	$29 million/yr		$29 million/yr	

Table 25. Breakdown of annual operating expenses for each process.

Process Alternative	Year	Raw Materials	Chemicals & Catalysts	Operating Labor	Maintenance	Utilities	Waste Disposal	Other Expenses	Total
Base Sn-Beta	1-20	$21,000,000	$4,600,000	$7,200,000	$5,100,000	$29,000,000	$1,300,000	$12,000,000	$79,000,000
Sn-Beta + $CaSO_4$	1-20	$21,000,000	$4,900,000	$7,200,000	$5,200,000	$29,000,000	$1,200,000	$12,000,000	$80,000,000

3.5 Revenues

Revenues for both process alternatives were generated by the sale of the produced organic acids and side-products as well as through recovered energy. Spot prices were applied to all six saleable products. Those include lactic acid at \$1.98/kg, levulinic acid at \$5.00/kg, acetic acid at \$0.55/kg, formic acid at \$0.68/kg, furfural at \$1.65/kg, and gypsum at \$0.04/kg [104-110]. Sale of products are expected to produce \$120 million and \$190 million in annual revenue for the Sn-Beta and Sn-Beta + $CaSO_4$, respectively.

In addition to revenue from the products, two operating credits also generate significant revenue. The first is the saturated water credit. This credit is available because the base process design produces more low-pressure steam then is consumed by that process. To avoid the production of useless steam, it was decided to use boiler-feed water rather than cooling water to cool the process stream before furfural extraction and to then heat it to just below its saturation point with the excess steam. The saturated water is the sent to the utility boiler to produce the grades of steam that are not in excess. Heating the water to its saturation point consumes about 21% of the energy required to make this steam. This energy recovery saves \$3.2 million and \$3.5 million per year for the two respective process alternatives in steam production costs and is represented as a revenue instead of just reducing the steam utility costs.

The other credit is the lignin energy credit. After corn stover hydrolysis, a lignin-rich stream is produced. Instead of disposing of the lignin, it is sent to the utility boiler to generate heat in place of natural gas. By burning 5,500 kg/hr of lignin, over 110 GJ/hr is produced. Since natural gas costs \$4.4/GJ, lignin energy generation saves about \$4.2 million per year for both process alternatives. Including the credits, the total annual revenue is \$140 million and \$200 million for the Sn-Beta and Sn-Beta + $CaSO_4$ alternatives, respectively.

Table 26. Revenue Projections for Base Sn-Beta and Sn-Beta + CaSO4 Alternatives.

		Base Sn-Beta Alternative		Sn-Beta + $CaSO_4$ Alternative	
Products	\$/kg	Amount (kg/yr)	Revenue (\$/yr)	Amount (kg/yr)	Revenue (\$/yr)
Lactic Acid	\$ 1.98	18,000,000	\$ 35,000,000	78,000,000	\$ 150,000,000
Levulinic Acid	\$ 5.00	15,000,000	\$ 74,000,000	4,200,000	\$ 21,000,000
Acetic Acid	\$ 0.55	3,000,000	\$ 1,700,000	5,100,000	\$ 2,800,000
Formic Acid	\$ 0.68	15,000,000	\$ 10,000,000	8,000,000	\$ 5,400,000
Furfural	\$ 1.65	1,600,000	\$ 2,600,000	1,600,000	\$ 2,600,000
Gypsum	\$ 0.04	1,000,000	\$ 36,000	--	--
Saturated Water Credit	--	--	\$ 3,100,000	--	\$ 3,500,000
Lignin Energy Credit	--	--	\$ 4,300,000	--	\$ 4,200,000
		Total	**\$130,000,000**		**\$ 190,000,000**

3.6 Overall Profitability

Cash flow sheets for both process alternatives can be seen in Tables 9 and 10 and indicate the overall process profitability of each. On the basis of a 20-year projected operating life, the base Sn-Beta process alternative has an estimated net present value of NPV@20%=\$29 million ± 40% while the Sn-Beta + $CaSO_4$ process alternative is projected to be valued at an NPV@20% of \$250 million ± 40% at a January 2019 basis date. An NPV@20% greater than \$0 indicates that the process exceeds the minimum discounted rate of return of 20%.

The primary difference between the two alternatives is the amount of revenue generated annually. The Sn-Beta + $CaSO_4$ alternative is estimated to produce \$60 million/yr more in revenue than the base Sn-Beta alternative. This difference is directly related to the efficiency of the Sn-Beta catalytic reaction, where the Sn-Beta + $CaSO_4$ converts almost twice as much carbon into saleable products than the base Sn-Beta (72% to 37%). Additionally, the majority of the revenue in the base Sn-Beta process is sourced from levulinic acid, which does not have a large market demand.

Table 27. Economic Cash Flow Sheet for the Base Sn-Beta Process Alternative ($ millions).
Location: Southeast North Dakota
Basis Date: January 2019

Year	Revenues	Operating Costs	Gross Profit	Depreciation	Federal Taxable Profit	Federal Income Tax	State Taxable Profit	State Income Tax	Nontaxable Charges	Net Profit	Present Value @ 20%
-2									($42)	($42)	($60)
-1									($42)	($42)	($50)
0									($60)	($60)	($60)
1	$130	($79)	$52	$16	$36	($8)	$29	($1.2)		$43	$36
2	$130	($79)	$52	$13	$39	($8)	$31	($1.3)		$43	$30
3	$130	($79)	$52	$12	$41	($9)	$32	($1.4)		$42	$24
4	$130	($79)	$52	$10	$42	($9)	$33	($1.4)		$42	$20
5	$130	($80)	$51	$9.0	$42	($9)	$33	($1.4)		$41	$16
6	$130	($79)	$52	$7.9	$44	($9)	$35	($1.5)		$41	$14
7	$130	($79)	$52	$7.0	$45	($10)	$36	($1.5)		$41	$11
8	$130	($79)	$52	$6.2	$46	($10)	$36	($1.6)		$41	$10
9	$130	($80)	$51	$5.4	$45	($10)	$36	($1.5)		$40	$8.0
10	$130	($79)	$52	$5.1	$47	($10)	$37	($1.6)		$41	$6.6
11	$130	($79)	$52	$5.1	$47	($10)	$37	($1.6)		$41	$5.5
12	$130	($79)	$52	$5.1	$47	($10)	$37	($1.6)		$41	$4.6
13	$130	($80)	$51	$5.1	$46	($10)	$36	($1.6)		$40	$3.7
14	$130	($79)	$52	$5.1	$47	($10)	$37	($1.6)		$41	$3.2
15	$130	($79)	$52	$5.1	$47	($10)	$37	($1.6)		$41	$2.6
16	$130	($79)	$52	$5.1	$47	($10)	$37	($1.6)		$41	$2.2
17	$130	($80)	$51	$5.1	$46	($10)	$36	($1.6)		$40	$1.8
18	$130	($79)	$52		$52	($11)	$41	($1.8)		$39	$1.5
19	$130	($79)	$52		$52	($11)	$41	($1.8)		$39	$1.2
20	$130	($79)	$52		$52	($11)	$41	($1.8)	$19	$58	$1.5

Note: Red numbers in parentheses represent negative values

NPV@20% $29

DCFROR 23%

Table 28. Economic Cash Flow Sheet for the Sn-Beta + CaSO4 Process Alternative ($ millions).

Location: Southeast North Dakota

Basis Date: January 2019

Year	Revenues	Operating Costs	Gross Profit	Depreciation	Federal Taxable Profit	Federal Income Tax	State Taxable Profit	State Income Tax	Nontaxable Charges	Net Profit	Present Value @ 20%
-2									($44)	($44)	($60)
-1									($44)	($44)	($50)
0									($70)	($70)	($70)
1	$190	($80)	$110	$16	$100	($21)	$77	($3.3)		$90	$75
2	$190	($80)	$110	$13	$100	($21)	$79	($3.4)		$89	$62
3	$190	($80)	$110	$12	$100	($21)	$81	($3.5)		$89	$52
4	$190	($80)	$110	$10	$100	($22)	$82	($3.5)		$89	$43
5	$190	($80)	$110	$9.3	$100	($22)	$82	($3.5)		$87	$35
6	$190	($80)	$110	$8.2	$110	($22)	$80	($3.6)		$88	$30
7	$190	($80)	$110	$7.2	$110	($22)	$80	($3.6)		$88	$25
8	$190	($80)	$110	$6.4	$110	($23)	$90	($3.7)		$88	$20
9	$190	($80)	$110	$5.6	$110	($22)	$80	($3.6)		$87	$17
10	$190	($80)	$110	$5.3	$110	($23)	$90	($3.7)		$87	$14
11	$190	($80)	$110	$5.3	$110	($23)	$90	($3.7)		$87	$12
12	$190	($80)	$110	$5.3	$110	($23)	$90	($3.7)		$87	$10
13	$190	($80)	$110	$5.3	$110	($23)	$80	($3.7)		$86	$8.1
14	$190	($80)	$110	$5.3	$110	($23)	$90	($3.7)		$87	$6.8
15	$190	($80)	$110	$5.3	$110	($23)	$90	($3.7)		$87	$5.7
16	$190	($80)	$110	$5.3	$110	($23)	$90	($3.7)		$87	$4.7
17	$190	($80)	$110	$5.3	$110	($23)	$80	($3.7)		$86	$3.9
18	$190	($80)	$110		$110	($24)	$90	($3.9)		$86	$3.2
19	$190	($80)	$110		$110	($24)	$90	($3.9)		$86	$2.7
20	$190	($80)	$110		$110	($24)	$90	($3.9)	$20	$110	$2.8

Note: Red numbers in parentheses represent negative values

NPV@20% $250

DCFROR 41%

4. Summary and Conclusions

The objective of this study was to determine the feasibility of catalytically transforming corn stover into lactic acid or both lactic and levulinic acids as primary products. Two separate process alternatives were examined. The first evaluated the conversion of extracted sugars to carboxylic acids solely through the use of Sn-Beta. This resulted in the co-production of lactic and levulinic acid as major products. The second process included calcium sulfate as part of the reaction mixture, significantly increasing the reaction selectivity towards lactic acid. Both process alternatives produced other valuable by-products that are worth purifying to a sellable grade such as formic and acetic acid, calcium sulfate, and furfural. Process plant throughput was determined by feeding the corn stover equivalent to corn kernel feed in a typical bioethanol plant. This set the input of both processes at 250,000 tons of corn stover per year.

Process separation steps utilized reactive extraction to remove carboxylic acids from the reaction solutions. To define optimal conditions for co-extraction of lactic, levulinic, acetic, and formic acids, both screening and optimization experiments were conducted. Based on the screening experiments, it was determined that the most significant factors are the aqueous/organic phase ratio, extractant concentration, acid composition and diluent type. Optimal conditions for both examined diluents, 1-octanol and isoamyl, using a central composite design. Based on these conditions, it was determined that extraction with a 1-octanol diluent was more economical based on the simpler downstream separation requirements.

It was found that both processes were profitable with the base Sn-Beta design being valued at an NPV@20% of $100 million while the Sn-Beta + $CaSO_4$ design reached an NPV@20% of $340 million. The lower profitability of the base case is due to the lower carbon conversion of carbohydrates to saleable carboxylic acid products. For the base case to be more profitable, levulinic acid would need to be sold at a price $8.65/kg higher than that of lactic acid.

A major hurdle for both processes is the demand for levulinic acid. Though global demand for levulinic acid is expected to increase 150 to 200-fold over the next 6 years, the global demand in 2020 is expected to be 4,100 metric tons[108, 111]. The base Sn-Beta alternative is expected to annually produce four times this demand and the Sn-Beta + $CaSO_4$ alternative will slightly exceed it. For demand to meet production, advances in levulinic acid applications are needed. One potential application is to convert levulinic acid to 1,4-pentanediol, which can be used as a monomer biodegradable high-strength polyester production [30].

Chapter V – Conclusions

Study Objectives

Chapter II

1. Advance the conversion efficiency of the reaction for both glucose and xylose through careful study of the effect of residence time, reaction temperature, and substrate sources on catalyst selectivity.
2. Evaluate the optimum reaction conditions for monomeric sugar mixtures that simulate the typical composition of sugars extracted from 1) microalgae and 2) corn stover.
3. Apply the optimum reaction conditions to actual sugar mixtures derived from microalgae and corn stover to evaluate effectiveness.

Chapter III

1. Screen the significance of organic/aqueous phase ratio, extractant concentration, extraction time, diluent, carboxylic acid concentration, and temperature for the reactive extraction of acetic, formic, lactic, and levulinic acid using triocylamine.
2. Determine the optimum operating conditions for reactive extraction using 1-octanol and isoamyl alcohol as active diluents.
3. Construct a simple process design to compare the economic cost of an 1-octanol and a isoamyl alcohol reactive extraction system.

Chapter IV

1. Determine the technological and economic feasibility of two processes designed from Sn-Beta reactions yields. The first utilizes just Sn-Beta to primarily produce

lactic and levulinic acid, while the second uses the co-catalysis of Sn-Beta and CaSO4 to almost exclusively produce lactic acid.

Conclusions
Chapter II

1. The optimal Sn-Beta conversion efficiency of xylose was found to occur at 180˚ C. This is lower than the previously found optimum for glucose conversion of 200˚ C, indicating a higher rate of xylose degradation reactions at higher temperatures. When mixed in a model corn stover extract solution (65 wt% glucose, 35 wt% xylose), optimal lactic acid yield occurred at 200˚ C.
2. For both glucose and xylose, minor changes in reaction yield occurred between 1-5 hours of run time, indicating that most conversion occurs within the first hour.
3. Microalgae extract yielded 8.2 carbon wt% lactic acid and 0.8 carbon wt% levulinic acid while corn stover extract had yields of 60 and 1.6 carbon wt%, respectively when conversion was performed with the Sn-Beta catalyst in the presence of $CaSO_4$. Both extracts experienced significantly lower yields of levulinic acid due to the inclusion of calcium sulfate in the reaction mixture. It is expected that the significant difference in lactic acid yields between the two extracts is a result of an unidentified catalytic inhibitor.
4. Further examination of the effect of calcium sulfate on the Sn-Beta reaction yield revealed that sulfate neutralizes the Brønsted acid sites on Sn-Beta, inhibiting the dehydration reaction to levulinic acid. This decrease in turn led to increased yield of the retro-aldol reaction, which forms lactic acid. Additionally, it was found

that calcium acts as a second Lewis acid, aiding in the retro-aldol reaction efficiency.

Chapter III

1. The most significant factors for co-reactive extraction of acetic, formic, lactic, and levulinic using TOA were, in order: diluent type, acid composition, and organic/aqueous phase ratio.
2. Co-extraction of the target carboxylic acid with isoamyl alcohol as a diluent was optimized at a lactic/levulinic acid ratio (acid composition) of 1.31, organic/aqueous phase ratio of 0.45, and extractant (TOA) concentration of 13 wt% in the organic phase.
3. Co-extraction of the target carboxylic acid with 1-octanol as a diluent was optimized at a lactic/levulinic ratio of 0.69, organic/aqueous phase ratio of 0.65, and extractant (TOA) concentration of 22 wt% in the organic phase.
4. A process designed to extract carboxylic acids produced from Sn-Beta using 1-octanol as a diluent is expected to save $7.1 million over a 20-year life cycle compared to a process using isoamyl alcohol as a diluent. This is due to formic acid forming an azeotrope with the reactive extraction diluent (isoamyl alcohol) and the back extraction diluent (water), resulting in an extra extractive distillation step to purify formic acid.

Chapter IV

1. The base Sn-Beta alternative, designed to more selectively produce levulinic acid, is estimated to be worth an NPV@20% of $29 million over a 20 year

project life cycle. Including $CaSO_4$ in the Sn-Beta reaction to more almost exclusively produce lactic acid increased the process NPV@20% to $250 million.

2. A major hurdle for both processes is the demand of levulinic acid. It is expect that in 2020 the global levulinic acid demand will reach 4,100 metric tons, which is roughly equivalent to the amount of levulinic formed in one commercial process facility using the Sn-Beta + $CaSO_4$ reaction alternative and only a quarter of that produced in a commercial facility using the base Sn-Beta case.

Recommendations

Chapter II:

Three recommendations for future work arise from the Sn-Beta reaction study. The first is that a more robust residence time study should be conducted. Glucose and xylose residence time studies investigated in Chapter II showed only a small variance between one and five hours, indicating that the majority of the reaction occurs under one hour. Reaction times under one hour are difficult to examine due to the batch reactor configuration. Slow temperature ramp times for the reactor result in one hour to heat and half an hour to cool, providing a dynamic temperature period that is longer than the desired reaction time. This makes it impossible to discern how much of the reaction occurred in the heat up time versus the set reaction period. These problems can be mitigated in a continuous flow reactor where the sugar solution is introduced to the catalyst at the reaction temperature (200˚ C), allowing for better control of the residence time.

A second recommendation is to study the catalyst deactivation in a flow reactor. Previous studies have shown that Sn-Beta loses 90% of its activity within 30 hours [112]. The addition of calcium sulfate to the reaction mixture has shown decreased sugar degradation reactions, resulting in higher lactic acid yields. These degradation reactions, which produce hydrochar, are likely the main source of deactivate reducing the ability of the reactants to adsorb onto the active sites. A study on this deactivation will provide more insight on how often the catalyst will need to be regenerated.

The final recommendation for Chapter II is to identify the side-products produced in microalgae extract solutions. Based on complimentary experiments, it is evident that the addition of calcium sulfate to the reaction mixture nearly eliminates dehydration reactions while simultaneously increasing the yield of lactic acid. However, the increased lactic acid yield was not observed in the microalgae experiments, while two unknown products which represent about 30 carbon wt% appeared. Identification of these products will help to provide insight on what is reducing the lactic acid yield. Primary identifications analyzed the complete reaction solution on gas chromatography-mass spectrometry (GC-MS) and electrospray ionization time of flight mass spectrometry (EIS-TOF-MS) but were inconclusive. Further identification should focus on isolating the two products using HPLC, then using mass spectrometry techniques for identification.

Chapter III:

Two recommendations for the reactive extraction of carboxylic acids with TOA can be made. The first is that when formic or acetic acid are present, 1-octanol should be used as the active diluent to aid TOA in extraction. 1-Octanol allows easier separation of formic and acetic acid from TOA than isoamyl alcohol. Conversely, if these light acids are not present isoamyl alcohol provides higher reactive extraction efficiency for carboxylic acids and should be used for lactic and levulinic acid extraction.

The second recommendation is to conduct further research on the back extraction with trimethylamine. Experiments should focus on finding optimum conditions for extraction of a representative sample of carboxylic acids in a 1-octanol/TOA solution. Emphasis should be placed on lactic and levulinic extraction since acetic and formic would be separated before this step.

Chapter IV:

The primary hurdle for both process alternatives is the global demand of levulinic acid. Therefore, further research should be focused on finding applications for levulinic acid. One application that will be studied at the University of North Dakota is to hydrogenate levulinic acid with 2-pentanol, forming two-equivalents of γ-valerolactone (GVL). GVL can then be further hydrogenated to 1,4-pentanediol, which can be used as a high-strength polyester.

Selective levulinic acid production has significant room for improvement for reaction conversion efficiency. In the base Sn-Beta process, only 37% of the inlet carbon from sugars is transformed into saleable carboxylic acid products. This is significantly lower than the 72% observed in the Sn-Beta + $CaSO_4$ alternative. Further research can be conducted to identify a similar catalyst modification that can emphasize the Brønsted acid sites and reduce sugar degradation to increase overall levulinic acid yield.

APPENDIX A - HPLC ANALYSIS

After each Sn-Beta catalytic reaction sample was produced, 10 mL of the sample was removed from the bulk and syringe filtered using a 0.2 micron filter (Agilent Technologies, Santa Clara USA). For most samples, 1 mL of the syringe filtered product was then placed in a 2 mL autosampler vial. Conversely, samples obtained from biomass extract, like corn stover and microalgae, were mixed at a ratio of 500 μL sample to 500 μL of the HPLC mobile phase (5 mM H_2SO_4) and syringe filtered for a second time. This allowed for potential contaminants that precipitate due to a change in pH to be removed before injection into the HPLC. Each autosampler vial was then capped using a crimping tool.

For quantification, eight calibration standards were created. Each standard contained the four target organic acids (lactic, levulinic, acetic, and formic) and the two most commonly analysed sugars (glucose and xylose). Also present in the solution was glycolic acid which often appeared at low levels from reaction samples. A separate calibration standard was used to analyse microalgae extract samples since primary component galactose eluted at the same time as xylose. The initial calibration solution had each target analyte present at a concentration of 1.25 mg/mL. This solution was labelled calibration standard A. Standard B was created by taking 2 mL of standard A and diluting it with 2 mL of 5 mM H_2SO_4, halving the concentration. This dilution was done a total of seven times to create standards B-H.

HPLC separation occurred using a hydrogen ion-exchange column, Hi-Plex-H (Agilent Technologies, Santa Clara USA) and a mobile phase of 5 mM H_2SO_4. Mobile phase was pumped through the column at 0.6 mL/min at 40°C, reaching a maximum column pressure of 5,100 kPa. Once separated, analytes were detected on a refractive index detector for up to 45 minutes after injection. Before each analysis, the reference cell for the refractive index detector was purged for 10 minutes to ensure there was no interference with the signal.

Each analysis set started with injection of two mobile phase blanks. This was done to ensure that a stable baseline was established and all contaminants from previous runs were eliminated. After blank runs were conducted, the eight calibration standards were subsequently analysed. If a new analysis set was conducted within a week of a previous analysis, only one calibration standard was run to verify a consistent calibration curve. After HPLC calibration, samples were immediately analysed. Each sample was injected for three separate runs on the HPLC to ensure limited error due to injection volume. For large analysis sets, a baseline check and calibration check tool place every 24 hours to ensure there was no measurement drift. This was done by injecting a blank solution followed by a single calibration standard and an additional blank. An example HPLC analysis sequence can be seen in Table 29.

Table 29. Example HPLC analysis sequence

Vial	Sample Name	Method	Injection/ Vial	Sample Amount (µL)	Injection Volume (µL)
1	01_Blank	40C_0.6ML_45MIN	1	1000	15
1	02_Blank	40C_0.6ML_45MIN	1	1000	15
91	03_AK_Cal A	40C_0.6ML_45MIN	1	1000	15
92	04_AK_Cal B	40C_0.6ML_45MIN	1	1000	15
93	05_AK_Cal C	40C_0.6ML_45MIN	1	1000	15
94	06_AK_Cal D	40C_0.6ML_45MIN	1	1000	15
95	07_AK_Cal E	40C_0.6ML_45MIN	1	1000	15
96	08_AK_Cal F	40C_0.6ML_45MIN	1	1000	15
97	09_AK_Cal G	40C_0.6ML_45MIN	1	1000	15
98	10_AK_Cal H	40C_0.6ML_45MIN	1	1000	15
41	11_AK485-02	40C_0.6ML_45MIN	3	1000	15
42	12_AK487-02	40C_0.6ML_45MIN	3	1000	15
43	13_AK489-02	40C_0.6ML_45MIN	3	1000	15
44	14_AK491-02	40C_0.6ML_45MIN	3	1000	15
45	15_AK493-02	40C_0.6ML_45MIN	3	1000	15
46	16_AK495-02	40C_0.6ML_45MIN	3	1000	15
1	17_Blank	40C_0.6ML_45MIN	1	1000	15
96	19_AK_Cal F	40C_0.6ML_45MIN	1	1000	15
1	20_Blank	40C_0.6ML_45MIN	1	1000	15
47	21_AK303-02	40C_0.6ML_45MIN	3	1000	15
48	22_AK311-02	40C_0.6ML_45MIN	3	1000	15
49	23_AK377-02	40C_0.6ML_45MIN	3	1000	15
50	24_AK379-02	40C_0.6ML_45MIN	3	1000	15
51	25_AK383-02	40C_0.6ML_45MIN	3	1000	15
52	26_AK497-02	40C_0.6ML_45MIN	3	1000	15
53	27_AK387-02	40C_0.6ML_45MIN	3	1000	15

APPENDIX B - DATA PROCESSING

Quantification of analytes in each sample was based off of areas determined using Agilent ChemStation software. Areas were determined based on the increase in signal relative to the mobile phase baseline. Since sulphuric acid is not present at normal levels in the sample, a drop in the baseline signal occurs at 8.55 minutes. To clean up the chromatograph, a blank run with this same signal drop was subtracted from the analysed chromatograph. This allowed for a better identification of the baseline by the ChemStation software. Figure 18 shows the raw chromatograph with the inverted peak and the process chromatograph for a carboxylic acid sample produced from Sn-Beta. Peaks were identified based on the retention time of pure analyte standard solutions. The typical retention time for each analyte can be seen on Table 30.

Peak areas were translated into concentrations based on calibration standards. Eight samples with known concentrations of each sample was analysed. A linear regression was than conducted to create a line of best fit to relate peak area to concentration. An example of this can be seen in Figure 19. To ensure an accurate correlation, an R^2 value greater than 0.999 was required to estimate the analyte concentration. Additionally, in order for analyte to be identified or quantified, their signal had to exceed an acceptable limit. This was based on the standard error of the equation's predicted area compared to its actual area [SE(y)]. In order for and analyte to be identified, its measured area had to be three times larger than its SE(y). Similarly, for the analyte to be quantified, it needed to be five times larger than its SE(y).

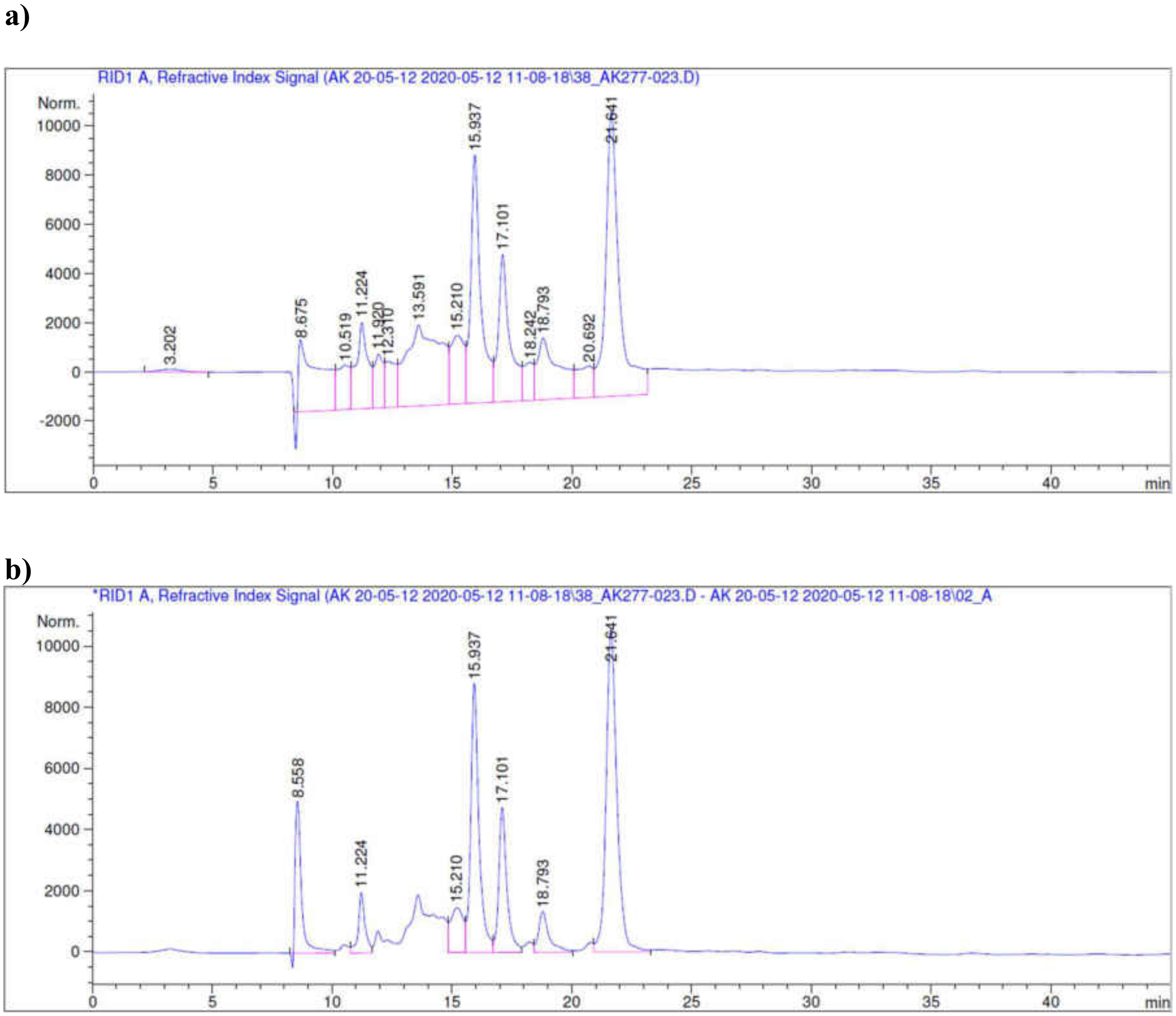

Figure 18. Example of unprocessed (a) and processed (b) HPLC chromatographs from laboratory samples

Table 30. Typical retention times for target analytes using Hi-Plex-H column.

Analyte	**Retention Time (min.)**
Glucose	11.53
Xylose	12.23
Glycolic Acid	15.33
Lactic Acid	15.94
Formic Acid	17.10
Acetic Acid	18.79
Levulinic Acid	21.44

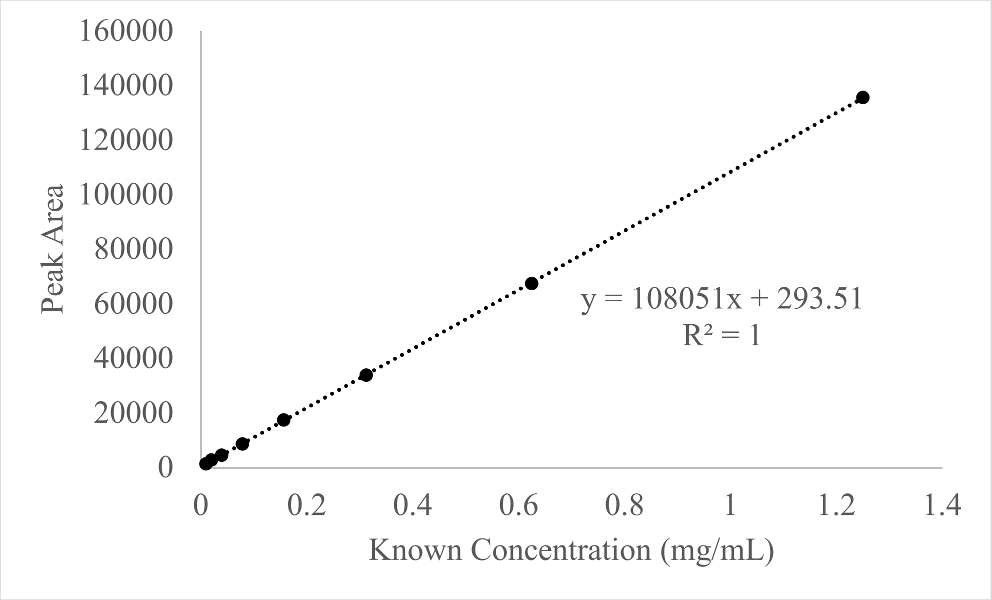

Figure 19. Example of a lactic acid calibration curve relating peak area to known concentration.

For Sn-Beta reactions, product yields were calculated based on the amount of carbon in the inlet sugar that ended up in final products. This was done by first the calculating the moles of the acid produced and dividing it by the moles of inlet sugars. The resulting mole fraction was than multiplied by the carbon ratio to yield the carbon yield. This carbon ratio was based on the number of carbon molecules that comprised the product divided by the number in reactant. For example, lactic acid (3 carbon) produced from glucose (6 carbon) has a carbon ratio of ½. The carbon yield equation can be seen below:

$$Carbon\ yield = \frac{Moles\ of\ Product}{Moles\ of\ Reactant} * \frac{Carbon\ \#\ of\ Product}{Carbon\ \#\ of\ Reactant}$$

Efficiency of reactive extraction experiments was measured by the distribution coefficient as defined as:

$$K_D = \frac{[\mathrm{HA}]_{\mathrm{org}}}{[\mathrm{HA}]_{\mathrm{aqu}}}$$

Where $[\mathrm{HA}]_{\mathrm{org}}$ is the concentration of the target acid in the organic phase produced during the extraction and $[\mathrm{HA}]_{\mathrm{aqu}}$ is the target acid concentration in the aqueous phase produced during the extraction. As final aqueous solutions analyze be HPLC, $[\mathrm{HA}]_{\mathrm{aqu}}$ was easily found. Estimation of $[\mathrm{HA}]_{\mathrm{org}}$ was accomplished by finding the weight loss of the carboxylic acid in the aqueous phase after extraction. Organic phase volume was estimated by assuming an ideal solution where the volume of the extractant and diluent are additive. The organic phase calculation can be summarized as below:

$$[HA]_{org} = \frac{V_{aqu.} * ([HA]_{aqu,initial} - [HA]_{aqu.,final})}{(\rho_{diluent} * m_{diluent}) + (\rho_{extractant} * m_{extractant})}$$

Where V_{aqu} is the volume of the aqueous phase, ρ is density, and m is mass.

APPENDIX C - ANOVA ANALYSIS FOR REACTIVE EXTRACTION EXPERIMENTS

A six factor, half factorial design was created and discussed in Chapter III to screen variables for the reactive extraction of carboxylic acids with trioctylamine. This experimental design allowed for identifications of the factors and interaction of factors that played the most significant role in extraction. Variable significance can be determined by calculating the standardized effect, which is depicted in the Pareto charts for each individual acid. The standardized effect is a t-statistic that determines if the null hypothesis (that the variable has no effect) can be rejected or not. In order for the null hypothesis to be rejected, the standardized effect must be greater than the set significance level, which is a function of the model's confidence level. Variables with the least significance can be eliminated from the model, increasing the precision by adding more degrees of freedom when calculating the model's error. One by one terms can be removed from the model until only statistically significant terms remain.

Sorted by acid, the following figures will first depict the unreduced model with the effect of each examined variables. This will be succeed by the reduced model were variables are eliminated until only significant ones remain. Graphs of the fitted means of the high and low values of main effects and two-way interactions, giving a visual representation of the variables effects. The final graphs are the normal probability plot and versus fit graphs. These graphs reaffirm the assumptions that the models are randomly distributed and have constant variance.

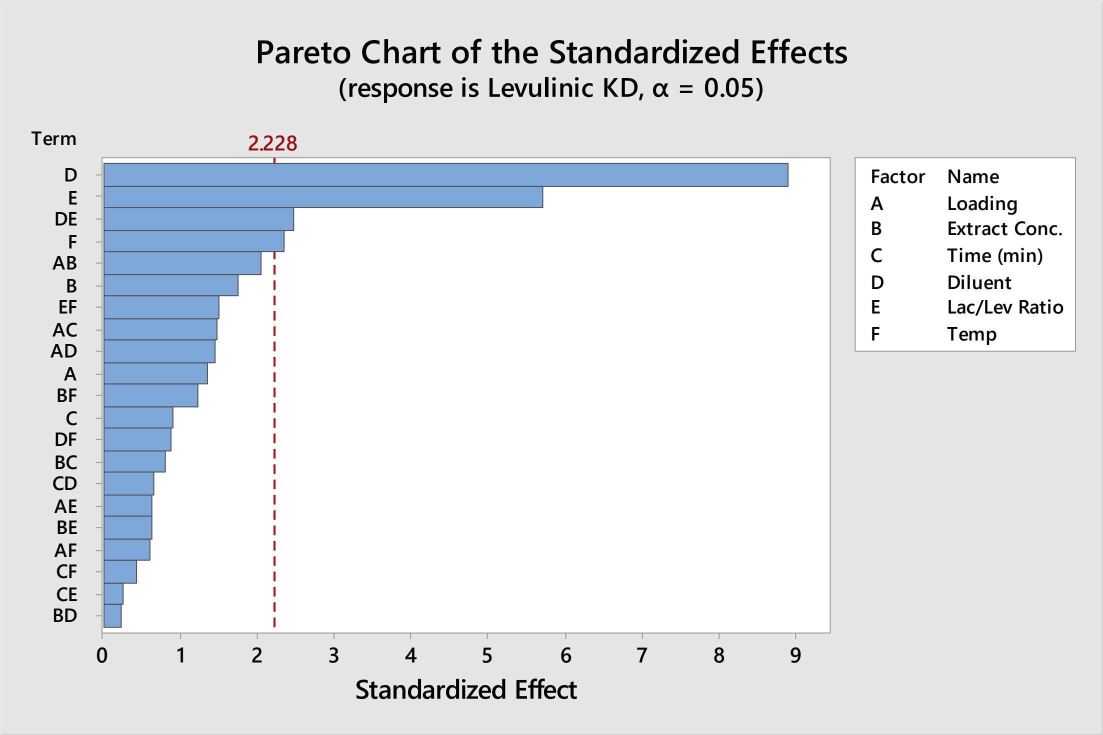

Figure 20. Unreduced model for the examined effects of levulinic acid reactive extraction.

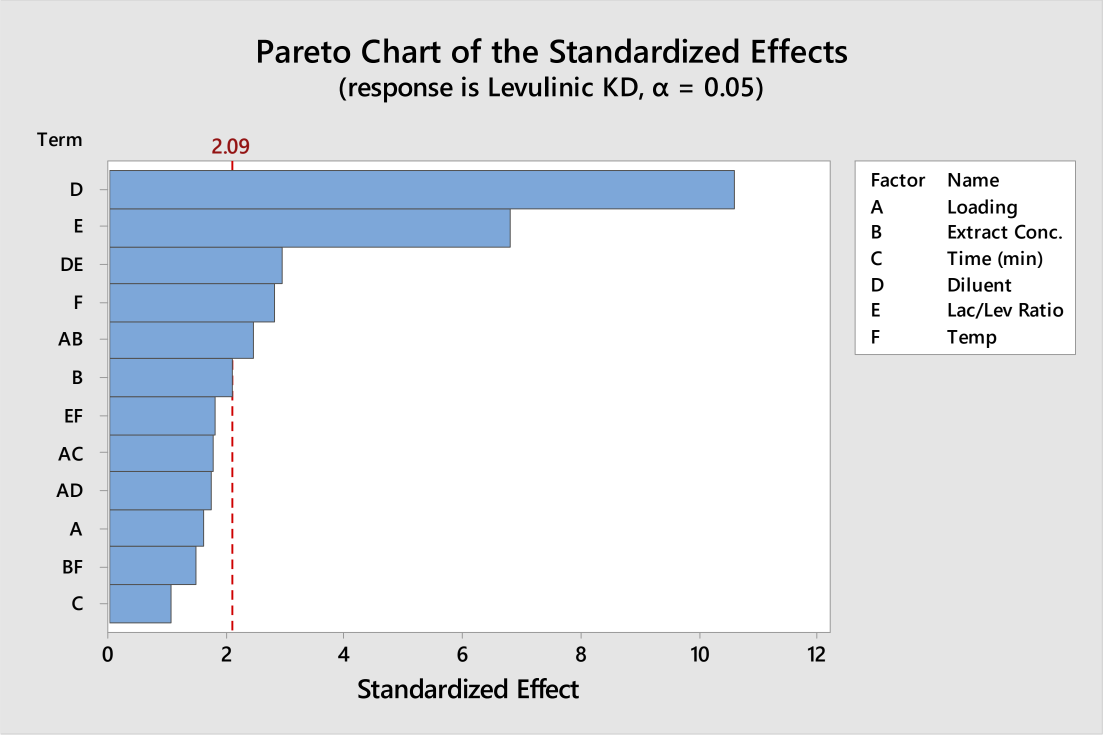

Figure 21. Reduced model for the examined effects of levulinic acid reactive extraction.

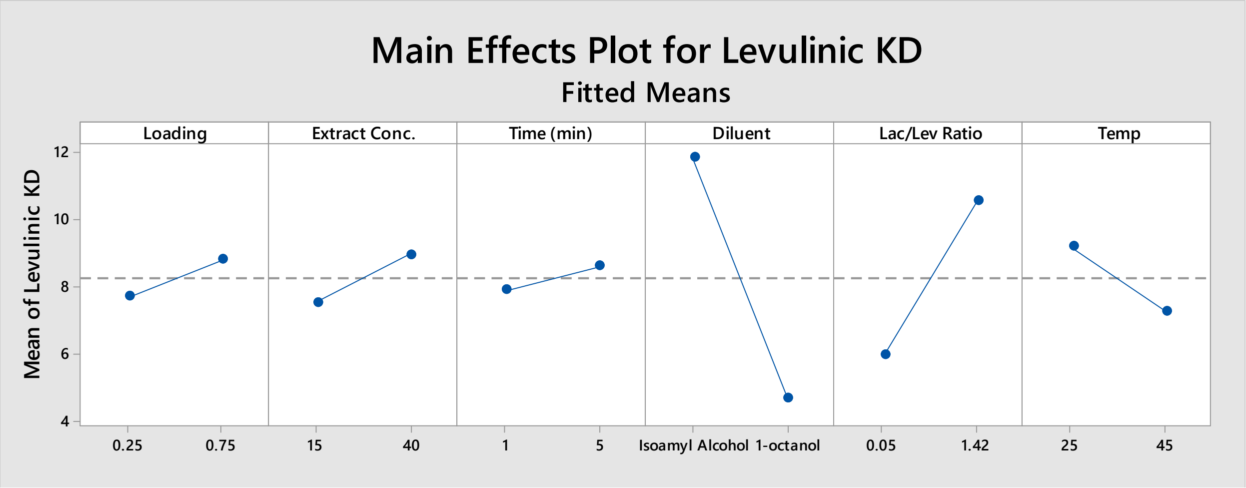

Figure 22. High and low means of significant main effects for levulinic acid reactive extraction.

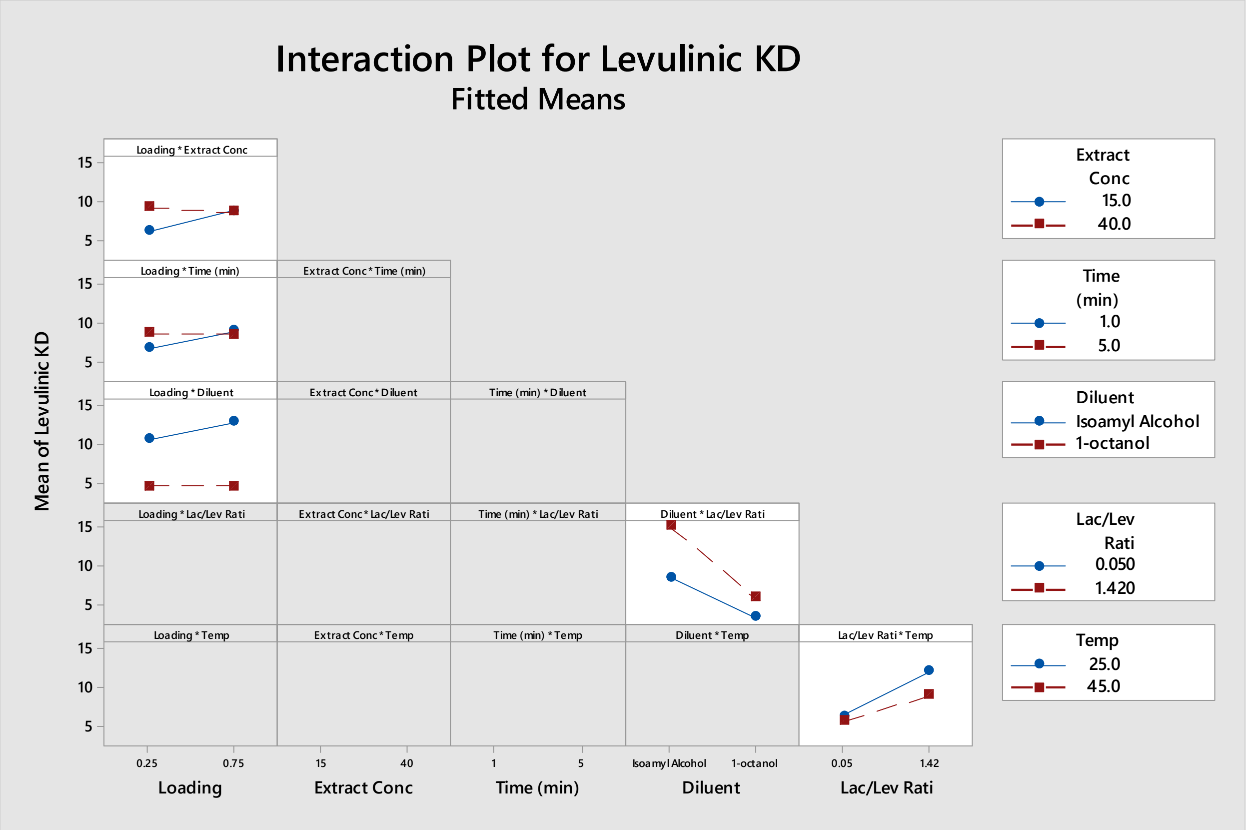

Figure 23. High and low means of significant two-interactions for levulinic acid reactive extraction.

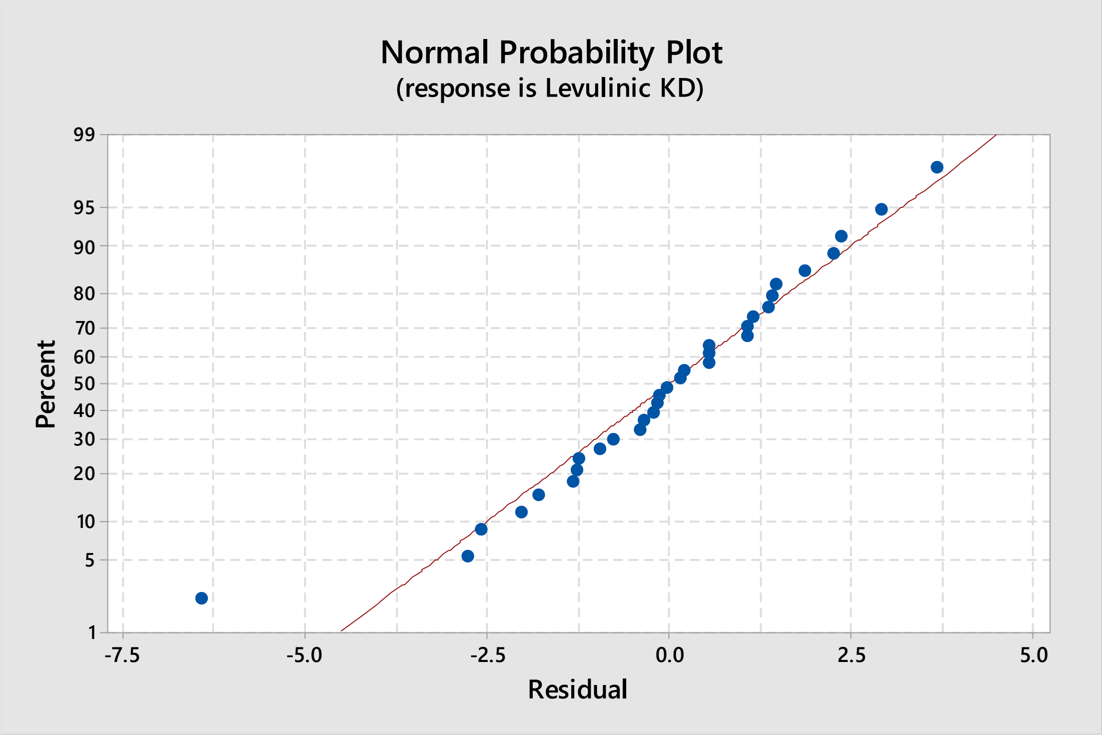

Figure 24. Normal probability distribution of levulinic acid extraction model.

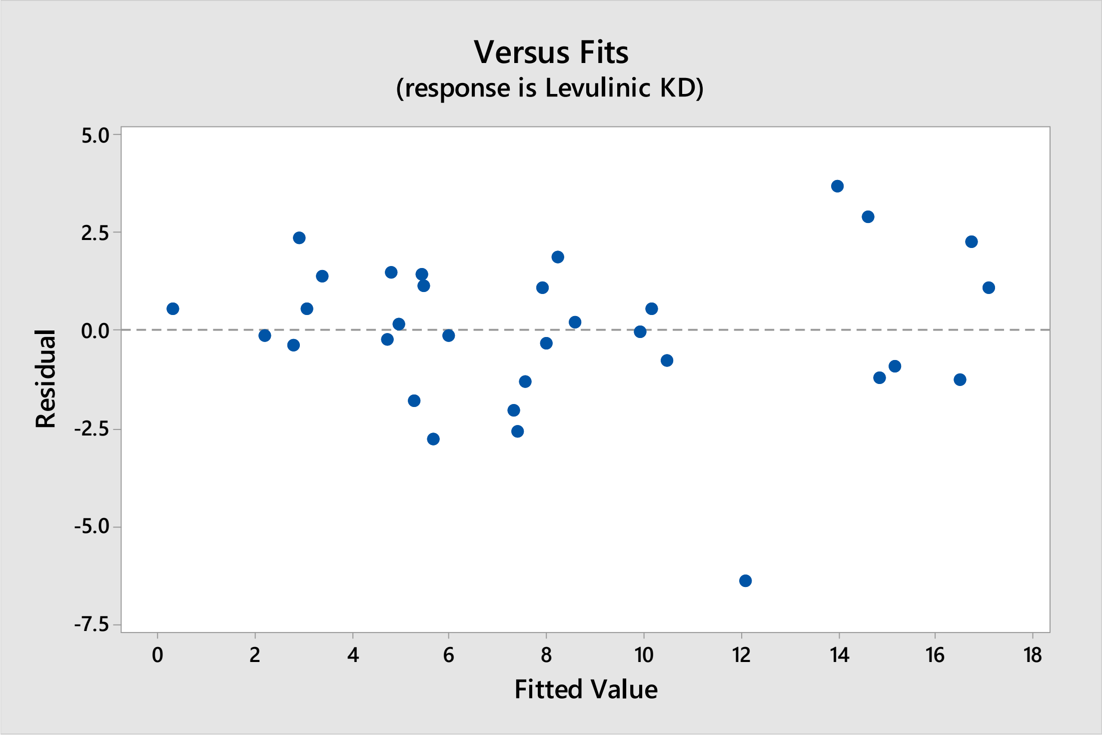

Figure 25. Versus fit plot of levulinic acid extraction model

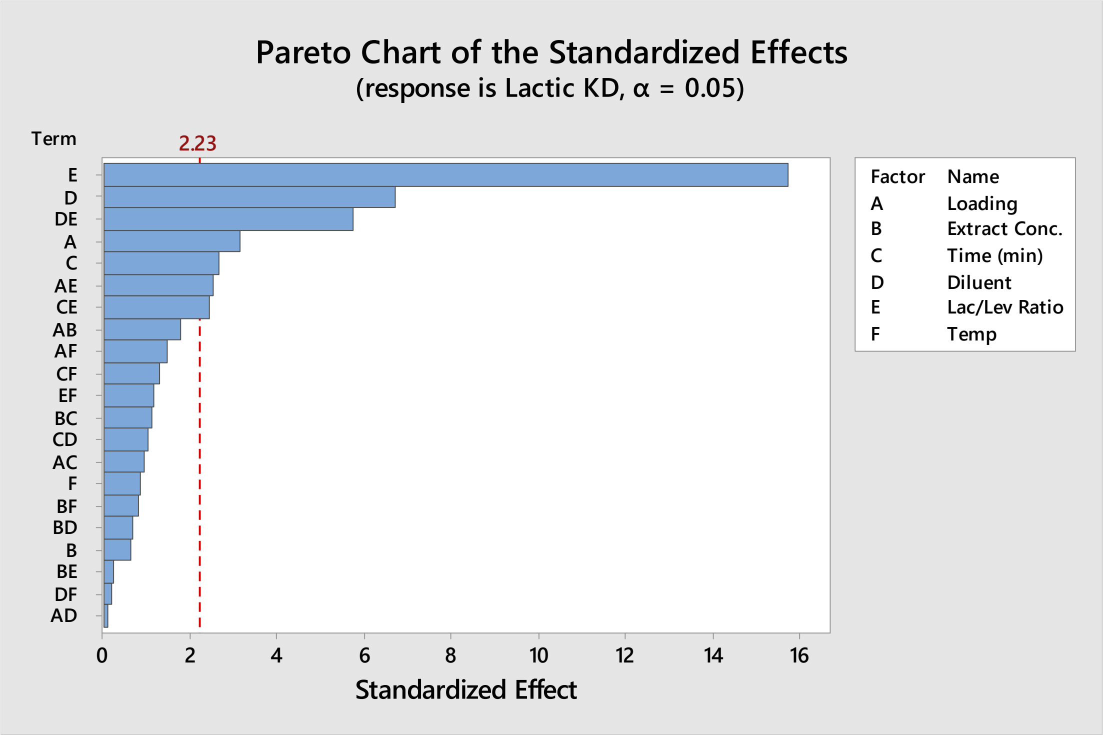

Figure 26. Unreduced model for the examined effects of lactic acid reactive extraction.

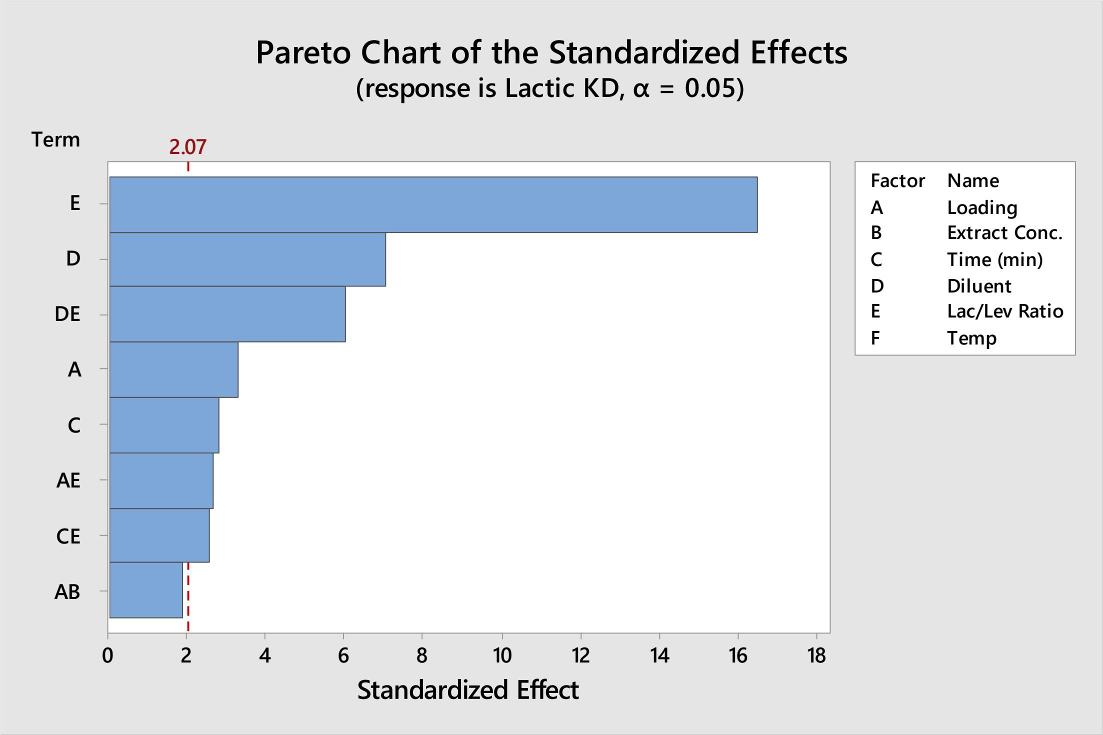

Figure 27. Reduced model for the examined effects of lactic acid reactive extraction.

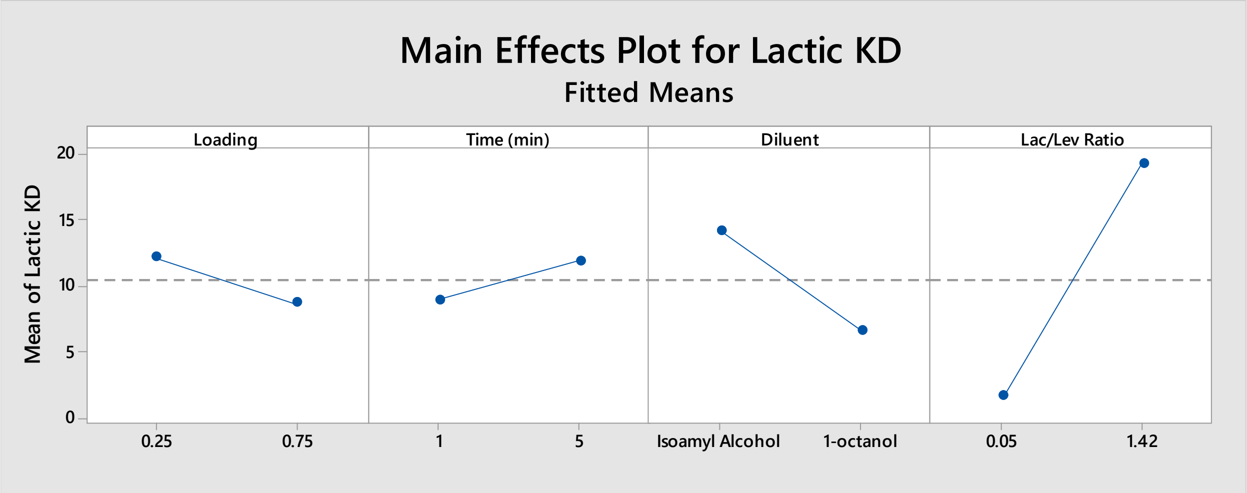

Figure 28. High and low means of significant main effects for lactic acid reactive extraction.

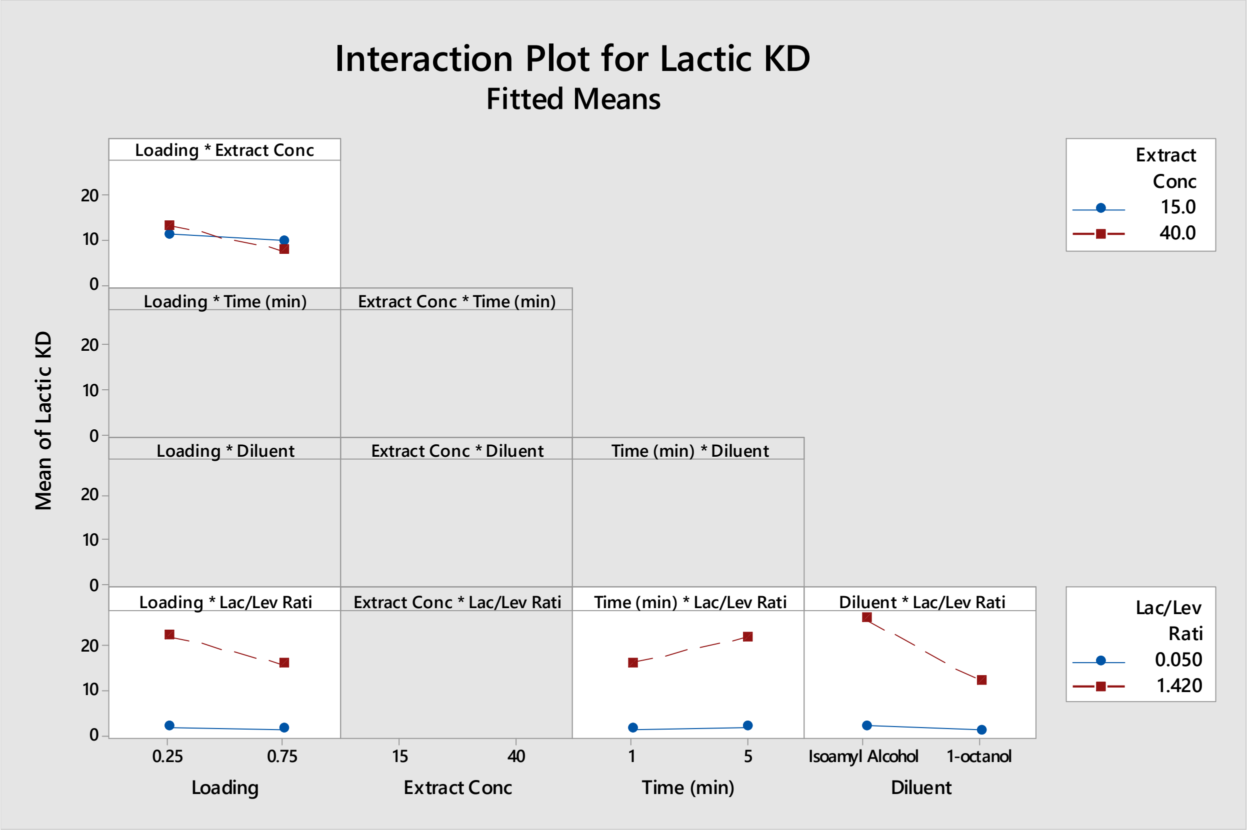

Figure 29. High and low means of significant two-interactions for lactic acid reactive extraction.

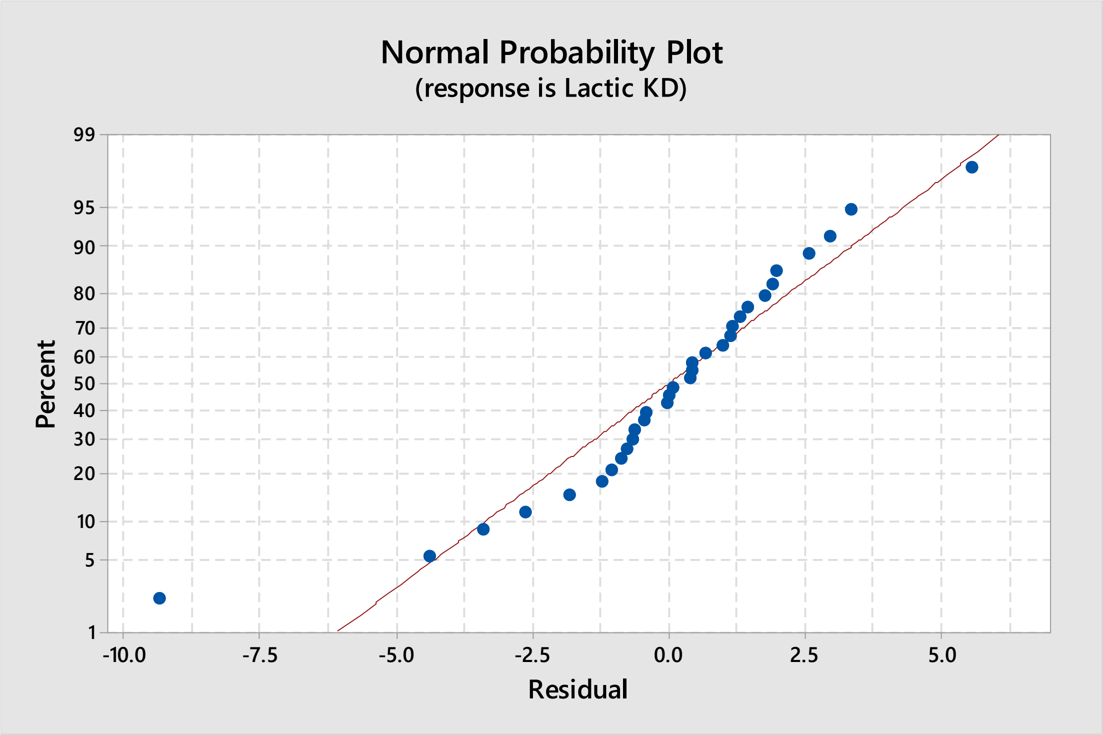

Figure 30. Normal probability distribution of lactic acid extraction model.

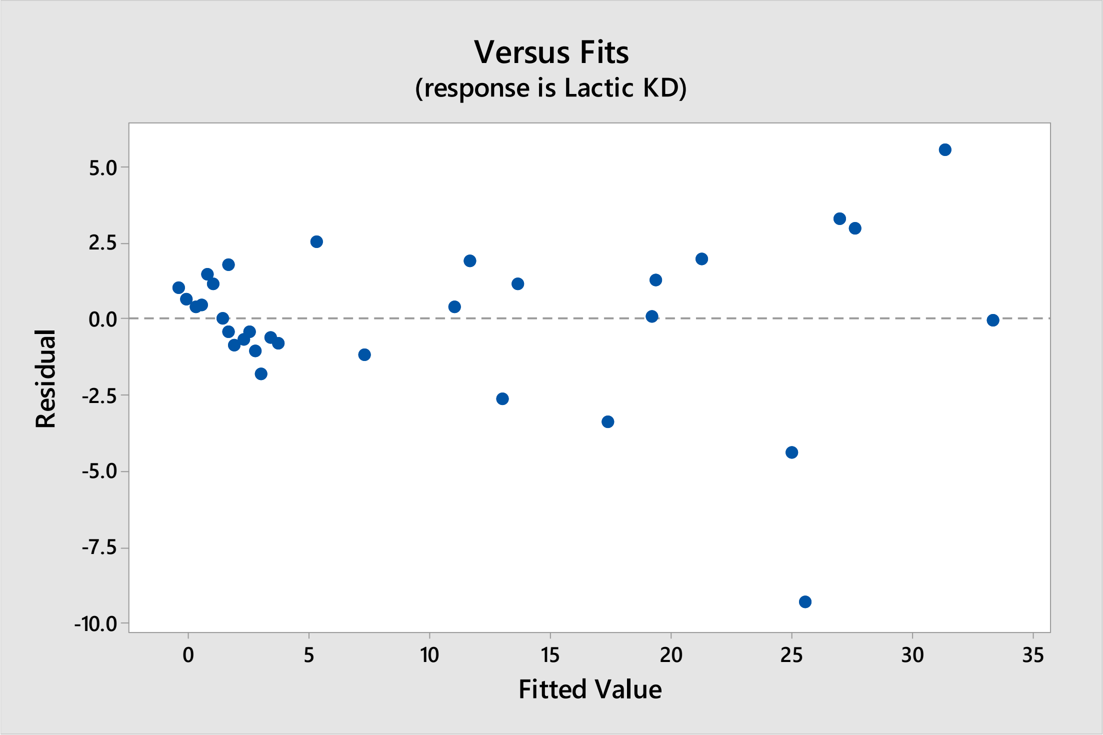

Figure 31. Versus fit plot of lactic acid extraction model

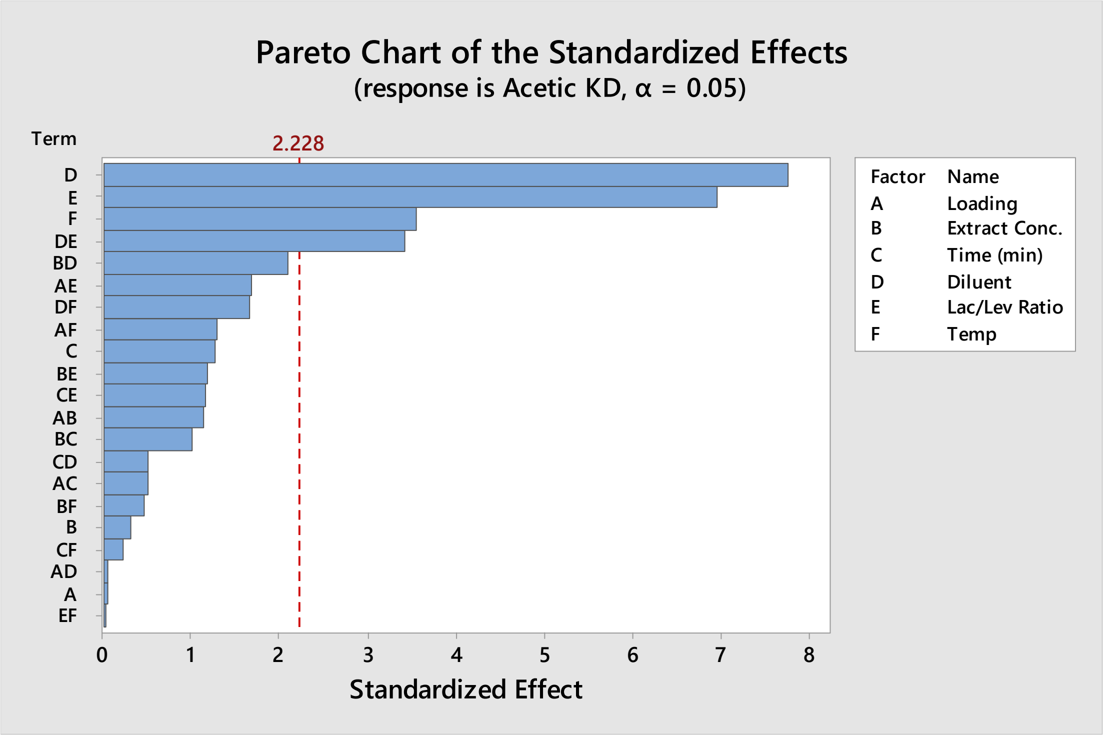

Figure 32. Unreduced model for the examined effects of acetic acid reactive extraction.

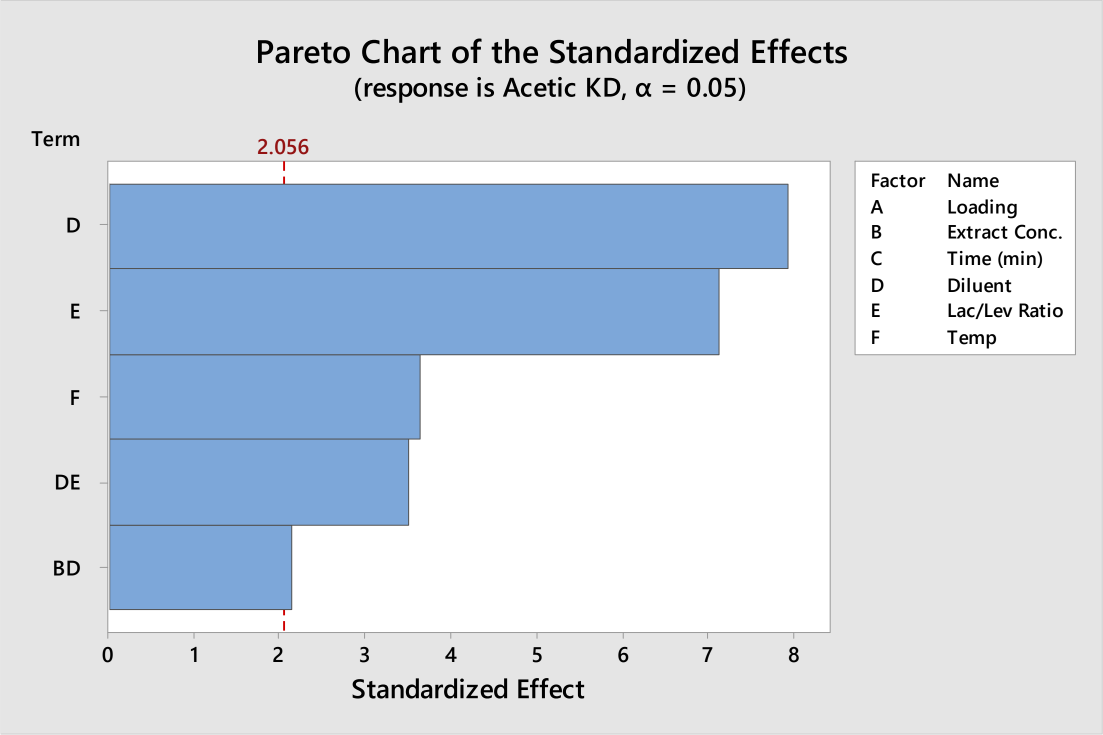

Figure 33. Reduced model for the examined effects of acetic acid reactive extraction

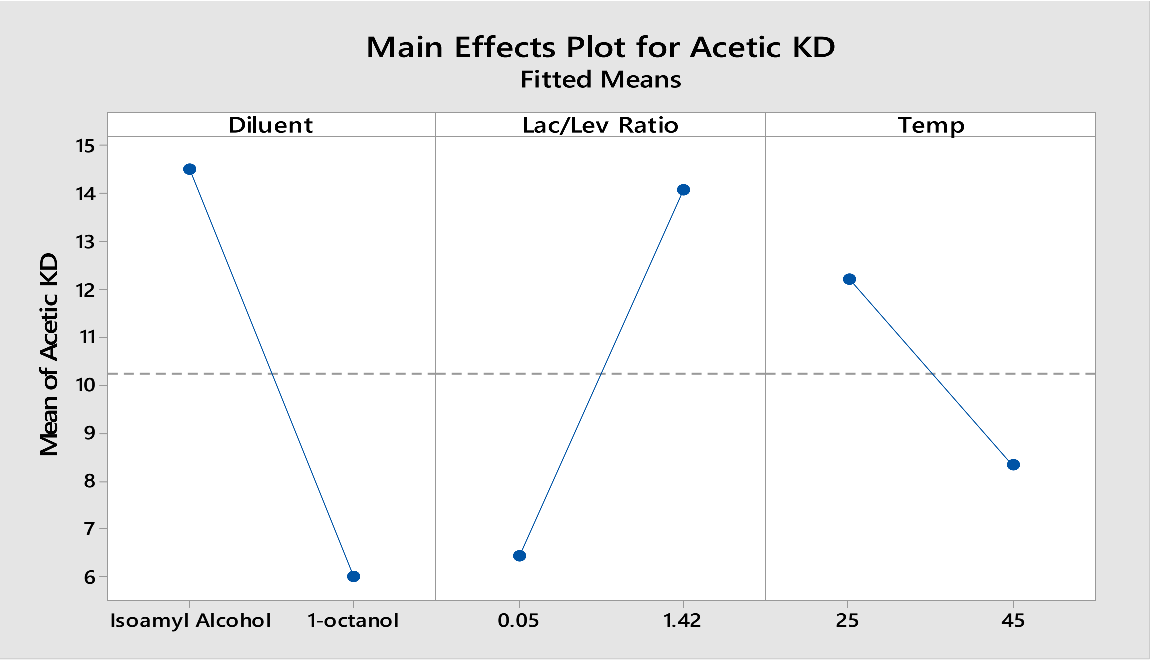

Figure 34. High and low means of significant main effects for acetic acid reactive extraction.

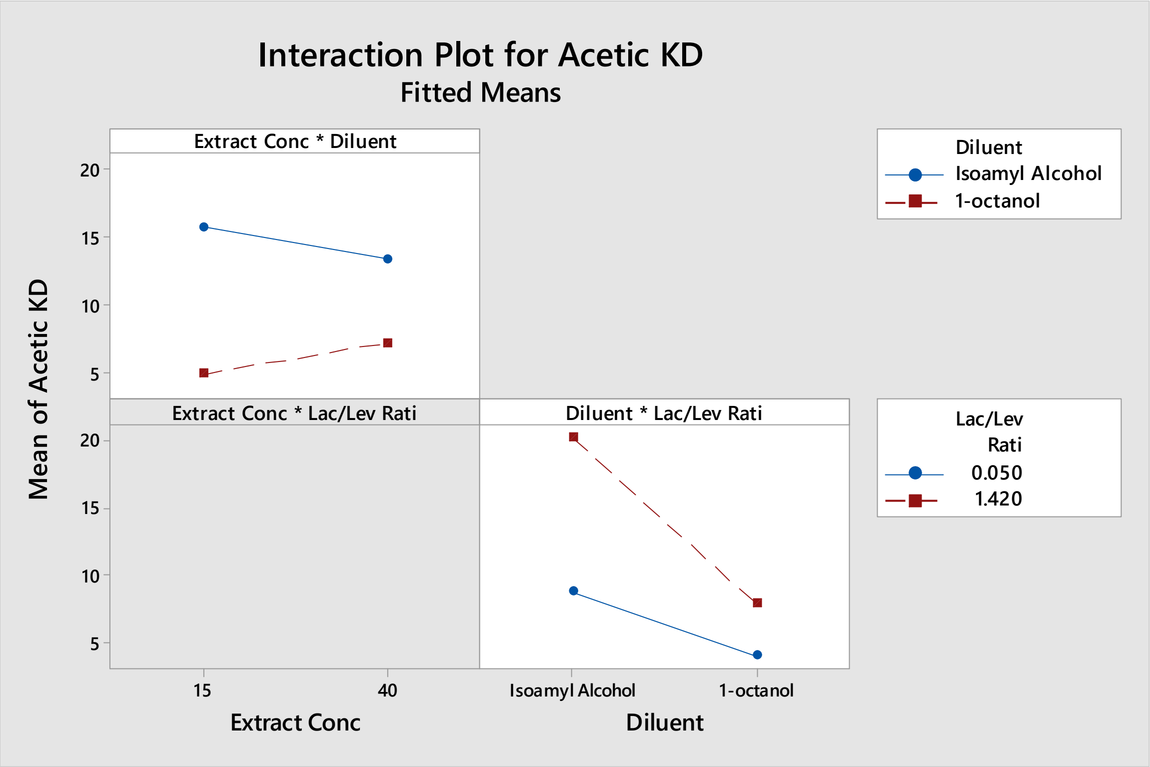

Figure 35. High and low means of significant two-interactions for acetic acid reactive extraction.

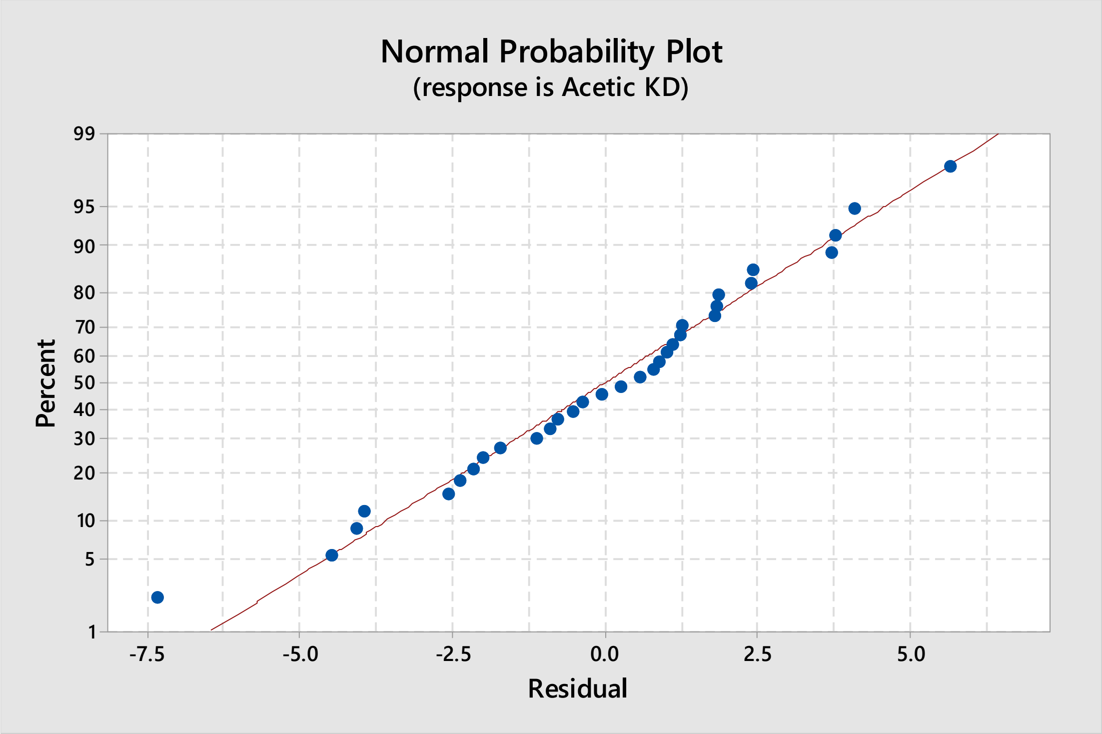

Figure 36. Normal probability distribution of acetic acid extraction model.

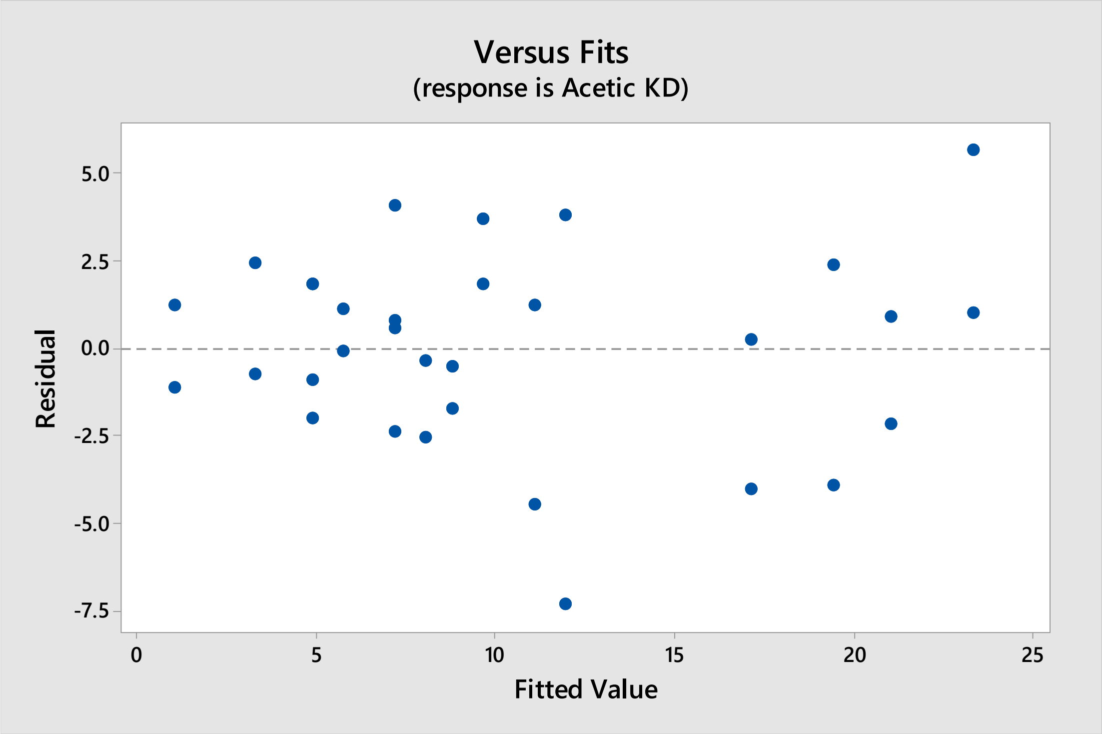

Figure 37. Versus fit plot of acetic acid extraction model

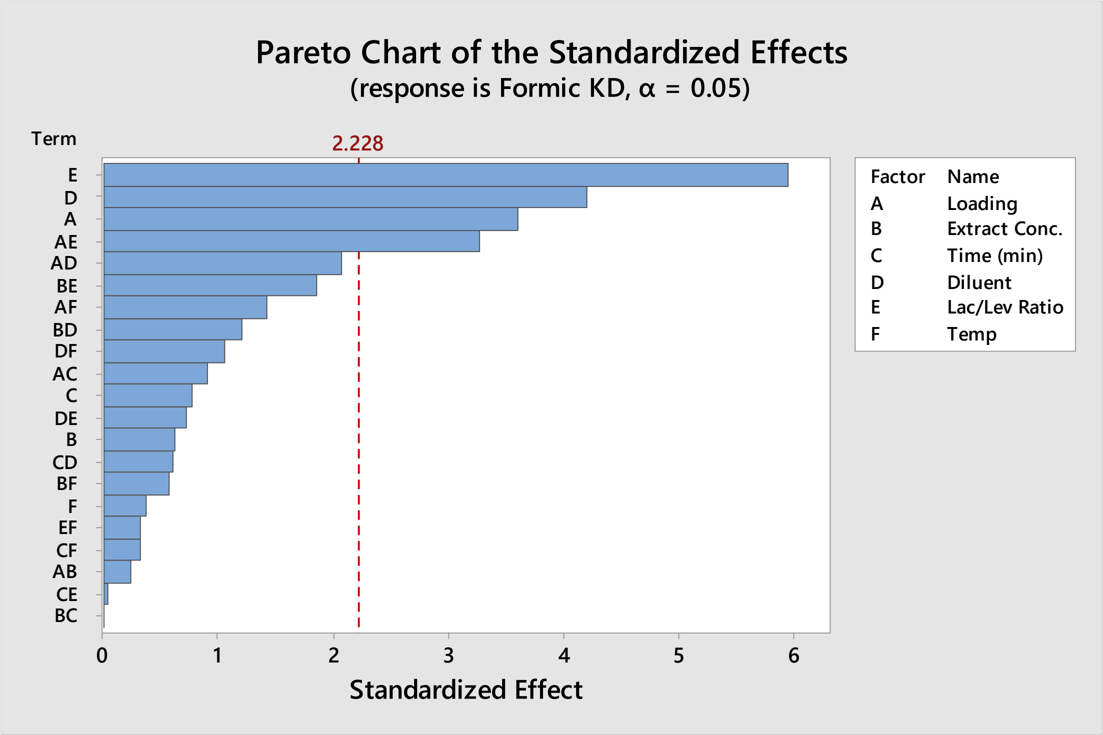

Figure 38. High and low means of significant two-interactions for formic acid reactive extraction.

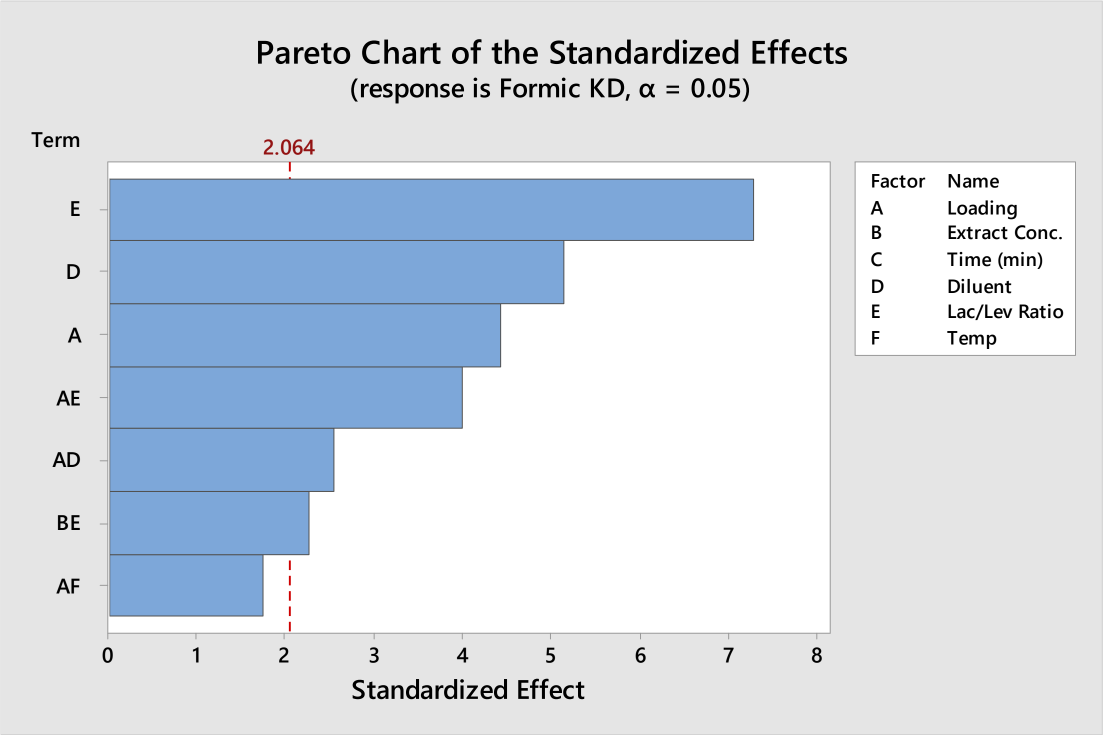

Figure 39. Reduced model for the examined effects of formic acid reactive extraction.

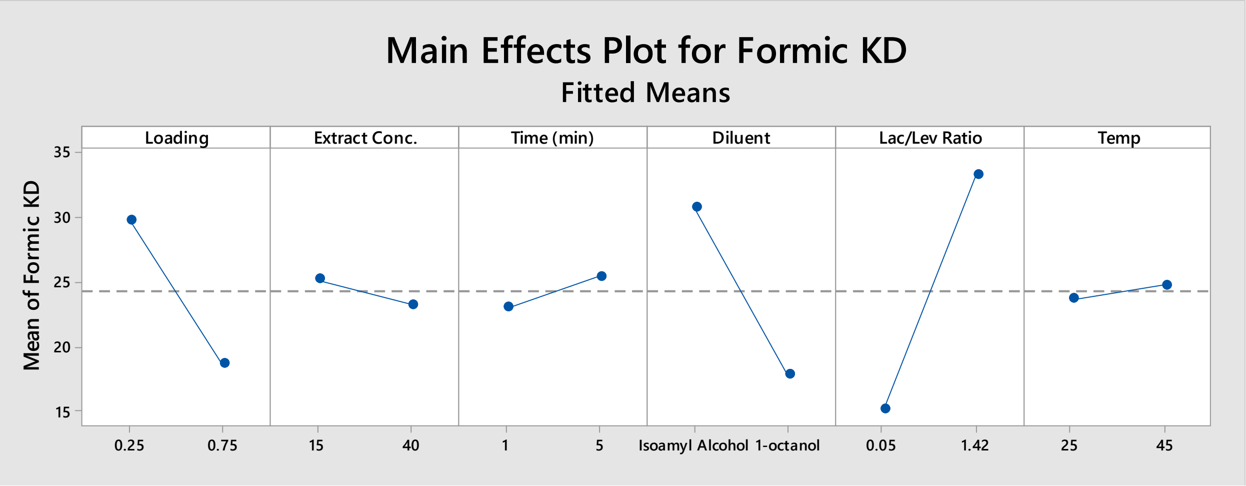

Figure 40. High and low means of significant main effects for formic acid reactive extraction.

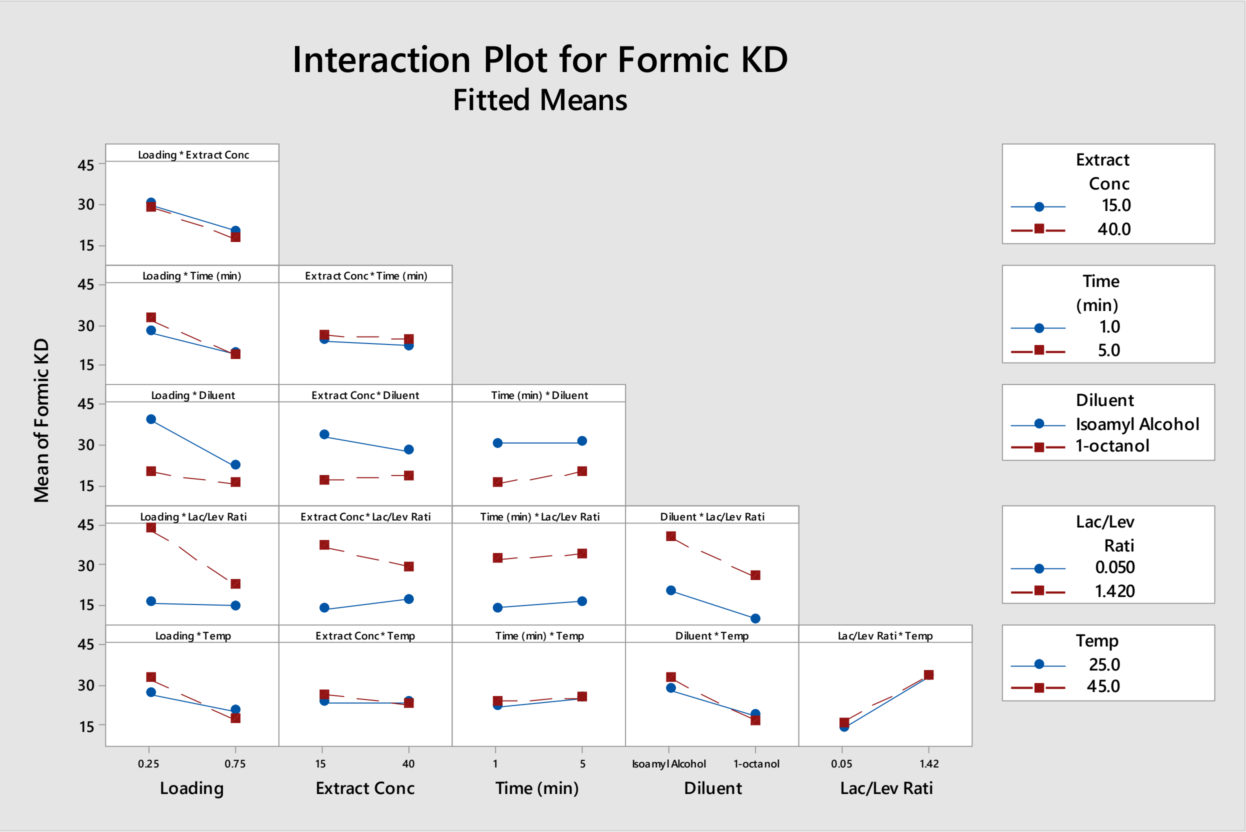

Figure 41. High and low means of significant two-interactions for formic acid reactive extraction.

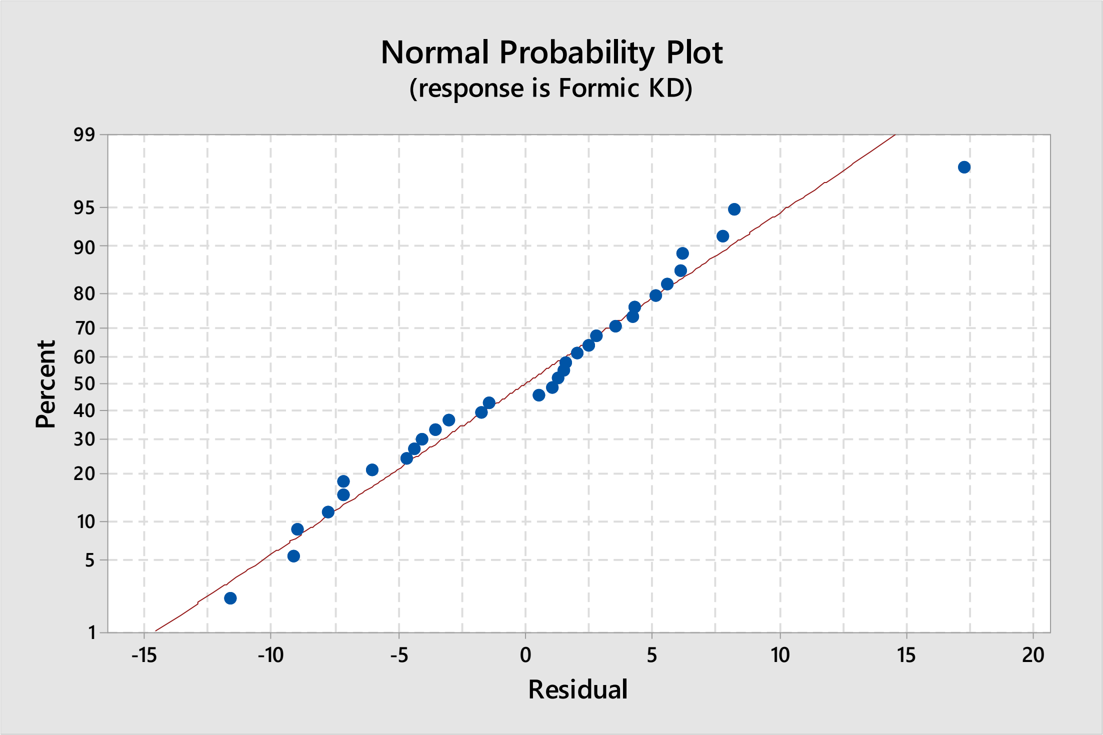

Figure 42. Normal probability distiributron of acetic acid extraction model.

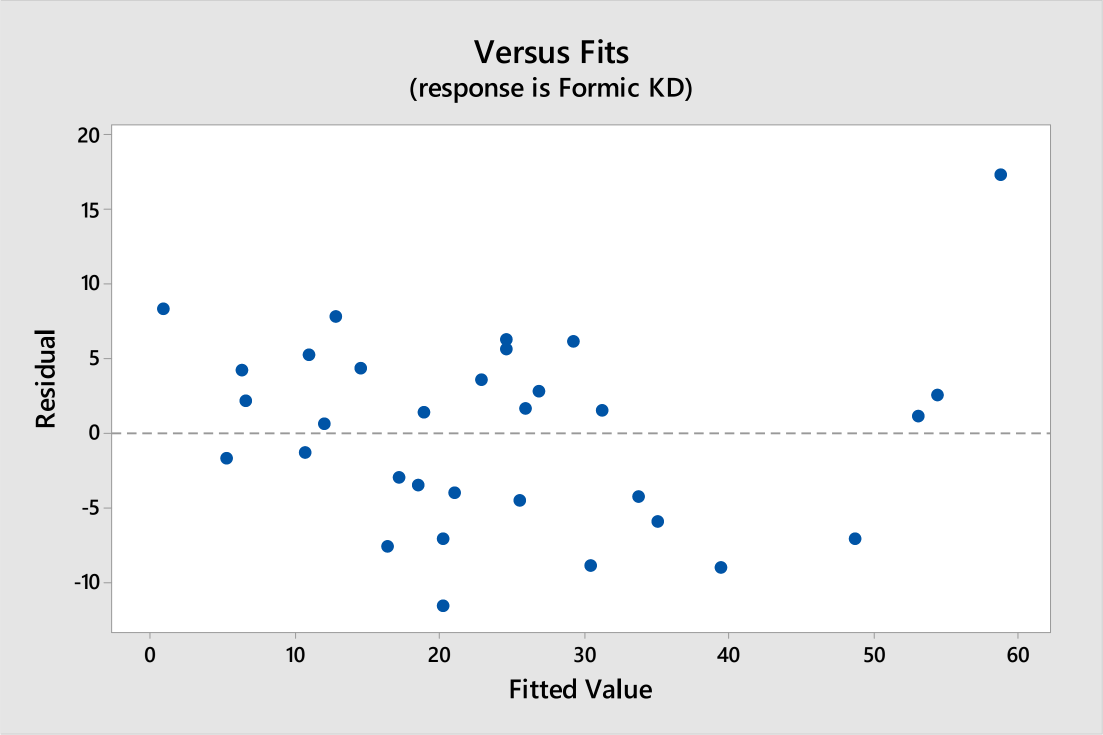

Figure 43. Versus fit plot of formic acid extraction model

APPENDIX D – EXTRACTION ALTERNATIVE ECONOMICS

Table 31. Capital Cost Summary for 1-octanol extraction alternative.

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 1 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-1	TOA Mixing Drum	2	Height = 8.0 m Diameter = 3.9 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$17,000	$26,000	2.5	1	5.3	$140,000	$280,000
D-2	TOA Settler	2	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Steel Cladding	$17,000	$26,200	2.5	1	5.3	$140,000	$280,000

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 2 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-3	Light Acids Column Top	1	Rectifying Diameter = 0.61m Height = 7 m Pressure = 410 kPa MOC = Carbon Steel with Stainless Clad	$9,800	$15,000	2.5	1.1	7.5	$110,000	$110,000
D-3	Light Acids Column Bottom	1	Stripping Diameter = 4.8 m Height = 6 m Pressure = 410 kPa MOC = Carbon Steel with Stainless Clad	$100,000	$160,000	2.5	1.1	7.5	$1,200,000	$1,200,000

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 3 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-3 Trays	Light Acids Column Trays Top	9	Rectifying Diameter = 0.61m Number of Trays = 20 MOC = Stainless Steel	$300	$460	2.2	1.4	3.08	$1,400	$13,000
D-3 Trays	Light Acids Column Trays Bottom	10	Stripping Diameter = 4.8 m Number of Trays = 15 MOC = Stainless Steel	$8,000	$12,000	2.2	1.07	2.354	$30,000	$300,000

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 4 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-4	TMA/Water Mixing Drums	4	Height = 7.8 m Diameter = 3.9 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	5.8	$130,000	$540,000
D-5	TMA/Water Settlers	4	Height = 2.1 m Diameter = 4.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	6.5	$150,200	$600,000

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 5 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-6	TMA/Water Column Top	1	Diameter = 2.0 m Height = 7.5 m # of Trays = 10 Pressure = 218 kPa MOC = Carbon Steel with Stainless Steel Cladding	$6,90	$11,000	2.5	1.1	7.5	$80,000	$80,000
D-6	TMA/Water Column Bottom	1	Diameter = 2.0 m Height = 7.5 m # of Trays = 10 Pressure = 218 kPa MOC = Carbon Steel with Stainless Clad	$17,000	$27,000	2.5	1.1	7.5	$200,000	$200,000

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 6 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-6 Trays	TMA/Water Column Trays Top	5	Diameter = 2.0 m Number of Trays = 5 Pressure = 218 kPa MOC = Stainless Steel	$1,300	$2,000	2.2	1.2	2.64	$5,300	$26,000
D-6 Trays	TMA/Water Column Trays Bottom	5	Diameter = 2.0 m Number of Trays = 10 Pressure = 218 kPa MOC = Stainless Steel	$3,500	$5,400	2.2	1.4	3.08	$17,000	$82,000

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 7 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-7	Acid/l-octanol Pressure Swing Adsorber	4	Total Height = 8.9 m Diameter = 3.5 m Maximum Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	\$63,000	\$97,000	2.5	1.1	6.5	\$630,000	\$2,500,000
E = Heat Exchangers										
E-1	Light Acids Condensor	2	Area = 21 m^2 Shell/Tube MOC = cs/Cu Max. Pressure = 365 kPa	\$4,500	\$7,000	1.25	1	3.5	\$24,000	\$48,900

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 8 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
E = Heat Exchangers										
E-2	Light Acids Reboiler	3	Area = 900 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 3300 kPa	$44,000	$68,000	1.25	1	3.5	$240,000	$710,000
E-3	Pre-Back Extraction Cooler	2	Area = 580 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 3300 kPa	$40,000	$62,000	1.25	1	3.5	$220,000	$430,000
E-4	TMA/Water Column Condesnsor	2	Area = 480 m^2 Shell/Tube MOC = cs/cs Max Pressure = 310 kPa	$22,000	$35,000	1.25	1	3.5	$120,000	$240,000

Table 31. Capital Cost Summary for 1-octanol extraction alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 9 of 9
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
M = Mixers										
E-5	TMA/Water Column Reboiler	2	Area = 300 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 3310 kPa	$15,000	$23,000	1.25	1	3.5	$82,000	$160,000

Total Bare Module Cost	$7,800,000
Contingency and Fee (18%)	$1,400,000
Total Module Cost	$9,100,000
Auxiliary Factor (30%)	$2,800,000
Fixed Capital Investment	$12,000,000
Working Capital	$1,800,000
Chemicals and Catalyst	$390,000
TCI	$14,000,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 1 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-1	TOA/Isoamyl Mixing Drums	2	Height = 4.8 m Diameter = 2.4 m MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1	5.3	$120,000	$240,000
D-2	TOA/Isoamyl Settlers	2	Height = 4.8 m Diameter =2.4 m MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1	5.3	$120,000	$240,000
D-3	Acid Splitter Column Top	1	Height = 7.4 m Diameter = 1.8 m Pressure=130 kPa MOC= Carbon Steel with Stainless Steel Cladding	$26,000	$39,000	2.5	1.1	7.5	$300,000	$300,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 2 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-3	Acid Splitter Column Bottom	1	Height = 4.4 m Diameter = 2.4 m Pressure=130 kPa MOC= Carbon Steel with Stainless Steel Cladding	$28,000	$43,000	2.5	1.1	7.5	$320,000	$320,000
D-3 Trays	Acid Splitter Column Trays Top	6	Diameter = 1.8 ft Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$2,000	$4,000	2.2	1.45	3.19	$10,000	$70,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 3 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-3 Trays	Acid Splitter Column Trays Bottom	11	Diameter = 2.4 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$4,000	$7,000	2.2	1.2	2.64	$20,000	$200,000
D-4	TMA/Water Mixing Drums	4	Diameter = 1.9 m Height = 3.8 m Pressure = 130 kPa MOC = Carbon Steel with Nickel Cladding	$17,000	$26,000	2.5	1.1	5.8	$150,000	$610,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 4 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-5	TMA/Water Settlers	4	Diameter = 1.9 m Height = 3.8 m Pressure = 120 kPa MOC = Carbon Steel with Stainless Steel Cladding	$17,000	$26,000	2.5	1.1	6.5	$170,000	$680,000
D-6	TMA Recovery Column Top	1	Diameter = 1.5 m Height = 1.9 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$12,000	$19,000	2.5	1.1	7.5	$140,000	$140,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 5 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-6	TMA Recovery Column Bottom	1	Diameter = 1.4 m Height = 21 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$95,000	$146,000	2.5	1.1	7.5	$1,090,000	$1,093,000
D-6 Trays	TMA Recovery Column Trays Top	1	Diameter = 1.5 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$1,000	$2,000	2.2	3.6	7.92	$10,000	$12,300

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 6 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-6 Trays	TMA Recovery Column Trays Bottom	34	Diameter = 1.4 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$2,000	$4,000	2.2	1	2.2	$8,300	$282,000
D-7	Extractive Distilation Column Top	1	Diameter = 1.7 m Height = 18 m Pressure = 63 psia MOC = Carbon Steel with Stainless Steel Cladding	$66,000	$101,000	2.5	1.1	7.5	$760,000	$760,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 7 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-7	Extractive Distilation Column Bottom	1	Diameter = 2.7 m Height = 18 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$120,000	$190,000	2.5	1.1	7.5	$1,400,000	$1,400,000
D-7 Trays	Extractive Distilation Column Trays Top	29	Diameter = 1.7 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$270	$410	2.2	1	2.2	$900	$26,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 8 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-7 Trays	Extractive Distilation Column Trays Bottom	29	Diameter = 2.7 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$530	$820	2.2	1	2.2	$1,800	$53,000
D-8	Extractant Recovery Column Top	1	Diameter = 0.58 m Height = 3.7 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$6,000	$9,000	2.5	1.1	7.5	$68,800	$69,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 9 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-8	Extractant Recovery Column Bottom	1	Diameter = 1.2 m Height = 5.0 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$32,000	$49,000	2.5	1.1	7.5	$369,200	$369,000
D-8 Trays	Extractant Recovery Column Trays Top	5	Diameter = 0.58 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$570	$890	2.2	1.5	3.3	$2,900	$15,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 11 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
D = Pressure Vessels										
D-8 Trays	Extractant Recovery Column Trays Bottom	7	Diameter = 1.2 m Pressure = 130 kPa MOC = Carbon Steel with Stainless Steel Cladding	$2,200	$3,370	2.2	1.4	3.08	$10,000	$73,000
E = Heat Exchangers										
E-1	Light Acids Condensor	2	Area = 580 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 450 psia	$4,500	$7,000	1.25	1	3.5	$24,000	$48,900

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 11 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
E = Heat Exchangers										
E-2	Acid Splitter Reboiler	3	Area= 750 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 120 kPa	$44,000	$68,000	1.25	1	3.5	$240,000	$720,000
E-3	TMA Column Condesnsor	3	Transfer Area = 745 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 130 kPa	$48,000	$74,000	1.25	1	3.5	$260,000	$780,000
E-4	TMA Column Reboiler	2	Transfer Area = 220 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 130 kPa	$4,800	$7,440	1.25	1	3.5	$26,000	$52,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 12 of 13
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
E = Heat Exchangers										
E-5	Extractive Distilation Column Condesnsor	2	Transfer Area = 380 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 130 kPa	$32,000	$50,000	1.25	1	3.5	$170,000	$350,000
E-6	Extractive Distilation Column Reboiler	5	Transfer Area = 870 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 130 kPa	$53,000	$81,000	1.25	1	3.5	$280,000	$1,400,000
E-7	Extractant Recovery Column Condesnsor	2	Transfer Area = 19 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 130 kPa	$5,300	$8,200	1.25	1	3.5	$29,000	$58,000

Table 32. Capital Cost Summary for Isoamyl Extraction Alternative

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 13 of 13

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
M = Mixers										
E-8	Extractant Recovery Column Column Reboiler	2	Transfer Area = 170 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 130 kPa	$900	$1,300	1.25	1	3.5	$5,000	$9,200
E-9	Extractant Cooler	2	Transfer Area = 31 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 250 psia	$7,200	$11,000	1.25	1	3.5	$39,000	$77,000

Total Bare Module Cost	$11,000,000
Contingency and Fee (18%)	$2,000,000
Total Module Cost	$13,000,000
Auxiliary Factor (30%)	$3,800,000
Fixed Capital Investment	$17,000,000
Working Capital	$2,500,000
Chemicals and Catalyst	$500,000
TCI	$20,000,000

Table 33. Utility requirements for extraction alternatives.

Utility Description	*1-octanol Alternative* Annual Cost ($/yr)	Requirement	*Isoamyl Alternative* Annual Cost ($/yr)	Requirement
Cooling Water	$560,000	Amount: 1.8 billion kg/yr Temperature: 30°C Pressure: 300 kPa	$2,100,000	Amount: 29 billion kg/yr Temperature: 30°C Pressure: 300 kPa
Boiler Feed Water	--	--	$278,000	Amount: 22 million kg/yr Temperature: 30°C Pressure: 450 kPa
Low Pressure Steam	--	--	$1,300,000	Amount: 50 million kg/yr Temperature: 140°C Pressure: 450 kPa
Medium-High Pressure Steam	--	--	$5,600,000	Amount: 300 million kg/yr Temperature: 239°C Pressure: 3,300 kPa
High Pressure Steam	$8,500,000	Amount: 440 million kg/yr Temperature: 400°C Pressure: 4,500 kPa	$2,000,000	Amount: 86 million kg/yr Temperature: 400°C Pressure: 4,500 kPa
Total	$9.0 million/yr		$11 million/yr	

Table 34. Chemical and Catalyst requirements for each acid extraction alternative

		1-octanol Alternative		*Isoamyl Alternative*	
Chemical/Catalyst	Cost ($/kg)	Initial Requirement (kg)	Make-Up Requirements (kg/year)	Initial Requirement (kg)	Make-Up Requirements (kg/year)
1-octanol	$0.44	67,000	10,000	--	--
13x Molecular Sieve	$1.32	45,000	2,300	--	--
Azelaic Acid	$9.09	--	--	9,100	1,400
Isoamyl Alcohol	$1.00	--	--	54,000	8,000
Trimethylamine (TMA)	$2.68	12,000	1,700	*15,000*	2,300
Trioctylamine (TOA)	$8.50	20,000	3,000	8,100	1,200
Total Cost		$390,000	$82,000/year	$500,000	$75,000

Table 35. Breakdown of annual operating expenses for each process.

Process Alternative	Year	Raw Materials	Chemicals & Catalysts	Operating Labor	Maintenance	Utilities	Total
1-octanol	1-20	--	\$82,000	\$400,000	\$480,000	\$9,000,000	\$10,000,000
Isoamyl Alcohol	1-20	--	\$74,000	\$400,000	\$4800,000	\$11,000,000	\$12,000,000

Table 36. Cash Flow Sheet for 1-octanol extraction alternative

Year	Revenues	Operating Costs	Gross Profit	Depreciation	Fed. Taxable Profit	Fed. Income Tax	State Taxable Profit	State Income Tax	Nontaxable Charges	Net Profit	Present Value @ 20%
-2									($4.0)	($4.0)	($5.7)
-1									($4.0)	($4.0)	($4.8)
0									($6.2)	($6.2)	($6.2)
1	$0	($10)	($10)	$2.3	($12)	$2.6	($9.7)	$0.42		($7.0)	($5.8)
2	$0	($10)	($10)	$1.1	($11)	$2.3	($8.8)	$0.38		($7.3)	($5.1)
3	$0	($10)	($10)	$1.0	($11)	$2.3	($8.7)	$0.37		($7.3)	($4.2)
4	$0	($10)	($10)	$0.88	($11)	$2.3	($8.6)	$0.37		($7.3)	($3.5)
5	$0	($10)	($10)	$0.78	($11)	$2.4	($8.9)	$0.38		($7.7)	($3.1)
6	$0	($10)	($10)	$0.69	($11)	$2.2	($8.4)	$0.36		($7.4)	($2.5)
7	$0	($10)	($10)	$0.61	($11)	$2.2	($8.4)	$0.36		($7.4)	($2.1)
8	$0	($10)	($10)	$0.54	($11)	$2.2	($8.3)	$0.36		($7.4)	($1.7)
9	$0	($10)	($10)	$0.47	($11)	$2.3	($8.7)	$0.37		($7.8)	($1.5)
10	$0	($10)	($10)	$0.44	($10)	$2.2	($8.3)	$0.36		($7.5)	($1.2)
11	$0	($10)	($10)	$0.44	($10)	$2.2	($8.3)	$0.36		($7.5)	($1.0)
12	$0	($10)	($10)	$0.44	($10)	$2.2	($8.3)	$0.36		($7.5)	($0.84)
13	$0	($10)	($10)	$0.44	($11)	$2.3	($8.6)	$0.37		($7.8)	($0.73)
14	$0	($10)	($10)	$0.44	($10)	$2.2	($8.3)	$0.36		($7.5)	($0.58)
15	$0	($10)	($10)	$0.44	($10)	$2.2	($8.3)	$0.36		($7.5)	($0.48)
16	$0	($10)	($10)	$0.44	($10)	$2.2	($8.3)	$0.36		($7.5)	($0.40)
17	$0	($10)	($10)	$0.44	($11)	$2.3	($8.6)	$0.37		($7.8)	($0.35)
18	$0	($10)	($10)		($10)	$2.1	($7.9)	$0.34		($7.6)	($0.28)
19	$0	($10)	($10)		($10)	$2.1	($7.9)	$0.34		($7.6)	($0.24)
20	$0	($10)	($10)		($10)	$2.1	($7.9)	$0.34	$1.8	($5.8)	($0.15)
										NPV@20%	($52)

Note: Red numbers in parentheses represent negative values

Table 37. Cash Flow Sheet for isoamyl alcohol extraction alternative

Year	Revenues	Operating Costs	Gross Profit	Depreciation	Fed. Taxable Profit	Fed. Income Tax	State Taxable Profit	State Income Tax	Nontaxable Charges	Net Profit	Present Value @ 20%
-2									($5.6)	($5.6)	($8.1)
-1									($5.6)	($5.6)	($6.7)
0									($8.6)	($8.6)	($8.6)
1	$0	($12)	($12)	$2.9	($15)	$3.2	($11.9)	$0.51		($8.6)	($7.1)
2	$0	($12)	($12)	$1.6	($14)	$2.9	($11.0)	$0.47		($8.9)	($6.2)
3	$0	($12)	($12)	$1.5	($14)	$2.9	($10.8)	$0.47		($8.9)	($5.2)
4	$0	($12)	($12)	$1.28	($14)	$2.8	($10.7)	$0.46		($8.9)	($4.3)
5	$0	($12)	($12)	$1.13	($13)	$2.8	($10.6)	$0.46		($9.0)	($3.6)
6	$0	($12)	($12)	$1.00	($13)	$2.8	($10.5)	$0.45		($9.0)	($3.0)
7	$0	($12)	($12)	$0.88	($13)	$2.8	($10.4)	$0.45		($9.0)	($2.5)
8	$0	($12)	($12)	$0.78	($13)	$2.7	($10.3)	$0.44		($9.1)	($2.1)
9	$0	($12)	($12)	$0.68	($13)	$2.7	($10.2)	$0.44		($9.1)	($1.8)
10	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($1.5)
11	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($1.2)
12	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($1.02)
13	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($0.85)
14	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($0.71)
15	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($0.59)
16	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($0.49)
17	$0	($12)	($12)	$0.64	($13)	$2.7	($10.2)	$0.44		($9.1)	($0.41)
18	$0	($12)	($12)		($12)	$2.6	($9.7)	$0.42		($9.3)	($0.35)
19	$0	($12)	($12)		($12)	$2.6	($9.7)	$0.42		($9.3)	($0.29)
20	$0	($12)	($12)		($12)	$2.6	($9.7)	$0.42	$2.5	($6.7)	($0.18)
										NPV@20%	($67)

Note: Red numbers in parentheses represent negative values

APPENDIX E – BASE SN-BETA PROCESS DESCRIPTION

Process Area 01: Corn Stover Pretreatment

To enhance the extraction of sugars from corn stover the feedstock needs to be treated with both 0.4 wt% sodium hydroxide (NaOH) and 1 wt% sulfuric acid (H_2SO_4). By adding a dilute NaOH solution acetate, ash, and other extractables can be removed from the stover, which in turn increases downstream hydrolysis yields. Due to corn stover being a lignocellulosic biomass it is largely composed of cellulose and hemicellulose. To maximize the sugar yield from cellulose, the hemicellulose needs to be removed first. This can be accomplished by adding H_2SO_4 to the corn stover.

Due to the corrosive nature of NaOH and H2SO4 in solution, the process streams are mixed in the process area (instead of offsite) to prevent excessive use of corrosion resistant materials. Concentrated sulfuric acid (96 wt%, steam 1) is pumped from off-site as stream 1 at 25°C and 310 kPa to the sulfuric acid mixing vessel (D-100 seen on 01-A-002/1). Here the sulfuric acid is mixed with process water (Stream 2), diluting the acid to 1 weight percent. The sulfuric acid mixture is then sent to acid hydrolysis (R-120). Solid NaOH is conveyed (J-107 A/B) (stream 4) from outside of the process area to be fed into the NaOH mixing vessel (D-105). Process water (stream 5) from wastewater treatment is fed into D-105 to dilute the NaOH to 0.4 wt%. For simplicity, NaOH is fed at atmospheric pressure (100 kPa) making D-105 operate at 100 kPa. The 0.4 wt% NaOH solution (Stream 6) is then pumped up to 45 psia and sent to the deacetylation vessel (D-110 A-C/D).

Dry corn stover enters the process (stream 10) by a belt conveyor (J-112 A-C/D) and is sent to a deacetylation vessel (D-110 A-C/D) which operates at 100 kPa and 80° C

(01-A-002/2). Here the corn stover is washed with the 0.4 wt% sodium hydroxide solution (stream 9) from D-100 to de-ash the corn. Low pressure steam, stream 501, is used to heat this vessel to operating temperature of 80° C. After 48 minutes of residence time the vessel is allowed to settle. The washed corn stover then leaves through a screw conveyor (J-111 A-C/D) and is then transferred to a belt conveyor (J-114 A-C/D)(stream 10). A black liquor stream (stream 11) is produced containing about 22% of the original biomass including 88% of the acetate, 75% of the ash, and 100% of the water-soluble extractives. Stream 8 is then sent to wastewater treatment before it is recycled back to the NaOH mixer (D-105). This process is conducted as a semi-batch operation and is further described on the first supplemental PFD (01-A-003).

The belt conveyors (J-114 A-C/D) leaving the four deacetylation vessels (stream 10) are fed to one conveyor belt (J-112) which feeds the acid presteamer (D-121) seen on 01-A-002/3. Medium-high pressure steam (stream 12) is directly added to the corn stover to heat it up to 160° C. Once heated, the stover is funneled into the acid hydrolysis reactor (R-120) (stream 14) and combined with 1 wt% sulfuric acid (stream 4). The high solids content within the reactor requires a screw of 29 kW to transfer it forward and increase the pressure from 82 kPa to 560 kPa. Medium-high pressure steam is directly added into the reactor for it to remain isothermal (stream 14). The main goal of the acid hydrolysis reactor is to break apart the hemi-cellulose in the corn stover, allowing access to the cellulose. Table 1 shows all of the reactions containing the major components of corn stover.

The partially hydrolyzed corn stover leaving the screw of R-120 (stream 15) is directly sent to a rotary vacuum filter (H-131 A-G on 01-A-002/4). Here the newly formed, water-soluble sugars are removed from the solid corn stover. Process water (stream 19) enters at a 2.5:1 weight ratio with the corn stover solides to ensure 99% product recovery. The filter operates at 55 kPa, providing a vacuum to pull the liquids away from the solids. Caked on solids are removed from the filter (stream 17) and are conveyed (J-134 A-G) to enzymatic hydrolysis. Stream 18 is routed to a knockout drum (D-132) to remove the air pulled from the atmosphere due to the vacuum. The air is routed to a compressor (G-135) that generates the vacuum and produces compressed air (Stream 23). The bottom stream out of the knockout drum is routed through a pump (L-133 A/B) that increases the pressure from 34 to 140 kPa. This bottom stream (stream 19) is the acid hydrolysate.

The acid hydrolysate (stream 19) is mixed with calcium carbonate within the neutralization reactor (R-150 on 01-A-002/5). The neutralization tank operates at 99° C and 240 kPa with a residence time of 5 minutes. Calcium carbonate (stream 27) reacts with sulfuric acid from stream 19 to form partially insoluble calcium sulfate, gaseous carbon dioxide, and water. Carbon dioxide is vented out the top of the drum (stream 22) and released to the atmosphere. The slurry exiting the bottom of the drum (stream 23) is pumped (L-151 A-B/C) to increase the pressure from 240 to 380 kPa.

The neutralization slurry (stream 23) is sent to a rotary vacuum filter (H-160) to remove solid gypsum (Stream 25 on 01-A-002/6). Gypsum consists of the produced calcium sulfate hydrated with two moles of water. Like the acid hydrolysate filter, process water (stream 24) is added at a 2.5:1 mass ratio to wash the solid gypsum caked on the filter, ensuring a high liquid product recovery. The gypsum filter operates under vacuum conditions at 55 kPa, pulling fluids to the gypsum filtration knockout drum (D-161)(Stream 26). Here the air from the atmosphere (stream 27) that is pulled through the filter is suctioned out the top of the drum through a compressor (G-164). The liquid, making up stream 28, is similarly suctioned out of the drum using a pump (L-165 A/B), increasing the pressure to 255 kPa. Gypsum (Stream 31) caked onto H-160 is removed from the filter and conveyed (J-162) to storage and is ready for sale.

Process Area 02: Glucose Recovery and Hydrolysate Concentration

Hemicellulose-free corn stover in stream 20 is transferred by multiple belt conveyors (J-134 A-G) from the acid rotary filters (H-131 A-G) to one, consolidated belt conveyor (J-204 on 02-A-002/1). The feed corn stover is then mixed with cellulase enzymes, stream 29, in a screw conveyor (J-202 A-H). The screw conveyor then feeds (stream 30) the continuous enzymatic reactor (R-200 A-H/I), which operates at 49°C and 280 kPa at a residence time of 24 hr. Water, stream 32, is added to the reactor to decrease the solid concentration. Due to solids concentration still being over 20%, the reactor solution is not readily pump-able. As gravity pulls the reactor solution down, enzymes work to break apart the cellulose in the corn stover making glucose. By the time the solution reaches the bottom, the broken down corn stover stream (stream 31) has a solid concentration low enough to be pumped. The reactor effluent (stream 31) is then sent to a hold-up tank (F-205), which can hold up to 12 hours of the liquid, as it waits to be sent to the following batch reactor. Since the tank operates at atmospheric pressure the exiting effluent (Stream 38) is pumped (L-201 A/B) out of the tank increasing the pressure to 88 psia.

Partially hydrolyzed corn stover (stream 31) from the pre-batch hold up tank (F-205) is sent to the batch enzymatic reactor (R-210 A-F/G seen on 02-A-002/2). Like the continuous reactor, the batch reactor operates at 49°C and 340 kPa but with a residence time of 60 hours. Due to the large residence time, six reactors are run in parallel through a semi-batch process. More information on the semi-batch process can be found on Drawing 02-A-003. While the reaction is running, the solution is continually stirred and temperature controlled (E-212 A-F/G/H) through a recycle line (stream 33). After the reaction has been completed, the products (stream 34) are moved to a hold up tank (F-213) to dampen out dynamics of the semi-batch system.

The goal of enzymatic hydrolysis is to break down the cellulose within the corn stover to more usable sugars. This de-polymerization reaction is the same that occurs within the acid hydrolysis reactor (R-120). Due to the use of cellulase enzymes, the only carbohydrates that react are glucan, producing glucose and glucose oligomers. These reactions and their conversions throughout the entire enzymatic reaction can be seen on Table 2.

Hydrolyzed corn stover (stream 34) from the post-batch hold up (F-213) is pumped (L-225 A-B/C) to a rotary vacuum filter (H-220 A-C seen on 02-A-002/3). Like the previous filters, process water (stream 35) enters at a 2.5:1 mass ratio with the solids caked on the filter, ensure 99% product recovery. Out of the filter, a lignin-rich stover is produced (stream 36) which is sent to the utility steam boiler and is used as a replacement for natural gas. The liquids (stream 39) are suctioned to a knockout drum (D-221) at 35 kPa. The top stream of the knockout drum is routed to a compressor (G-223) that generated the vacuum and produces compressed air (stream 44). The bottom stream (stream 39) out of the knockout drum is pulled into a pump (L-222 A/B) that changes the pressure from 35 to 260 kPa. This bottom stream (stream 39) that is produced is the enzymatic hydrolysate.

The enzymatic hydrolysate (stream 39) and neutralized acid hydrolysate (stream 28) are combined in a holdup drum (D-230) to help dampen out the dynamics of the two streams (02-A-002/4). Demineralized water (stream 40) is also added here to dilute the sugar stream to optimum Sn-Beta reaction conditions (1.6 wt%). This combined stream (stream 41) is then sent to the Sn-Beta catalytic reactors in process area 03.

Process Area 03: Catalytic Decomposition and Furfural Extraction

Combined hydrolysate (stream 41) from process area 02 is sent to a pump (L-301 A/B), which increases the stream pressure from 62 to 2100 kPa (03-A-002/1). After the pump, stream 41 is sent to a heat exchanger (E-302 A/B), where it is heated with medium-high pressure steam from 27 to 200 °C. The process fluid is then sent to a catalytic reactor comprised of Sn^{2+} doped beta zeolites (R-300 A-P/Q/R). With a two-hour residence time, glucose and xylose are selectively converted into the desired products of lactic, levulinic, formic, and acetic acid. All major reactions in the catalytic reactor can be seen in Table 3. Boiler feed water enters the jacket of the reactor to keep the reactor isothermal. This water is then vaporized producing low-moderate steam. Sugar that is not converted to an organic acid product is assumed to be hydrochar, which coats the surface of the catalyst. Due to this, 18 reactors are available with two being dedicated for hydrochar removal. To remove this hydrochar, compressed air is injected into the bottom of the reactor at 260°C and 1800 kPa. Oxygen in the air then reacts with the hydrochar forming carbon dioxide, which is then vented to the atmosphere. This process occurs for each reactor every eight days in a semi-batch sequence.

Sugar decomposition products are sent from R-300 to a heat exchanger (E-311 A/B) to cool the stream from 200 to 35°C (03-A-002/2). The cooled sugar decomposition products are sent to a flash drum (D-310) which reduces the pressure from 1780 kPa to a more operable 172 kPa (stream 46). Since the stream is cooled down to 35˚ C in E-311 for downstream operations, the rapid decrease in pressure will not result in a phase change at normal conditions. If vapor were to be produced it would likely contain formic acid and water. The vapor is then run through a chemical filter (D-312) which removes formic acid. Water remaining is vented to the atmosphere.

To prevent products from exceeding purity requirements, furfural needs to be extracted from the depressurized stream (stream 46). Furfural is a by-product produced from the dehydration of xylan in the acid hydrolysis reactor. Toluene is preferred over other solvents for this extraction due to its high distribution coefficient for furfural (KD =3) and essentially no affinity for organic acids and water. Furfural's high affinity to toluene results in only three stages being required to remove more than 99% of the furfural. Due to the low number of stages required to complete the separation, a series of mixer-settlers are utilized instead of a trayed column.

The depressurized decomposition products stream (stream 46) is mixed with toluene in the first stage extraction mixing drum (D-320 on 03-A-002/3). The first stage transfers 84% of furfural to toluene which is then phase separated (stream 51) from the aqueous process stream (stream 52) in D-321. The aqueous process stream is pumped (L-323) up to 170 kPa from 100 kPa and sent to the second stage extractor (D-325). The same steps occur for both the second and third stages which removes 13% and 2.5% of the overall furfural, respectively. Each mixing drum (D-320, D-325, and D-330) has a residence time of 5 minutes and operates at 35˚ C and 170 kPa. The subsequent settlers

(D-321, D-326, and D-331) have a residence time of 5 minutes and operate at 35˚ C and 100 kPa.

Furfural extractant streams (streams 51, 55, and 60) are combined in a hold-up drum (D-340) to dampen out the dynamics of the three extraction steps (03-A-002/5). All three steams enter at 35˚ C and 100 kPa so the hold-up drum operates at those conditions as well. The combined extractant (stream 61) is pumped out of the drum through L-341 A/B, which increases the stream pressure from 60 kPa to 1500 kPa.

Stream 61 is then sent to the toluene recovery column (D-350) which regenerates the toluene solvent and produces technical-grade furfural (03-A-002/6). This stream enters the toluene recovery column on stage 24 of the 29-stage column. A 99% pure toluene vapor (stream 61) leaves the top of the column and is totally condensed with cooling water in E-351 A/B. The condensate is then sent to a reflux drum (D-352) to help dampen out column dynamics. After the drum, the stream is split returning two-thirds of the toluene back to the column (stream 63) and sending the rest (stream 64) to a downstream hold-up drum (D-360). The reflux pump (L-353 A/B) increases the reflux pressure from 150 to 250 kPa, allowing it to re-enter the column. Out of the bottom of the column, a furfural-rich stream is sent to a thermosyphon reboiler (E-354 A/B). The reboiler operates with a boil-up ratio of 2.3, vaporizing most of the stream (stream 66) to return to the column. The remaining liquid (stream 67) out of the reboiler is recovered as technical-grade furfural and is transferred to storage to be sold.

Recovered toluene (stream 64) from D-350 enters the toluene recycle drum (D-360) at 110 °F and 140 kPa (03-A-002/7). When necessary, make-up toluene (stream 68) is added to the system here entering at 25 °C and 310 kPa. The combined stream (stream 57) leaves the drum at 110 °C and 117 kPa where it is pumped (L-361 A/B) up to 207 kPa. It is then cooled in E-202 A/B to 170 °C and is recycled back to the first-stage furfural mixer.

Compressed air required for removing hydrochar within the catalytic reactor is generated using a three-stage air compressor (G-370 on 03-A-002/8). Air enters as stream 42 at 70% humidity (average yearly humidity in North Dakota), and 24 °C, at 103 kPa. It's compressed to 184 °C and 280 kPa in the first stage of the compressor. The air is then cooled to 35 °C in E-371 A/B and is run through a knockout drum (D-372) to remove any condensed water. Under normal conditions none of the water is expected to condense. The stream is then cycled through the second stage which changes the air conditions to 172 °C and 780 kPa. The air is then partially cooled to 93 °C in E-373 A/B. The stream is not completely cooled because it is desired to have the air at 260 °C to calcine the reactor. At that temperature high pressure steam can no longer heat it so the compressor is operated less efficiently to reach that temperature. The air is sent to the third stage where it is compressed to 260 °C and 1800 kPa. The compression ratio is smaller in the final stage (2.3 compared to 2.7) because the third stage has to overcome more heat than the other two stages. To maintain the same duty in each stage (2,300 kW), the pressure increase was lowered in the final stage. The compressed air is then sent to the tin (II) zeolite packed bed reactor to regenerate and remove the kW from the catalysts.

Process Area 04: Acetic and Formic Acid Purification

The furfural-lean product stream (stream 59) from the third stage furfural extraction settler (D-331) is mixed with a trioctylamine/l-octanol solution (stream 69) in D-400 (04-A-002/1). Here trioctylamine forms an organic soluble-complex with the organic acids (acetic, formic, lactic, and levulinic acid), removing them from the aqueous phase. After being mixed, the solution (stream 73) is sent to D-401 were it is allowed to settle and the phases are separated. The aqueous phase with the un-extracted acids are pumped (L-403 A/B) from 110 kPa up to 230 kPa and sent to the second stage TOA mixing drum (D-405). This process is repeated for the second stage.

Trioctylamine was chosen as the extractant due to its higher recoveries of lactic and levulinic acid then other tertiary amines. Due to the high viscosity of TOA it needs to be diluted to make it pump-able. Of possible diluents, 1- octanol was chosen over methyl isobutyl ketone since it doesn’t form an azeotrope with any of the products, allowing for easier recovery. Distribution coefficients for each acid were based on experimentally found values.

First and second stage TOA acid extractant streams (streams 74 and 76) are combined in the TOA hold up drum (D-420) to smoothen out flow variations (03-A-002/2). Each extractant stream enters the drum at 35° C and 110 kPa so the drum operates at those conditions as well. The combined extractants are then pumped (L-421 A/B) up to a pressure of 450 kPa and sent to the volatile acid column (D-430).

In D-430, acetic and formic acid are distilled and removed from the heavy acids. This is done to prevent the mixture of water and formic acid in the back-extraction step, which is required to separate the heavier acids from the solvent. Formic acid and water form an azeotrope making it difficult to separate. Out the top of the column flows an l-octanol saturated volatile acid stream (stream 79). A third of the stream is condensed in E-431 A/B and is recycled back into the column (stream 81). The vapor (stream 80) is then sent on to D-440 to remove the 1-octanol. Heavy Acid extractant leaves the bottom of the column (stream 82) at 202 °C and 410 kPa where it enters a thermosyphon reboiler (E-434 A/B). A third of the stream is vaporized and returned to the column (stream 82) at

202 °C. The remaining liquid (stream 84) is cooled with boiler feed water in the heavy acid extractant cooler (E-435 A/B) from 202 °C to 35 °C and is sent to Process Area 05.

The distillate vapor (stream 79) from the volatile acid column is sent to molecular sieve column (D-440 A-D) to remove the remaining l-octanol and acids (04-A-002/4). Here the 13x molecular sieve (10 Å) takes advantage of the difference in molecular diameter between 1-octanol (~16 Å) and acetic acid (~8 Å). l-Octanol is then rejected by the molecular sieve and leaves D-440 at 99 °C and 160 kPa. The 1-octanol vapor is then condensed (stream 86) and cooled to 35 °C in E-443 A/B with cooling water. It is then sent to the TOA acid extractant hold up drum (D-530) to be recycled.

Both acetic and formic acid adsorbs to the molecular sieve. Once the molecular sieve becomes saturated the vapor is pulled out and is condensed (E-442 A/B) to 99 °C and 55 kPa with cooling water. The condensate (stream 85) is suctioned through a pump (L-441 A/B), leaving at 230 kPa, and continues to be further purified in a distillation column.

The volatile acids (stream 85) from the molecular sieve are sent Formic/Acetic Column (D-450 on 04-A-002/5). Technical-Grade Formic Acid (stream 89) leaves the top the column as a vapor and is condensed within E-451 A/B. The Formic Acid stream is then sent through the volatile acid column reflux drum (D-452) to dampen out column dynamics. From there 83% of the stream (stream 88) is sent back into the column while the rest (stream 89) is sent to storage to be sold as Technical-Grade Formic Acid. Out the bottom of the column leaves a mostly pure acetic acid stream (stream 90) at 121 °C and 403 kPa. The stream is sent to a thermosyphon reboiler (E-454 A/B) where 95% of the stream is vaporized (stream 91) and returns to the column at 121 °C and 402 kPa. The liquid (stream 92) leaves the reboiler at 121 °C and 365 kPa and is sent to storage as Technical-Grade Acetic Acid.

Process Area 05: Lactic and Levulinic Acid Purification

The heavy extractant stream from volatile acid recovery column (stream 84) containing 1-octanol, trioctylamine, lactic, and levulinic acid, which remains after the volatile acid column, is sent to D-500 (05-A-002/1). The organic process stream is mixed with a solution of trimethylamine (TMA) and water (stream 100). TMA forms a stronger attraction to lactic and levulinic acid than TOA, creating a water-soluble complex, transferring the acids back to the aqueous phase. TMA has a high affinity to both lactic and levulinic acid (estimated KD of 7 and 10, respectively) resulting in only 4 stages being required to achieve nearly 100% recovery.

After the streams are mixed in D-500, they are sent to D-501 to settle. The organic phase (stream 94) leaves the settler and is sent to the second stage mixing drum (D-505). This process is repeated for a total of four stages. Each stage removes about 90% of the acid that enters, giving an overall lactic and levulinic acid yield of 99.9%.

The aqueous phases (streams 94, 97, 107, & 110) from each extraction step are combined in the TMA extraction hold-up drum (D-520 on 05-A-002/3) operating at 35°C and 120 kPa. The outlet stream (stream 112) is then pumped (L-521 A/B) out of the drum, increasing the pressure to 230 kPa, and is sent to the back extractant recovery column (D-540).

The organic phase remaining (stream 111) after the fourth stage of back extraction is sent to the TOA hold-up drum (D-530 on 05-A-002/4). Here stream 111 is combined with the 1-octanol (stream 86) recovered from molecular sieve column (D-440). When needed, losses of TOA and 1-octanol are made up at this point. D-530 operates at the same conditions as the entering streams 86 and 111, 35 °F and 140 kPa. The solvent solution is pumped (L-531 A/B) out of the drum and recycled back to the first-stage TOA mixing drum (D-400).

The combined TMA back extractant stream (stream 112) enters the back extractant solvent recovery column (D-540 on 05-A-002/5) at stage 5 of 10. Vaporized TMA and water leaves the top of the column and is totally condensed with cooling water in E-541 A/B. The condensed stream is then sent to the column's reflux drum (D-542) to dampen out column dynamics. Of the stream exiting the drum, 4% is pumped back into the column to be recycled while the rest is sent to a hold-up drum (D-550). Lactic and levulinic acid leaves the bottom of the column at 224 °C and 218 kPa where it enters a thermosiphon reboiler (E-544 A/B). Over 80% of the stream is vaporized with medium-high pressure steam and returns to the column at 224 °C and 217 kPa. The remaining liquid (stream 120) leaves the reboiler at 224 °C and 180 kPa and is sent on to another distillation column to be further purified.

Recovered TMA and water, which leaves as the distillate in D-540, is sent to a hold-up drum (D-550 on 05-A-002/6). Make-up TMA and water are also added to the drum when needed. Since the primary stream entering the drum is the recovered TMA and water, the drum operates at its conditions 54 °C and 140 kPa. The stream is pumped (L-551 A/B) out of the drum, increasing the pressure to 260 kPa, and is sent to a heat exchanger (E-552 A/B). E-552 cools the stream from 54° C to the extraction temperature of 35° F. Once cooled, the stream is recycled back to the first-stage TMA mixing drum (D-500).

Lactic and levulinic acid (stream 120) from the back extractant recovery column is pumped (L-565) up to the 14th tray of the lactic/levulinic acid column (D-560 on 05-A-002/7). Vaporized polymer-grade lactic acid leaves the top of the column and is completely condensed by boiler-feed water in E-561 A/B. The condensed stream is sent through the column's reflux drum (D-562). From there, half of the stream is recycled back into the column and the other half is sent to storage to be sold as Polymer-Grade Lactic Acid. Out the bottom of the column leaves a food-grade levulinic acid stream. The stream then enters a thermosyphon reboiler which utilizes high-pressure steam to vaporize 74% of the entering stream. The produced vapor is recycled back into the column. The remaining liquid leaves the reboiler at 278° C and 360 kPa and is sent to storage as Food-Grade Levulinic Acid.

APPENDIX F – SN-BETA + $CaSO_4$ PROCESS DESCRIPTION

Process Area 01: Corn Stover Pretreatment

To enhance the extraction of sugars from corn stover the feedstock needs to be treated with both 0.4 wt% sodium hydroxide (NaOH) and 1 wt% sulfuric acid (H_2SO_4). By adding a dilute NaOH solution acetate, ash, and other extractables can be removed from the stover, which in turn increases downstream hydrolysis yields. Due to corn stover being a lignocellulosic biomass it is largely composed of cellulose and hemicellulose. To maximize the sugar yield from cellulose, the hemicellulose needs to be removed first. This can be accomplished by adding H_2SO_4 to the corn stover.

Due to the corrosive nature of NaOH and H_2SO_4 in solution, the process streams are mixed in the process area (instead of offsite) to prevent excessive use of corrosion resistant materials. Concentrated sulfuric acid (96 wt%, steam 1) is pumped from off-site as stream 1 at 25°C and 310 kPa to the sulfuric acid mixing vessel (D-100 seen on 01-A-002/1). Here the sulfuric acid is mixed with process water (Stream 2), diluting the acid to 1 weight percent. The sulfuric acid mixture is then sent to acid hydrolysis (R-120). Solid NaOH is conveyed (J-107 A/B) (stream 4) from outside of the process area to be fed into the NaOH mixing vessel (D-105). Process water (stream 5) from wastewater treatment is fed into D-105 to dilute the NaOH to 0.4 wt%. For simplicity, NaOH is fed at atmospheric pressure (100 kPa) making D-105 operate at 100 kPa. The 0.4 wt% NaOH solution (Stream 6) is then pumped up to 45 psia and sent to the deacetylation vessel (D-110 A-C/D).

Dry corn stover enters the process (stream 10) by a belt conveyor (J-112 A-C/D) and is sent to a deacetylation vessel (D-110 A-C/D) which operates at 100 kPa and 80° C (01-A-002/2). Here the corn stover is washed with the 0.4 wt% sodium hydroxide solution (stream 9) from D-100 to de-ash the corn. Low pressure steam, stream 501, is used to heat this vessel to operating temperature of 80° C. After 48 minutes of residence time the vessel is allowed to settle. The washed corn stover then leaves through a screw conveyor (J-111 A-C/D) and is then transferred to a belt conveyor (J-114 A-C/D)(stream 10). A black liquor stream (stream 11) is produced containing about 22% of the original biomass including 88% of the acetate, 75% of the ash, and 100% of the water-soluble extractives. Stream 8 is then sent to wastewater treatment before it is recycled back to the NaOH mixer (D-105). This process is conducted as a semi-batch operation and is further described on the first supplemental PFD (01-A-003).

The belt conveyors (J-114 A-C/D) leaving the four deacetylation vessels (stream 10) are fed to one conveyor belt (J-112) which feeds the acid presteamer (D-121) seen on 01-A-002/3. Medium-high pressure steam (stream 12) is directly added to the corn stover to heat it up to 160° C. Once heated, the stover is funneled into the acid hydrolysis reactor (R-120) (stream 14) and combined with 1 wt% sulfuric acid (stream 4). The high solids content within the reactor requires a screw of 29 kW to transfer it forward and increase the pressure from 82 kPa to 560 kPa. Medium-high pressure steam is directly added into the reactor for it to remain isothermal (stream 14). The main goal of the acid hydrolysis reactor is to break apart the hemi-cellulose in the corn stover, allowing access to the cellulose. Table 1 shows all of the reactions containing the major components of corn stover.

The partially hydrolyzed corn stover leaving the screw of R-120 (stream 15) is directly sent to a rotary vacuum filter (H-131 A-G on 01-A-002/4). Here the newly formed, water-soluble sugars are removed from the solid corn stover. Process water (stream 19) enters at a 2.5:1 weight ratio with the corn stover solides to ensure 99%

product recovery. The filter operates at 55 kPa, providing a vacuum to pull the liquids away from the solids. Caked on solids are removed from the filter (stream 17) and are conveyed (J-134 A-G) to enzymatic hydrolysis. Stream 18 is routed to a knockout drum (D-132) to remove the air pulled from the atmosphere due to the vacuum. The air is routed to a compressor (G-135) that generates the vacuum and produces compressed air (Stream 23). The bottom stream out of the knockout drum is routed through a pump (L-133 A/B) that increases the pressure from 34 to 140 kPa. This bottom stream (stream 19) is the acid hydrolysate.

The acid hydrolysate (stream 19) is mixed with calcium carbonate within the neutralization reactor (R-150 on 01-A-002/5). The neutralization tank operates at 99° C and 240 kPa with a residence time of 5 minutes. Calcium carbonate (stream 27) reacts with sulfuric acid from stream 19 to form partially insoluble calcium sulfate, gaseous carbon dioxide, and water. Carbon dioxide is vented out the top of the drum (stream 22) and released to the atmosphere. The slurry exiting the bottom of the drum (stream 23) is pumped (L-151 A-B/C) to increase the pressure from 240 to 380 kPa.

The neutralization slurry (stream 23) is sent to a rotary vacuum filter (H-160) to remove solid gypsum (Stream 25 on 01-A-002/6). Gypsum consists of the produced calcium sulfate hydrated with two moles of water. Like the acid hydrolysate filter, process water (stream 24) is added at a 2.5:1 mass ratio to wash the solid gypsum caked on the filter, ensuring a high liquid product recovery. The gypsum filter operates under vacuum conditions at 55 kPa, pulling fluids to the gypsum filtration knockout drum (D-161)(Stream 26). Here the air from the atmosphere (stream 27) that is pulled through the filter is suctioned out the top of the drum through a compressor (G-164). The liquid, making up stream 28, is similarly suctioned out of the drum using a pump (L-165 A/B), increasing the pressure to 255 kPa. Gypsum (Stream 31) caked onto H-160 is removed from the filter and conveyed (J-162) to storage and is ready for sale.

Process Area 02: Glucose Recovery and Hydrolysate Concentration

Hemicellulose-free corn stover in stream 20 is transferred by multiple belt conveyors (J-134 A-G) from the acid rotary filters (H-131 A-G) to one, consolidated belt conveyor (J-204 on 02-A-002/1). The feed corn stover is then mixed with cellulase enzymes, stream 29, in a screw conveyor (J-202 A-H). The screw conveyor then feeds (stream 30) the continuous enzymatic reactor (R-200 A-H/I), which operates at 49°C and 280 kPa at a residence time of 24 hr. Water, stream 32, is added to the reactor to decrease the solid concentration. Due to solids concentration still being over 20%, the reactor solution is not readily pump-able. As gravity pulls the reactor solution down, enzymes work to break apart the cellulose in the corn stover making glucose. By the time the solution reaches the bottom, the broken down corn stover stream (stream 31) has a solid concentration low enough to be pumped. The reactor effluent (stream 31) is then sent to a hold-up tank (F-205), which can hold up to 12 hours of the liquid, as it waits to be sent to the following batch reactor. Since the tank operates at atmospheric pressure the exiting effluent (Stream 38) is pumped (L-201 A/B) out of the tank increasing the pressure to 88 psia.

Partially hydrolyzed corn stover (stream 31) from the pre-batch hold up tank (F-205) is sent to the batch enzymatic reactor (R-210 A-F/G seen on 02-A-002/2). Like the continuous reactor, the batch reactor operates at 49°C and 340 kPa but with a residence time of 60 hours. Due to the large residence time, six reactors are run in parallel through a semi-batch process. More information on the semi-batch process can be found on Drawing 02-A-003. While the reaction is running, the solution is continually stirred and temperature controlled (E-212 A-F/G/H) through a recycle line (stream 33). After the reaction has been completed, the products (stream 34) are moved to a hold up tank (F-213) to dampen out dynamics of the semi-batch system.

The goal of enzymatic hydrolysis is to break down the cellulose within the corn stover to more usable sugars. This de-polymerization reaction is the same that occurs within the acid hydrolysis reactor (R-120). Due to the use of cellulase enzymes, the only carbohydrates that react are glucan, producing glucose and glucose oligomers. These reactions and their conversions throughout the entire enzymatic reaction can be seen on Table 2.

Hydrolyzed corn stover (stream 34) from the post-batch hold up (F-213) is pumped (L-225 A-B/C) to a rotary vacuum filter (H-220 A-C seen on 02-A-002/3). Like the previous filters, process water (stream 35) enters at a 2.5:1 mass ratio with the solids caked on the filter, ensure 99% product recovery. Out of the filter, a lignin-rich stover is produced (stream 36) which is sent to the utility steam boiler and is used as a replacement for natural gas. The liquids (stream 39) are suctioned to a knockout drum (D-221) at 35 kPa. The top stream of the knockout drum is routed to a compressor (G-223) that generated the vacuum and produces compressed air (stream 44). The bottom stream (stream 39) out of the knockout drum is pulled into a pump (L-222 A/B) that changes the pressure from 35 to 260 kPa. This bottom stream (stream 39) that is produced is the enzymatic hydrolysate.

The enzymatic hydrolysate (stream 39) and neutralized acid hydrolysate (stream 28) are combined in a holdup drum (D-230) to help dampen out the dynamics of the two streams (02-A-002/4). Water saturated with calcium sulfate (stream 40) is recycled from process area 05 here to dilute the sugar stream to optimum Sn-Beta reaction conditions (1.6 wt%). Calcium sulfate is brought into the process area on a conveyer belt (J-231) and added to the D-320 with a screw conveyor (J-232) (Stream 41). This additional of calcium sulfate greatly increases the selectivity of the Sn-Beta toward lactic acid. This combined stream (stream 42) is then sent to the Sn-Beta catalytic reactors in process area 03.

Process Area 03: Catalytic Decomposition and Furfural Extraction

Combined hydrolysate (stream 42) from process area 02 is sent to a pump (L-301 A/B), which increases the stream pressure from 62 to 2100 kPa (03-A-002/1). After the pump, stream 41 is sent to a heat exchanger (E-302 A/B), where it is heated with medium-high pressure steam from 27 to 200 °C. The process fluid is then sent to a catalytic reactor comprised of Sn^{2+} doped beta zeolites (R-300 A-P/Q/R). With a two-hour residence time, glucose and xylose are selectively converted into the desired products of lactic, levulinic, formic, and acetic acid. All major reactions in the catalytic reactor can be seen in Table 3. Boiler feed water enters the jacket of the reactor to keep the reactor isothermal. This water is then vaporized producing low-moderate steam. Sugar that is not converted to an organic acid product is assumed to be hydrochar, which coats the surface of the catalyst. Due to this, 18 reactors are available with two being dedicated for hydrochar removal. To remove this hydrochar, compressed air is injected into the bottom of the reactor at 260°C and 1800 kPa. Oxygen in the air then reacts with the hydrochar forming carbon dioxide, which is then vented to the atmosphere. This process occurs for each reactor every eight days in a semi-batch sequence.

Sugar decomposition products are sent from R-300 to a heat exchanger (E-311 A/B) to cool the stream from 200 to 35°C (03-A-002/2). The cooled sugar decomposition products are sent to a flash drum (D-310) which reduces the pressure from 1780 kPa to a more operable 172 kPa (stream 47). Since the stream is cooled down to 35˚ C in E-311 for downstream operations, the rapid decrease in pressure will not result in a phase change at normal conditions. If vapor were to be produced it would likely contain formic acid and water. The vapor is then run through a chemical filter (D-312) which removes formic acid. Water remaining is vented to the atmosphere.

To prevent products from exceeding purity requirements, furfural needs to be extracted from the depressurized stream (stream 46). Furfural is a by-product produced from the dehydration of xylan in the acid hydrolysis reactor. Toluene is preferred over other solvents for this extraction due to its high distribution coefficient for furfural (KD =3) and essentially no affinity for organic acids and water. Furfural's high affinity to toluene results in only three stages being required to remove more than 99% of the furfural. Due to the low number of stages required to complete the separation, a series of mixer-settlers are utilized instead of a trayed column.

The depressurized decomposition products stream (stream 47) is mixed with toluene in the first stage extraction mixing drum (D-320 on 03-A-002/3). The first stage transfers 84% of furfural to toluene which is then phase separated (stream 52) from the aqueous process stream (stream 53) in D-321. The aqueous process stream is pumped (L-323) up to 170 kPa from 100 kPa and sent to the second stage extractor (D-325). The same steps occur for both the second and third stages which removes 13% and 2.5% of the overall furfural, respectively. Each mixing drum (D-320, D-325, and D-330) has a residence time of 5 minutes and operates at 35˚ C and 170 kPa. The subsequent settlers

(D-321, D-326, and D-331) have a residence time of 5 minutes and operate at 35˚ C and 100 kPa.

Furfural extractant streams (streams 51, 55, and 60) are combined in a hold-up drum (D-340) to dampen out the dynamics of the three extraction steps (03-A-002/5). All three steams enter at 35˚ C and 100 kPa so the hold-up drum operates at those conditions as well. The combined extractant (stream 62) is pumped out of the drum through L-341 A/B, which increases the stream pressure from 60 kPa to 1500 kPa.

Stream 62 is then sent to the toluene recovery column (D-350) which regenerates the toluene solvent and produces technical-grade furfural (03-A-002/6). This stream enters the toluene recovery column on stage 24 of the 29-stage column. A 99% pure toluene vapor (stream 63) leaves the top of the column and is totally condensed with cooling water in E-351 A/B. The condensate is then sent to a reflux drum (D-352) to help dampen out column dynamics. After the drum, the stream is split returning two-thirds of the toluene back to the column (stream 64) and sending the rest (stream 65) to a downstream hold-up drum (D-360). The reflux pump (L-353 A/B) increases the reflux pressure from 150 to 250 kPa, allowing it to re-enter the column. Out of the bottom of the column, a furfural-rich stream is sent to a thermosyphon reboiler (E-354 A/B). The reboiler operates with a boil-up ratio of 2.3, vaporizing most of the stream (stream 67) to return to the column. The remaining liquid (stream 68) out of the reboiler is recovered as technical-grade furfural and is transferred to storage to be sold.

Recovered toluene (stream 65) from D-350 enters the toluene recycle drum (D-360) at 110 °F and 140 kPa (03-A-002/7). When necessary, make-up toluene (stream 69) is added to the system here entering at 25 °C and 310 kPa. The combined stream (stream 48) leaves the drum at 110 °C and 117 kPa where it is pumped (L-361 A/B) up to 207 kPa. It is then cooled in E-202 A/B to 170 °C and is recycled back to the first-stage furfural mixer.

Compressed air required for removing hydrochar within the catalytic reactor is generated using a three-stage air compressor (G-370 on 03-A-002/8). Air enters as stream 42 at 70% humidity (average yearly humidity in North Dakota), and 24 °C, at 103 kPa. It's compressed to 184 °C and 280 kPa in the first stage of the compressor. The air is then cooled to 35 °C in E-371 A/B and is run through a knockout drum (D-372) to remove any condensed water. Under normal conditions none of the water is expected to condense. The stream is then cycled through the second stage which changes the air conditions to 172 °C and 780 kPa. The air is then partially cooled to 93 °C in E-373 A/B. The stream is not completely cooled because it is desired to have the air at 260 °C to calcine the reactor. At that temperature high pressure steam can no longer heat it so the compressor is operated less efficiently to reach that temperature. The air is sent to the third stage where it is compressed to 260 °C and 1800 kPa. The compression ratio is smaller in the final stage (2.3 compared to 2.7) because the third stage has to overcome more heat than the other two stages. To maintain the same duty in each stage (2,300 kW), the pressure increase was lowered in the final stage. The compressed air is then sent to the tin (II) zeolite packed bed reactor to regenerate and remove the kW from the catalysts.

Process Area 04: Acetic and Formic Acid Purification

The furfural-lean product stream (stream 60) from the third stage furfural extraction settler (D-331) is mixed with a trioctylamine/l-octanol solution (stream 70) in D-400 (04-A-002/1). Here trioctylamine forms an organic soluble-complex with the organic acids (acetic, formic, lactic, and levulinic acid), removing them from the aqueous phase. After being mixed, the solution (stream 74) is sent to D-401 were it is allowed to settle and the phases are separated. The aqueous phase with the un-extracted acids are pumped (L-403 A/B) from 110 kPa up to 230 kPa and sent to the second stage TOA mixing drum (D-405). This process is repeated for five stages.

Trioctylamine was chosen as the extractant due to its higher recoveries of lactic and levulinic acid then other tertiary amines. Due to the high viscosity of TOA it needs to be diluted to make it pump-able. Of possible diluents, 1- octanol was chosen over methyl isobutyl ketone since it doesn't form an azeotrope with any of the products, allowing for easier recovery. Distribution coefficients for each acid were based on experimentally found values. Inclusion of calcium sulfate in the aqueous solution decreases the extraction efficiency of each acid by 90% compared to pure aqueous solutions.

TOA acid extractant streams (streams 76, 78, 86, 88, and 93) are combined in the TOA hold up drum (D-420) to smoothen out flow variations (03-A-002/4), operating at 35° C and 110 kPa. The combined extractants are then pumped (L-421 A/B) up to a pressure of 450 kPa and sent to the volatile acid column (D-430).

In D-430, acetic and formic acid are distilled and removed from the heavy acids. This is done to prevent the mixture of water and formic acid in the back-extraction step, which is required to separate the heavier acids from the solvent. Formic acid and water form an azeotrope making it difficult to separate. Out the top of the column flows an l-octanol saturated volatile acid stream (stream 97). A third of the stream is condensed in E-431 A/B and is recycled back into the column (stream 98). The vapor (stream 97) is then sent on to D-440 to remove the 1-octanol. Heavy Acid extractant leaves the bottom of the column (stream 99) at 202 °C and 410 kPa where it enters a thermosyphon reboiler (E-434 A/B). A third of the stream is vaporized and returned to the column (stream 100) at 202 °C. The remaining liquid (stream 101) is cooled with boiler feed water in the heavy acid extractant cooler (E-435 A/B) from 202 °C to 35 °C and is sent to Process Area 05.

The distillate vapor (stream 97) from the volatile acid column is sent to molecular sieve column (D-440 A-D) to remove the remaining l-octanol and acids (04-A-002/4). Here the 13x molecular sieve (10 Å) takes advantage of the difference in molecular diameter between 1-octanol (~16 Å) and acetic acid (~8 Å). l-Octanol is then rejected by the molecular sieve and leaves D-440 at 99 °C and 160 kPa. The 1-octanol vapor is then condensed (stream 103) and cooled to 35 °C in E-443 A/B with cooling water. It is then sent to the TOA acid extractant hold up drum (D-530) to be recycled.

Both acetic and formic acid adsorbs to the molecular sieve. Once the molecular sieve becomes saturated the vapor is pulled out and is condensed (E-442 A/B) to 99 °C and 55 kPa with cooling water. The condensate (stream 102) is suctioned through a pump (L-441 A/B), leaving at 230 kPa, and continues to be further purified in a distillation column.

The volatile acids (stream 102) from the molecular sieve are sent Formic/Acetic Column (D-450 on 04-A-002/7). Technical-Grade Formic Acid (stream 104) leaves the top the column as a vapor and is condensed within E-451 A/B. The Formic Acid stream is then sent through the volatile acid column reflux drum (D-452) to dampen out column dynamics. From there 83% of the stream (stream 105) is sent back into the column while the rest (stream 106) is sent to storage to be sold as Technical-Grade Formic Acid. Out the bottom of the column leaves a mostly pure acetic acid stream (stream 90) at 121 °C and 403 kPa. The stream is sent to a thermosyphon reboiler (E-454 A/B) where 95% of the stream is vaporized (stream 108) and returns to the column at 121 °C and 402 kPa.

The liquid (stream 109) leaves the reboiler at 121 °C and 365 kPa and is sent to storage as Technical-Grade Acetic Acid.

Process Area 05: Lactic and Levulinic Acid Purification

The heavy extractant stream from volatile acid recovery column (stream 101) containing 1-octanol, trioctylamine, lactic, and levulinic acid, which remains after the volatile acid column, is sent to D-500 (05-A-002/1). The organic process stream is mixed with a solution of trimethylamine (TMA) and water (stream 110). TMA forms a stronger attraction to lactic and levulinic acid than TOA, creating a water-soluble complex, transferring the acids back to the aqueous phase. TMA has a high affinity to both lactic and levulinic acid (estimated KD of 7 and 10, respectively) resulting in only 4 stages being required to achieve nearly 100% recovery.

After the streams are mixed in D-500, they are sent to D-501 to settle. The organic phase (stream 115) leaves the settler and is sent to the second stage mixing drum (D-505). This process is repeated for a total of four stages. Each stage removes about 90% of the acid that enters, giving an overall lactic and levulinic acid yield of 99.9%.

The aqueous phases (streams 115, 118, 124, & 127) from each extraction step are combined in the TMA extraction hold-up drum (D-520 on 05-A-002/3) operating at 35°C and 120 kPa. The outlet stream (stream 129) is then pumped (L-521 A/B) out of the drum, increasing the pressure to 230 kPa, and is sent to the back extractant recovery column (D-540).

The organic phase remaining (stream 128) after the fourth stage of back extraction is sent to the TOA hold-up drum (D-530 on 05-A-002/4). Here stream 128 is combined with the 1-octanol (stream 103) recovered from molecular sieve column (D-440). When needed, losses of TOA and 1-octanol are made up at this point. D-530 operates at the same conditions as the entering streams 103 and 128, 35 °C and 140 kPa. The solvent solution is pumped (L-531 A/B) out of the drum and recycled back to the first-stage TOA mixing drum (D-400).

The combined TMA back extractant stream (stream 129) enters the back extractant solvent recovery column (D-540 on 05-A-002/5) at stage 5 of 10. Vaporized TMA and water leaves the top of the column and is totally condensed with cooling water in E-541 A/B. The condensed stream is then sent to the column's reflux drum (D-542) to dampen out column dynamics. Of the stream exiting the drum, 4% is pumped back into the column to be recycled while the rest is sent to a hold-up drum (D-550). Lactic and levulinic acid leaves the bottom of the column at 224 °C and 218 kPa where it enters a thermosiphon reboiler (E-544 A/B). Over 80% of the stream is vaporized with medium-high pressure steam and returns to the column at 224 °C and 217 kPa. The remaining liquid (stream 137) leaves the reboiler at 224 °C and 180 kPa and is sent on to another distillation column to be further purified.

Recovered TMA and water, which leaves as the distillate in D-540, is sent to a hold-up drum (D-550 on 05-A-002/6). Make-up TMA and water are also added to the drum when needed. Since the primary stream entering the drum is the recovered TMA and water, the drum operates at its conditions 54 °C and 140 kPa. The stream is pumped (L-551 A/B) out of the drum, increasing the pressure to 260 kPa, and is sent to a heat exchanger (E-552 A/B). E-552 cools the stream from 54° C to the extraction temperature of 35° F. Once cooled, the stream is recycled back to the first-stage TMA mixing drum (D-500).

Lactic and levulinic acid (stream 137) from the back extractant recovery column is pumped (L-565) up to the 14th tray of the lactic/levulinic acid column (D-560 on 05-A-002/7). Vaporized polymer-grade lactic acid leaves the top of the column and is completely condensed by boiler-feed water in E-561 A/B. The condensed stream is sent through the column's reflux drum (D-562). From there, half of the stream is recycled back into the column and the other half is sent to storage to be sold as Polymer-Grade Lactic Acid. Out the bottom of the column leaves a food-grade levulinic acid stream. The stream then enters a thermosyphon reboiler which utilizes high-pressure steam to vaporize 74% of the entering stream. The produced vapor is recycled back into the column. The remaining liquid leaves the reboiler at 278° C and 360 kPa and is sent to storage as Food-Grade Levulinic Acid.

APPENDIX G – MAJOR EQUIPMENT LIST BASE SN-BETA ALTERNATIVE

Table 38. Major Equipment List for the Base Sn-Beta Process

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
D = Pressure Vessels		
D-100	Sulfuric Acid Mixing Vessel	Height = 2.0 m Diameter = 1.0 m Temperature = 25 °C Pressure = 310 kPa Volume = 1.6 m^3 Residence Time = 5 minutes MOC = Carbon Steel with Nickel Clad
D-105	NaOH Mixing Vessel	Height = 2.1 m Diameter = 4.2 m Temperature = 25 °C Pressure = 100 kPa Volume = 14.5 m^3 Residence Time = 5 minutes MOC = Carbon Steel
D-110 A-C/D	Deacetylation Vessel	Height = 8.5 m Diameter = 4.3 m Temperature = 80° C Pressure= 110 kPa Residence Time = 1 hour Volume = 120 m^3 MOC= Carbon Steel with Stainless Clad
D-121	Acid Presteamer	Height = 5.8 m Diameter = 2.9 m Temperature = 238 °C Pressure = 3300 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-132	Acid Filtration Knockout Drum	Diameter = 4.0 m Height = 7.9 m Temperature = 93 °C Pressure = 55 kPa MOC = Carbon Steel with Nickel Cladding
D-161	Gypsum Filtration Knockout Drum	Diameter = 3.1 m Height = 6.1 m Temperature = 93 °C Pressure = 55 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
G = Compressors		
G-135	Acid Filtration Air Compressor	Duty = 13 kW Fluid = Air Volumetric Flow Rate = 2.9 liters/min Temperature In = 93 °C Temperature Out = 360 °C Pressure In = 34 kPa Pressure Out = 140 kPa MOC = Carbon Steel
G-164	Gypsum Filtration Air Compressor	Duty = 4.4 kW MOC = Carbon Steel Fluid= Air Volumetric Flow Rate = 1.1 liters/minute Temperature In = 93°C Temperature Out = 360 °C Pressure In = 35 kPa Pressure Out = 140 kPa

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
H = Filters		
H-131 A-G	Acid Rotary Vacuum Filter	Duty = 1.5 kW Volume = 15 m^3 Disks = 6 Disk Size = 2.9 MOC = Nickel Alloy
H-161 A/B	Gypsum Rotary Vacuum Filter	Duty = 0.74 kW Volume = 5.1 m^3 Disks = 2 Disk Size = 1.9 m MOC = Carbon Steel
J = Conveyors		
J-108	NaOH Feed Belt Conveyor	Power = 0.76 kW Length= 150 m Width= 0.61 m Temperature = 25 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-111 A-C/D	Deacetylation Screw Conveyor	Power = 3.7 kW Temperature = 80° C Pressure = 100 kPa Length = 15 m Diameter = 0.45 m MOC = Stainless Steel
J-112 A-C/D	Pre-Deacetylation Corn Stover Belt Conveyor	Power = 2.2 kW Length= 15 m Width= 0.45 m Temperature = 80 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-114 A-C/D	Post-Deacetylation Corn Stover Belt	Power = 2.2 kW Length= 15 m Width= 0.45 m Temperature = 80 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-122 A/B	Pre-Hydrolysis Corn Stover Belt Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 82 °C Pressure = 100 kPa MOC = Elastomer (rubber) material

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
J = Conveyors		
J-134 A-G	Acid Filtration Corn Stover Belt Conveyor	Power = 1.5 kW Length= 150 m Width= 0.61 m Temperature = 49 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-152 A/B	Calcium Carbonate Feed Conveyor	Power = 1.5 kW Length= 150 m Width= 0.61 m Temperature = 25 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-153 A/B	Calcium Carbonate Screw Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 25 °C Pressure = 100 kPa MOC = Stainless Steel
J-162 A/B	Gypsum Filtration Corn Stover Belt Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 49 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
L = Pumps		
L-107 A/B	NaOH Feed Pump	Power = 12 kW Volumetric Flow Rate = 2,000 liters/min Fluid= 0.4 wt% NaOH solution Inlet Pressure = 83 kPa Outlet Pressure = 310 kPa Temperature = 25 °C MOC = Stainless Steel
L-113 A-C/D	Pre-Deacetylation Corn Stover Pump	Power = 0.74 kW Volumetric Flow Rate = 76 liters/minute Fluid = Black Liquor Recycle Inlet Pressure = 100 kPa Outlet Pressure = 170 kPa Temperature = 80 °C MOC = Cast Iron

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
L = Pumps		
L-116 A/B	Black Liquor Pump	Power = 3.7 kW Volumetric Flow Rate = 2,100 liters/minute Fluid = Black Liquor Inlet Pressure = 69 kPa Outlet Pressure = 138 kPa Temperature = 80 °C MOC = Cast Iron
L-133 A/B	Acid Filtration Pump	Power = 19 kW Volumetric Flow Rate = 3,100 liters/min. Fluid = Acid Hydrolysate (0.4 wt% H_2SO_4) Inlet Pressure = 34 kPa Outlet Pressure = 275 kPa Temperature = 93 °C MOC = Nickel Alloy
L-151 A-B/C	Acid Phase Separation Slurry Pump	Power = 11 kW Volumetric Flow Rate = 870 liters/minute Fluid = Neutralized Acid Slurry Inlet Pressure = 240 kPa Outlet Pressure = 380 kPa Temperature = 99 °C MOC = Cast Iron
L-165 A/B	Gypsum Filtration Pump	Power = 46 kW Volumetric Flow Rate = 8,300 liters/minute Fluid = Neutralized Acid Hydrolysate Inlet Pressure = 35 kPa Outlet Pressure = 255 kPa Temperature = 93 °C MOC = Cast Iron

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
M = Mixers		
M-101	Sulfuric Acid Mixer	Agitator Power = 2.1 kW Blade Length= 1.2 m Temperature= 25 °C Pressure= 310 kPa MOC= Stainless Steel
M-106	NaOH Mixer	Agitator Power = 13 kW Blade Length= 1.5 m Temperature= 25°C Pressure= 100 kPa MOC= Carbon Steel
M-115 A-C/D	Deacetylation Mixer	Agitator Power = 96 kW Blade Length= 1.5 m Temperature= 80°C Pressure= 100 kPa MOC= Carbon Steel
R = Reactor		
R-120	Acid Hydrolysis Reactor	Length = 10 m Diameter = 1.0 m Temperature = 160 °C Pressure= 560 kPa Duty = 29 kW MOC = Carbon Steel with Nickel Clad
R-150	Neutralization Reactor	Height = 5.8 m Diameter = 11 m Residence Time = 5 minutes Temperature = 99 °C Pressure = 240 kPa MOC = Carbon Steel with Nickel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
D = Pressure Vessels		
D-221	Enzymatic Filtration Knockout Drum	Diameter = 3.1 m Height = 6.7 m Temperature= 29°C Pressure= 55 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-230	Hydrolysate Mixing Drum	Diameter = 2.5 m Height = 4.9 m Temperature = 37 °C Pressure = 220 kPa MOC = Stainless Steel
E = Heat Exchangers		
E-212 A-F/G/H	Batch Hydrolysis Recycle Heat Exchanger	Duty = 0.057 GJ/hr Transfer Area = 1.7 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 47 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Slurry Mixture Inlet Temperature = 51 °C Outlet Temperature = 47 °C Max. Pressure = 500 kPa MOC = Carbon Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
F = Tanks		
F-205	Pre-Batch Hold Up	Diameter = 9.1 m Height = 18 m Temperature = 49 °C Pressure = 100 kPa MOC = Carbon Steel
F-213	Post-Batch Hold Up	Diameter = 9.1 m Height = 18 m Temperature = 49 °C Pressure = 100 kPa MOC = Carbon Steel
G = Compressors		
G-223	Enzymatic Filtration Air Compressor	Duty = 13 kW MOC = Carbon Steel Fluid = Air Volumetric Flow Rate = 3.2 liters/minute Temperature In = 29 °C Temperature Out = 254 °C Pressure In = 35 kPa Pressure Out = 140 kPa
H = Filters		
H-220 A-C	Enzymatic Rotary Vacuum Filter	Duty = 1.5 kW Volume = 500 m^2 Disks = 6 Disk Size = 1.9 m MOC = Carbon Steel with Stainless Steel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
J = Conveyors		
J-202 A-H	Pre-Enzymatic Screw Conveyor	Power = 2.2 kW Temperature = 49° F Pressure = 300 kPa Length = 15 m Diameter = 0.45 m MOC = Stainless Steel
J-204 A/B	Corn Stover Belt Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 49 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-224 A-C	Lignin Rich Belt Conveyor	Power = 1.5 kW Temperature = 29° C Length= 150 m Width= 0.61 m Pressure = 100 kPa MOC = Elastomer (rubber) material
L = Pumps		
L-201 A/B	Enzymatic Slurry Pump	Power = 33 kW Volumetric Flow Rate = 2,600 liters/minute Fluid = Partially Hydrolyzed Corn Stover Inlet Pressure = 100 kPa Outlet Pressure = 610 kPa Temperature = 49 °C Process Fluid = Neutralized Acid Slurry MOC = Cast Iron
L-211 A-F/G/H	Batch Hydrolysis Recycle Pump	Power = 1.5 kW Volumetric Flow Rate = 370 liters/minute Fluid= Hydrolyzed Corn Stover Recycle Inlet Pressure = 340 kPa Outlet Pressure = 500 kPa Temperature = 51 °C MOC = Cast Iron

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
L = Pumps		
L-222 A/B	Enzymatic Filtration Pump	Power = 4.5 kW Volumetric Flow Rate = 1,100 liters/min. Fluid = Enzymatic Hydrolysate Inlet Pressure = 35 kPa Outlet Pressure = 360 kPa Temperature = 29°C MOC = Cast Iron
L-225 A/B	Enzymatic Filtration Feed Slurry Pump	Power = 6.7 kW Volumetric Flow Rate = 3,500 liters/min. Fluid = Hydrolyzed Corn Stover Inlet Pressure = 100 kPa Outlet Pressure = 340 kPa Temperature = 49 °C MOC = Cast Iron
M = Mixers		
M-214 A-F	Batch Hydrolysis Mixer	Agitator Power = 430 kW Blade Length= 2.7 m Temperature= 49°C Pressure= 340 kPa MOC= Carbon Steel
R = Reactors		
R-200 A-H/I	Cont.inuous, High-Solids Enzymatic Reactor	Height = 17 m Diameter = 4.3 m Temperature = 49 °C Pressure = 300 kPa Residence Time = 24 hr MOC = Carbon Steel with Stainless Steel Cladding
R-210 A-F/G	Batch Enzymatic Reactor	Height = 16 m Diameter = 8.2 m Temperature = 49 °C Pressure = 340 kPa Residence Time = 60 hr MOC = Carbon Steel with Stainless Steel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	**Equipment Name**	**Equipment Specifications**
Process Area 3		
D = Pressure Vessels		
D-310	Furfural Flash Drum	Height = 3.7 m Diameter = 7.4 m Temperature = 35 °C Pressure = 172 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-320	1st Stage Furfural Extraction Mixing Drum	Height = 7.4 m Diameter = 7.3 m Temperature = 35°C Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-321	1st Stage Furfural Extraction Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-325	2nd Stage Furfural Extraction Mixing Drum	Height = 7.4 m Diameter = 7.3 m Temperature = 35°C Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-326	2nd Stage Furfural Extraction Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 110 kPa MOC = Carbon Steel with Stainless Clad
D-330	3rd Stage Furfural Extraction Mixing Drum	Height = 7.4 m Diameter = 7.3 m Temperature = 35°C Pressure = 170 kPa MOC = Carbon Steel with Stainless Clad
D-330	3rd Stage Furfural Extraction Mixing Drum	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 110 kPa MOC = Carbon Steel with Stainless Clad

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
D = Pressure Vessels		
D-340	Furfural Extraction Hold Up Drum	Height = 0.79 m Diameter = 0.39 m Temperature = 35 °C Pressure = 60 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-350	Toluene Recovery Column	Diameter= 0.23 m Height = 12 m Number of Trays = 29 Temperature = 160°C Pressure = 217 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-352	Toluene Recovery Column Reflux Drum	Height = 0.28 m Length= 0.85 m Temperature = 110 °C Pressure = 159 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-360	Toluene Recycle Drum	Height = 0.94 m Diameter = 0.3 m Temperature = 110 °C Pressure = 117 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-372	1st Compressor Inter stage Knockout Drum	Height = 3.6 m Diameter = 7.3 m Temperature = 35 °C Pressure = 270 kPa MOC = Carbon Steel
D-374	2nd Compressor Inter stage Knockout Drum	Height = 2.8 m Diameter = 5.5 m Temperature = 93 °C Pressure = 760 kPa MOC = Carbon Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
E = Heat Exchangers		
E-302 A/B	Catalytic Decomposition Preheater	Duty = 94 GJ/hr Area = 710 m^2 Heat Transfer Coefficient = 750 W/(m^2-°C) **Shell Side** Medium-High Pressure Steam Inlet Temperature = 238°C Outlet Temperature = 238°C Max Pressure = 3300 kPa MOC = Carbon Steel **Tube Side** Concentrated Hydrolysate Inlet Temperature = 27°C Outlet Temperature = 200 °C Max. Pressure = 2100 kPa MOC = Carbon Steel
E-311 A-F/G/H	Pre-Furfural Extraction Cooler	Duty = 110 GJ/hr Area = 710 m^2 Heat Transfer Coefficient = 965 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30°C Outlet Temperature = 175°C Max Pressure = 920 kPa MOC = Carbon Steel **Tube Side** Sugar Decomposition Products Inlet Temperature = 200°C Outlet Temperature = 35°C Max. Pressure = 1800 kPa MOC = Carbon Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
E = Heat Exchangers		
E-351 A/B	Toluene Recovery Condenser	Duty = 0.20 GJ/hr Transfer Area = 3.9 m^2 Heat Transfer Coefficient = 750 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 105°C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Toluene Inlet Temperature = 116 °C Outlet Temperature = 110 °C Max. Pressure = 179 kPa MOC = Carbon Steel
E-354 A/B	Toluene Recovery Reboiler	Duty = 0.30 GJ/hr Transfer Area = 2.4 m^2 Heat Transfer Coefficient = 800 W/(m^2-°C) **Shell Side** Furfural Inlet Temperature = 160 °C Outlet Temperature = 160 °C Max Pressure = 217 kPa MOC = Carbon Steel with Stainless Steel Cladding **Tube Side** Moderate Pressure Steam Inlet Temperature = 204 °C Outlet Temperature = 204 °C Max. Pressure = 1700 kPa MOC = Carbon Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
E = Heat Exchangers		
E-362 A/B	Toluene Cooler	Duty = 0.044 MJ/ hr Transfer Area = 9.3 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 105 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Recovered Toluene Inlet Temperature = 110 °C Outlet Temperature = 35 °C Max. Pressure = 207 kPa MOC = Carbon Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
E = Heat Exchangers		
E-371 A-B/C	1st Air Compressor Inter Stage Cooler	Duty = 8.2 GJ/hr Area = 470 m^2 Heat Transfer Coefficient = 220 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30 °C Outlet Temperature = 148 °C Max Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Air Inlet Temperature = 184 °C Outlet Temperature = 35 °C Max. Pressure = 280 kPa MOC = Carbon Steel
E-373 A/B	2nd Air Compressor Inter Stage Cooler	Duty = 4.9 GJ/hr Area = 79 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Boiler feed water Inlet Temperature = 30 °C Outlet Temperature = 148 °C Max Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Air Inlet Temperature = 172 °C Outlet Temperature = 93 °C Max. Pressure = 780 kPa MOC = Carbon Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
G = Compressors		
G-370	Three-Stage Air Compressor	Duty = 6,900 kW MOC = Carbon Steel Fluid = Air Volumetric Flow Rate = 69,000 liters/min. Temperature In = 24 °C Temperature Out = 260 °C Pressure In = 103 kPa Pressure Out = 1800 kPa
L = Pumps		
L-301 A/B	Pre-Catalytic Decomposition Pump	Power = 140 kW Volumetric Flow Rate = 4,600 liters/minute Fluid= Concentrated Hydrolysate Inlet Pressure = 62 kPa Outlet Pressure = 2100 kPa Temperature = 27 °C MOC = Cast Iron
L-323 A/B	1st Stage Furfural Pump	Power = 4.8 kW Volumetric Flow Rate = 2,700 liters/miN. Fluid= Organic Acid Mix Inlet Pressure = 100 kPa Outlet Pressure = 170 kPa Temperature = 35 °C MOC = Copper Alloy
L-328 A/B	2nd Stage Furfural Pump	Power = 4.8 kW Volumetric Flow Rate = 2,700 liters/minute Fluid= Organic Acid Mix Inlet Pressure = 100 kPa Outlet Pressure = 170 kPa Temperature = 35 °C MOC = Copper Alloy
L-333 A/B	3rd Stage Furfural Pump	Power = 8.7 kW Volumetric Flow Rate = 2,700 liters/minute Fluid= Organic Acid Mix Inlet Pressure = 100 kPa Outlet Pressure = 228 kPa Temperature = 35 °C MOC = Copper Alloy

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
L = Pumps		
L-341 A/B	Furfural Extractant Pump	Power = 0.74 kW Volumetric Flow Rate = 9.5 liters/minute Fluid= Combined Furfural Extractant Inlet Pressure = 35 kPa Outlet Pressure = 220 kPa Temperature = 35 °C MOC = Cast Iron
L-353 A/B	Toluene Recovery Reflux Pump	Power = 0.74 kW Volumetric Flow Rate = 13 liters/minute Fluid= Toluene Recycle Inlet Pressure = 150 kPa Outlet Pressure = 250 kPa Temperature = 110 °C MOC = Cast Iron
L-361 A/B	Toluene Pump	Power = 0.74 kW Volumetric Flow Rate = 6.8 liters/minute Fluid= Toluene Inlet Pressure = 117 kPa Outlet Pressure = 207 kPa Temperature = 110 °C MOC = Cast Iron
M = Mixers		
M-322	1st Stage Furfural Mixer	Agitator Power = 39 kW Blade Length= 1.2 m Temperature= 35 °C Pressure= 170 kPa MOC = Stainless Steel
M-327	2nd Stage Furfural Mixer	Agitator Power = 39 kW Blade Length= 1.2 m Temperature= 35 °C Pressure= 170 kPa MOC = Stainless Steel
M-332	3rd Stage Furfural Mixer	Agitator Power = 39 kW Blade Length= 1.2 m Temperature= 35 °C Pressure= 170 kPa MOC = Stainless Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
R = Reactors		
R-300 A-P/Q/R	Sn+2 Beta Zeolite Packed Bed Reactor	Diameter = 1.8 m Height = 15 m Temperature = 200 °C Pressure = 2,070 kPa Residence Time = 2hr Quantity of Catalyst = 4,100 kg MOC = Carbon Steel with Stainless Steel Cladding
Process Area 4		
D = Pressure Vessels		
D-400	1st Stage TOA Mixing Drum	Height = 8.3 m Diameter = 4.1 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-401	1st Stage TOA Settler	Height = 4.6 m Diameter = 2.3 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-405	2nd Stage TOA Mixing Drum	Height = 8.3 m Diameter = 4.1 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-406	2nd Stage TOA Settler	Height = 4.6 m Diameter = 2.3 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
D = Pressure Vessels		
D-420	TOA Acid Extractant Hold Up Drum	Height = 4.4 m Diameter = 2.2 m Pressure = 110 kPa Temperature = 35 °C MOC = Carbon Steel with Stainless Steel Cladding
D-430	Volatile Acid Recovery Column	Rectifying Diameter = 0.58 m Stripping Diameter = 3.9 m Height = 13 m Number of Trays = 18 Temperature = 202°C Pressure = 410 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-432	Volatile Acid Column Reflux Drum	Height = 1.5 m Diameter = 0.73 m Pressure = 345 kPa Temperature = 99 °C MOC = Carbon Steel with Stainless Steel Cladding
D-440 A-D	Acid/l-Octanol Pressure Swing Adsorber	Total Height = 8.9 m Diameter = 3.5 m Packing height = 7.1 m Type of Molecular Sieves = 13x Temperature = 99 °C Minimum Pressure = 55 kPa Maximum Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
D = Pressure Vessels		
D-450	Formic/Acetic Column	Diameter = 1.3 m Height = 43 m Number of Trays = 70 Temperature = 121°C Pressure = 400 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-452	Formic/Acetic Acid Column Reflux Drum	Height = 2.2 m Diameter = 1.1 m Pressure = 340 kPa Temperature = 99°C MOC = Carbon Steel with Stainless Steel Cladding
E = Heat Exchangers		
E-431 A/B	Volatile Acid Column Condenser	Duty = 1.5 GJ/hr Area = 19 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 94 °C Max Pressure = 365 kPa MOC = Carbon Steel **Tube Side** 1-Octonal Saturated Volatile Acids Inlet Temperature = 99 °C Outlet Temperature = 99 °C Max. Pressure = 365 kPa MOC = Copper

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	**Equipment Name**	**Equipment Specifications**
Process Area 4		
E = Heat Exchangers		
E-434 A/B	Volatile Acid Column Reboiler	Duty = 56 GJ/hr Area = 590 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Moderate Pressure Steam Inlet Temperature = 239 °C Outlet Temperature = 239 °C Max. Pressure = 3300 kPa MOC = Carbon Steel **Tube Side** Heavy Acid Extractant Inlet Temperature = 202 °C Outlet Temperature = 202°C Max Pressure = 410 kPa MOC = Nickel
E-435 A-C/D	Heavy Acid Extractant Cooler	Duty = 65 GJ/hr Area = 740 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30 °C Outlet Temperature = 148 °C Max. Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Liquid Acid Mixture Inlet Temperature = 202 °C Outlet Temperature = 35 °C Max Pressure = 370 kPa MOC = Nickel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
E = Heat Exchangers		
E-442 A/B	Volatile Acid Cooler	Duty = 0.40 GJ/hr Area = 81 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Tube Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 95 °C Max. Pressure = 310 kPa MOC = Carbon Steel **Shell Side** Acetic and Formic Acids Inlet Temperature = 99 °C Outlet Temperature = 35 °C Max Pressure = 55 kPa MOC = Teflon
E-443 A/B	Recovered l-Octanol Cooler	Duty = 0.17 GJ/hr Area = 35 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 95 °C Max. Pressure = 310 kPa MOC = Carbon Steel **Tube Side** 1-Octonal Inlet Temperature = 99 °C Outlet Temperature = 35 °C Max Pressure = 160 kPa MOC = Copper

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
E = Heat Exchangers		
E-451 A/B	Formic/Acetic Column Condenser	Duty = 6.0 GJ/hr Area = 91 m^2 Heat Transfer Coeff. = 350 W/(m^2-K) **Shell Side** Cooling Water Inlet Temperature = 30°C Outlet Temperature = 95 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Formic Acid Inlet Temperature = 99 °C Outlet Temperature = 99 °C Max. Pressure = 360 kPa MOC = Copper
E-454 A/B	Formic/Acetic Column Reboiler	Duty = 3.5 GJ/hr Area = 27 m^2 Heat Transfer Coeff. = 350 W/(m^2-°C) **Shell Side** Low-Pressure Steam Inlet Temperature = 148 °C Outlet Temperature = 142 °C Max. Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Acetic Acid Inlet Temperature = 121 °C Outlet Temperature = 121 °C Max Pressure = 403 kPa MOC =Nickel
L = Pumps		
L-403 A/B	1st Stage TOA Pump	Power = 8.3 kW Volumetric Flow Rate = 2,800 liters/minute Fluid= Organic Acid Mix Inlet Pressure = 220 kPa Outlet Pressure = 350 kPa Temperature = 95 °C MOC = Copper Alloy

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
L = Pumps		
L-421 A/B	TOA Hold Up Pump	Power = 13 kW Volumetric Flow Rate = 1,400 liters/minute Inlet Pressure = 89 kPa Fluid= TOA Acid Extractant Outlet Pressure = 450 kPa Temperature = 35 °C MOC = Cast Iron
L-433 A/B	Volatile Acid Column Reflux Pump	Power = 0.74 kW Volumetric Flow Rate = 7.8 liters/minute Fluid= Formic and Acetic Acid Inlet Pressure in = 338 kPa Outlet Pressure out = 796 kPa Temperature = 99 °C MOC = Copper Alloy
L-441 A/B	Pressure Swing Pump	Power = 0.74 kW Volumetric Flow Rate = 34 liters/minute Fluid= Acetic and Formic Acid Inlet Pressure = 40 kPa Outlet Pressure = 440 kPa Temperature = 35 °C MOC = Copper Alloy
L-453 A/B	Formic/Acetic Reflux Pump	Power = 0.74 kW Volumetric Flow Rate = 42 liters/minute Fluid= Formic Acid Inlet Pressure = 330 kPa Outlet Pressure = 470 kPa Temperature = 99 °C MOC = Copper Alloy

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
M = Mixers		
M-402	1st Stage TOA Mixer	Duty = 54 kW Diameter = 1.4 m MOC = Stainless Steel
M-407	2nd Stage TOA Mixer	Duty = 54 kW Diameter = 1.4 m MOC = Stainless Steel
Process Area 5		
D = Pressure Vessels		
D-500	1st Stage TMA Mixing Drum	Height = 6.9 m Diameter = 3.5 m Temperature = 35 °C Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-501	1st Stage TMA Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-505	2nd Stage TMA Mixing Drum	Height = 6.9 m Diameter = 3.5 m Temperature = 35 °C Pressure = 280 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-506	2nd Stage TMA Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 250 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
D = Pressure Vessels		
D-510	3rd Stage TMA Mixing Drum	Height = 6.9 m Diameter = 3.5 m Temperature = 35 °C Pressure = 210 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-511	3rd Stage TMA Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-515	4th Stage TMA Mixing Drum	Height = 6.9 m Diameter = 3.5 m Temperature = 35 °C Pressure = 210 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-516	4th Stage TMA Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-520	TMA Acid Extractant Hold Up Drum	Height = 2.3 m Diameter = 4.7 m Pressure = 120 kPa Temperature = 35 °C MOC = Carbon Steel with Stainless Steel Cladding
D-530	TOA Acid Extactant Hold up Drum	Height = 9.6 m Diameter = 5.9 m Pressure = 100 kPa Temperature = 35 °C MOC = Carbon Steel with Stainless Steel Cladding

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
D = Pressure Vessels		
D-540	Back Extractant Solvent Recovery Column	Diameter = 1.3 m Height = 7.5 m Number of Trays = 10 Temperature = 224 °C Pressure = 218 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-542	Back Extractant Recovery Reflux Drum	Height = 2.1 m Diameter = 1.0 m Pressure = 170 kPa Temperature = 54 °C MOC = Carbon Steel with Stainless Steel Cladding
D-550	TMA Acid Extractant Hold Up Drum	Height = 3.2 m Diameter = 1.6 m Pressure = 120 kPa Temperature = 54 °C MOC = Carbon Steel with Stainless Steel Cladding
D-560	Lactic/Levulinic Acid Column	Diameter = 0.83 m Height = 26 m Number of Trays = 40 Temperature = 395 °C Pressure = 395 kPa MOC = Carbon Steel
D-562	Lactic/Levulinic Reflux Drum	Height = 0.98 m Diameter = 0.49 m Pressure = 340 kPa Temperature = 232 °C MOC = Carbon Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
E = Heat Exchangers		
E-541 A/B	Back Extractant Recovery Condenser	Duty = 14 GJ/hr Area = 270 m^2 Heat Transfer Coefficient = 850 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 290 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** TMA and Water Inlet Temperature = 54 °C Outlet Temperature = 54 °C Max. Pressure = 190 kPa MOC = Carbon Steel
E-544 A-C/D	Back Extractant Recovery Reboiler	Duty = 17 GJ/hr Area = 120 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Medium-High Pressure Steam Inlet Temperature = 238 °C Outlet Temperature = 232 °C Max. Pressure = 3310 kPa MOC = Carbon Steel **Tube Side** Lactic and Levulinic Acid Inlet Temperature = 224 °C Outlet Temperature = 224 °C Max Pressure = 218 kPa MOC = Nickel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
E = Heat Exchangers		
E-552 A/B	TMA/Water Cooler	Duty = 0.91 GJ/hr Area = 160 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 49 °C Max. Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Trimethylamine and Water Inlet Temperature = 54 °C Outlet Temperature = 35 °C Max Pressure = 260 kPa MOC = Carbon Steel
E-561 A/B	Lactic/Levulinic Condenser	Duty = 3.0 GJ/hr Area = 110 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30 °C Outlet Temperature = 175 °C Max Pressure = 930 kPa MOC = Carbon Steel **Tube Side** Lactic Acid Inlet Temperature = 232 °C Outlet Temperature = 232 °C Max. Pressure = 360 kPa MOC = Nickel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
E = Heat Exchangers		
E-564 A/B	Lactic/Levulinic Reboiler	Duty = 3.6 GJ/hr Area = 16 m^2 Heat Transfer Coeff. = 350 W/(m^2-°C) **Shell Side** High Pressure Steam Inlet Temperature = 400 °C Outlet Temperature = 283 °C Max. Pressure = 4500 kPa MOC = Carbon Steel **Tube Side** Levulinic Acid Inlet Temperature = 278 °C Outlet Temperature = 278 °C Max Pressure = 395 kPa MOC = Nickel
L = Pumps		
L-513 A/B	3rd Stage TMA Pump	Power = 4.2 kW Volumetric Flow Rate = 2,400 liters/min. Fluid= Mixed Acids Inlet Pressure = 145 kPa Outlet Pressure = 210 kPa Temperature = 35 °C MOC = Cast Iron
L-518 A/B	4th Stage TMA Pump	Power = 4.2 kW Volumetric Flow Rate = 2,400 liters/minute Fluid= Mixed Acids Inlet Pressure = 140 kPa Outlet Pressure = 2,000 kPa Temperature = 35 °C MOC = Cast Iron
L-521 A/B	TMA Hold Up Drum Pump	Power = 13 kW Volumetric Flow Rate = 4,200 liters/minute Fluid= TMA Back Extractant Inlet Pressure = 120 kPa Outlet Pressure = 230 kPa Temperature = 35 °C MOC = Stainless Steel

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
L = Pumps		
L-531 A/B	TOA Hold Up Drum Pump	Power = 8.0 kW Volumetric Flow Rate = 2,900 liters/minute Fluid= TOA/ 1-Octanol Solution Inlet Pressure = 100 kPa Outlet Pressure = 210 kPa Temperature = 35 °C MOC = Cast Iron
L-543 A/B	Back Extractant Recovery Reflux Pump	Power = 1.7 kW Volumetric Flow Rate = 1,200 liters/minute Fluid= TMA/Water Recycle Inlet Pressure = 160 kPa Outlet Pressure = 240 kPa Temperature = 54 °C MOC = Stainless Steel
L-551 A/B	TMA Hold Up Drum Pump	Duty = 11 kW Volumetric Flow Rate = 3,100 liters/minute Fluid= TMA/ Water Solution Inlet Pressure = 120 kPa Outlet Pressure = 260 kPa Temperature = 54 °C MOC = Stainless Steel
L-563 A/B	Lactic/Levulinic Reflux Pump	Duty = 0.74 kW Volumetric Flow Rate = 30 liters/minute Fluid= Lactic Acid Recycle Inlet Pressure = 330 kPa Outlet Pressure = 500 kPa Temperature = 232 °C MOC = Cast Iron
L-565 A/B	Lactic/Levulinic Feed Pump	Duty = 0.74 kW Volumetric Flow Rate = 57 liters/minute Fluid= Lactic and Levulinic Acid Inlet Pressure = 180 kPa Outlet Pressure = 430 kPa Temperature = 224 °C MOC = Cast Iron

Table 38. Major Equipment for the Base Sn-Beta Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
M = Mixers		
M-502	1st Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel
M-507	2nd Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel
M-512	3rd Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel
M-517	4th Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel

APPENDIX H – MAJOR EQUIPENT LIST FOR SN-BETA + $CaSO_4$ PROCESS

Table 39. Major Equipment List for the Sn-Beta + CaSO4 Process

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
D = Pressure Vessels		
D-100	Sulfuric Acid Mixing Vessel	Height = 2.0 m Diameter = 1.0 m Temperature = 25 °C Pressure = 310 kPa Volume = 1.6 m^3 Residence Time = 5 minutes MOC = Carbon Steel with Nickel Clad
D-105	NaOH Mixing Vessel	Height = 2.1 m Diameter = 4.2 m Temperature = 25 °C Pressure = 100 kPa Volume = 14.5 m^3 Residence Time = 5 minutes MOC = Carbon Steel
D-110 A-C/D	Deacetylation Vessel	Height = 8.5 m Diameter = 4.3 m Temperature = 80° C Pressure= 110 kPa Residence Time = 1 hour Volume = 120 m^3 MOC= Carbon Steel with Stainless Steel Cladding
D-121	Acid Presteamer	Height = 5.8 m Diameter = 2.9 m Temperature = 238 °C Pressure = 3300 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-132	Acid Filtration Knockout Drum	Diameter = 4.0 m Height = 7.9 m Temperature = 93 °C Pressure = 55 kPa MOC = Carbon Steel with Nickel Cladding
D-161	Gypsum Filtration Knockout Drum	Diameter = 3.1 m Height = 6.1 m Temperature = 93 °C Pressure = 55 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	**Equipment Name**	**Equipment Specifications**
Process Area 1		
G = Compressors		
G-135	Acid Filtration Air Compressor	Duty = 13 kW Fluid = Air Volumetric Flow Rate = 2.9 liters/min Temperature In = 93 °C Temperature Out = 360 °C Pressure In = 34 kPa Pressure Out = 140 kPa MOC = Carbon Steel
G-164	Gypsum Filtration Air Compressor	Duty = 4.4 kW MOC = Carbon Steel Fluid= Air Volumetric Flow Rate = 1.1 liters/minute Temperature In = 93°C Temperature Out = 360 °C Pressure In = 35 kPa Pressure Out = 140 kPa

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
H = Filters		
H-131 A-G	Acid Rotary Vacuum Filter	Duty = 1.5 kW Volume = 15 m^3 Disks = 6 Disk Size = 2.9 MOC = Nickel Alloy
H-161 A/B	Gypsum Rotary Vacuum Filter	Duty = 0.74 kW Volume = 5.1 m^3 Disks = 2 Disk Size = 1.9 m MOC = Carbon Steel
J = Conveyors		
J-108	NaOH Feed Belt Conveyor	Power = 0.76 kW Length= 150 m Width= 0.61 m Temperature = 25 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-111 A-C/D	Deacetylation Screw Conveyor	Power = 3.7 kW Temperature = 80° C Pressure = 100 kPa Length = 15 m Diameter = 0.45 m MOC = Stainless Steel
J-112 A-C/D	Pre-Deacetylation Corn Stover Belt Conveyor	Power = 2.2 kW Length= 15 m Width= 0.45 m Temperature = 80 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-114 A-C/D	Post-Deacetylation Corn Stover Belt	Power = 2.2 kW Length= 15 m Width= 0.45 m Temperature = 80 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-122 A/B	Pre-Hydrolysis Corn Stover Belt Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 82 °C Pressure = 100 kPa MOC = Elastomer (rubber) material

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
J = Conveyors		
J-134 A-G	Acid Filtration Corn Stover Belt Conveyor	Power = 1.5 kW Length= 150 m Width= 0.61 m Temperature = 49 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-152 A/B	Calcium Carbonate Feed Conveyor	Power = 1.5 kW Length= 150 m Width= 0.61 m Temperature = 25 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-153 A/B	Calcium Carbonate Screw Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 25 °C Pressure = 100 kPa MOC = Stainless Steel
J-162 A/B	Gypsum Filtration Corn Stover Belt Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 49 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
L = Pumps		
L-107 A/B	NaOH Feed Pump	Power = 12 kW Volumetric Flow Rate = 2,000 liters/min Fluid= 0.4 wt% NaOH solution Inlet Pressure = 83 kPa Outlet Pressure = 310 kPa Temperature = 25 °C MOC = Stainless Steel
L-113 A-C/D	Pre-Deacetylation Corn Stover Pump	Power = 0.74 kW Volumetric Flow Rate = 76 liters/minute Fluid = Black Liquor Recycle Inlet Pressure = 100 kPa Outlet Pressure = 170 kPa Temperature = 80 °C MOC = Cast Iron

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 1		
L = Pumps		
L-116 A/B	Black Liquor Pump	Power = 3.7 kW Volumetric Flow Rate = 2,100 liters/minute Fluid = Black Liquor Inlet Pressure = 69 kPa Outlet Pressure = 138 kPa Temperature = 80 °C MOC = Cast Iron
L-133 A/B	Acid Filtration Pump	Power = 19 kW Volumetric Flow Rate = 3,100 liters/min. Fluid = Acid Hydrolysate (0.4 wt% H_2SO_4) Inlet Pressure = 34 kPa Outlet Pressure = 275 kPa Temperature = 93 °C MOC = Nickel Alloy
L-151 A-B/C	Acid Phase Separation Slurry Pump	Power = 11 kW Volumetric Flow Rate = 870 liters/minute Fluid = Neutralized Acid Slurry Inlet Pressure = 240 kPa Outlet Pressure = 380 kPa Temperature = 99 °C MOC = Cast Iron
L-165 A/B	Gypsum Filtration Pump	Power = 46 kW Volumetric Flow Rate = 8,300 liters/minute Fluid = Neutralized Acid Hydrolysate Inlet Pressure = 35 kPa Outlet Pressure = 255 kPa Temperature = 93 °C MOC = Cast Iron

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	**Equipment Name**	**Equipment Specifications**
Process Area 1		
M = Mixers		
M-101	Sulfuric Acid Mixer	Agitator Power = 2.1 kW Blade Length= 1.2 m Temperature= 25 °C Pressure= 310 kPa MOC= Stainless Steel
M-106	NaOH Mixer	Agitator Power = 13 kW Blade Length= 1.5 m Temperature= 25°C Pressure= 100 kPa MOC= Carbon Steel
M-115 A-C/D	Deacetylation Mixer	Agitator Power = 96 kW Blade Length= 1.5 m Temperature= 80°C Pressure= 100 kPa MOC= Carbon Steel
R = Reactor		
R-120	Acid Hydrolysis Reactor	Length = 10 m Diameter = 1.0 m Temperature = 160 °C Pressure= 560 kPa Duty = 29 kW MOC = Carbon Steel with Nickel Clad
R-150	Neutralization Reactor	Height = 5.8 m Diameter = 11 m Residence Time = 5 minutes Temperature = 99 °C Pressure = 240 kPa MOC = Carbon Steel with Nickel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
D = Pressure Vessels		
D-221	Enzymatic Filtration Knockout Drum	Diameter = 3.1 m Height = 6.7 m Temperature= 29°C Pressure= 55 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-230	Hydrolysate Mixing Drum	Diameter = 2.5 m Height = 4.9 m Temperature = 37 °C Pressure = 220 kPa MOC = Stainless Steel
E = Heat Exchangers		
E-212 A-F/G/H	Batch Hydrolysis Recycle Heat Exchanger	Duty = 0.057 GJ/hr Transfer Area = 1.7 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 47 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Slurry Mixture Inlet Temperature = 51 °C Outlet Temperature = 47 °C Max. Pressure = 500 kPa MOC = Carbon Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
F = Tanks		
F-205	Pre-Batch Hold Up	Diameter = 9.1 m Height = 18 m Temperature = 49 °C Pressure = 100 kPa MOC = Carbon Steel
F-213	Post-Batch Hold Up	Diameter = 9.1 m Height = 18 m Temperature = 49 °C Pressure = 100 kPa MOC = Carbon Steel
G = Compressors		
G-223	Enzymatic Filtration Air Compressor	Duty = 13 kW MOC = Carbon Steel Fluid = Air Volumetric Flow Rate = 3.2 liters/minute Temperature In = 29 °C Temperature Out = 254 °C Pressure In = 35 kPa Pressure Out = 140 kPa
H = Filters		
H-220 A-C	Enzymatic Rotary Vacuum Filter	Duty = 1.5 kW Volume = 500 m^2 Disks = 6 Disk Size = 1.9 m MOC = Carbon Steel with Stainless Steel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
J = Conveyors		
J-202 A-H	Pre-Enzymatic Screw Conveyor	Power = 2.2 kW Temperature = 49° F Pressure = 300 kPa Length = 15 m Diameter = 0.45 m MOC = Stainless Steel
J-204 A/B	Corn Stover Belt Conveyor	Power = 1.5 kW Length= 15 m Width= 0.61 m Temperature = 49 °C Pressure = 100 kPa MOC = Elastomer (rubber) material
J-224 A-C	Lignin Rich Belt Conveyor	Power = 1.5 kW Temperature = 29° C Length= 150 m Width= 0.61 m Pressure = 100 kPa MOC = Elastomer (rubber) material
J-231	Calcium Sulfate Feed Conveyor	Power = 1.5 kW Temperature = 25° C Length= 150 m Width= 0.61 m Pressure = 100 kPa MOC = Elastomer (rubber) material
J-232	Calcium Sulfate Screw Conveyor	Power = 1.5 kW Temperature = 25° F Pressure = 255 kPa Length = 15 m Diameter = 0.61 m MOC = Stainless Steel
L = Pumps		
L-201 A/B	Enzymatic Slurry Pump	Power = 33 kW Volumetric Flow Rate = 2,600 liters/minute Fluid = Partially Hydrolyzed Corn Stover Inlet Pressure = 100 kPa Outlet Pressure = 610 kPa Temperature = 49 °C Process Fluid = Neutralized Acid Slurry MOC = Cast Iron

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 2		
L = Pumps		
L-211 A-F/G/H	Batch Hydrolysis Recycle Pump	Power = 1.5 kW Volumetric Flow Rate = 370 liters/minute Fluid= Hydrolyzed Corn Stover Recycle Inlet Pressure = 340 kPa Outlet Pressure = 500 kPa Temperature = 51 °C MOC = Cast Iron
L-222 A/B	Enzymatic Filtration Pump	Power = 4.5 kW Volumetric Flow Rate = 1,100 liters/min. Fluid = Enzymatic Hydrolysate Inlet Pressure = 35 kPa Outlet Pressure = 360 kPa Temperature = 29°C MOC = Cast Iron
L-225 A/B	Enzymatic Filtration Feed Slurry Pump	Power = 6.7 kW Volumetric Flow Rate = 3,500 liters/min. Fluid = Hydrolyzed Corn Stover Inlet Pressure = 100 kPa Outlet Pressure = 340 kPa Temperature = 49 °C MOC = Cast Iron
M = Mixers		
M-214 A-F	Batch Hydrolysis Mixer	Agitator Power = 430 kW Blade Length= 2.7 m Temperature= 49°C Pressure= 340 kPa MOC= Carbon Steel
R = Reactors		
R-200 A-H/I	Cont.inuous, High-Solids Enzymatic Reactor	Height = 17 m Diameter = 4.3 m Temperature = 49 °C Pressure = 300 kPa Residence Time = 24 hr MOC = Carbon Steel with Stainless Clad
R-210 A-F/G	Batch Enzymatic Reactor	Height = 16 m Diameter = 8.2 m Temperature = 49 °C Pressure = 340 kPa Residence Time = 60 hr MOC = Carbon Steel with Stainless Clad

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
D = Pressure Vessels		
D-310	Furfural Flash Drum	Height = 3.7 m Diameter = 7.4 m Temperature = 35 °C Pressure = 172 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-320	1st Stage Furfural Extraction Mixing Drum	Height = 7.4 m Diameter = 7.3 m Temperature = 35°C Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-321	1st Stage Furfural Extraction Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 1100 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-325	2nd Stage Furfural Extraction Mixing Drum	Height = 7.4 m Diameter = 7.3 m Temperature = 35°C Pressure = 170 kPa MOC = Carbon Steel with Stainless Clad
D-326	2nd Stage Furfural Extraction Settler	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 1100 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-330	3rd Stage Furfural Extraction Mixing Drum	Height = 7.4 m Diameter = 7.3 m Temperature = 35°C Pressure = 170 kPa MOC = Carbon Steel with Stainless Clad
D-330	3rd Stage Furfural Extraction Mixing Drum	Height = 1.6 m Diameter = 3.2 m Temperature = 35 °C Pressure = 1100 kPa MOC = Carbon Steel with Stainless Clad

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
D = Pressure Vessels		
D-340	Furfural Extraction Hold Up Drum	Height = 0.79 m Diameter = 0.39 m Temperature = 35 °C Pressure = 60 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-350	Toluene Recovery Column	Diameter= 0.23 m Height = 12 m Number of Trays = 29 Temperature = 160°C Pressure = 217 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-352	Toluene Recovery Column Reflux Drum	Height = 0.28 m Length= 0.85 m Temperature = 110 °C Pressure = 159 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-360	Toluene Recycle Drum	Height = 0.94 m Diameter = 0.3 m Temperature = 110 °C Pressure = 117 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-372	1st Compressor Inter stage Knockout Drum	Height = 2.7 m Diameter = 5.3 m Temperature = 35 °C Pressure = 270 kPa MOC = Carbon Steel
D-374	2nd Compressor Inter stage Knockout Drum	Height = 2.0 m Diameter = 4.0 m Temperature = 93 °C Pressure = 760 kPa MOC = Carbon Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
E = Heat Exchangers		
E-302 A/B	Catalytic Decomposition Preheater	Duty = 94 GJ/hr Area = 710 m^2 Heat Transfer Coefficient = 750 W/(m^2-°C) **Shell Side** Medium-High Pressure Steam Inlet Temperature = 238°C Outlet Temperature = 238°C Max Pressure = 3300 kPa MOC = Carbon Steel **Tube Side** Concentrated Hydrolysate Inlet Temperature = 27°C Outlet Temperature = 200 °C Max. Pressure = 2100 kPa MOC = Carbon Steel
E-311 A-F/G/H	Pre-Furfural Extraction Cooler	Duty = 110 GJ/hr Area = 710 m^2 Heat Transfer Coefficient = 965 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30°C Outlet Temperature = 175°C Max Pressure = 920 kPa MOC = Carbon Steel **Tube Side** Sugar Decomposition Products Inlet Temperature = 200°C Outlet Temperature = 35°C Max. Pressure = 1800 kPa MOC = Carbon Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	**Equipment Name**	**Equipment Specifications**
Process Area 3		
E = Heat Exchangers		
E-351 A/B	Toluene Recovery Condenser	Duty = 0.20 GJ/hr Transfer Area = 3.9 m^2 Heat Transfer Coefficient = 750 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 105°C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Toluene Inlet Temperature = 116 °C Outlet Temperature = 110 °C Max. Pressure = 179 kPa MOC = Carbon Steel
E-354 A/B	Toluene Recovery Reboiler	Duty = 0.30 GJ/hr Transfer Area = 2.4 m^2 Heat Transfer Coefficient = 800 W/(m^2-°C) **Shell Side** Furfural Inlet Temperature = 160 °C Outlet Temperature = 160 °C Max Pressure = 217 kPa MOC = Carbon Steel with Stainless Steel Cladding **Tube Side** Moderate Pressure Steam Inlet Temperature = 204 °C Outlet Temperature = 204 °C Max. Pressure = 1700 kPa MOC = Carbon Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
E = Heat Exchangers		
E-362 A/B	Toluene Cooler	Duty = 0.044 MJ/ hr Transfer Area = 9.3 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 105 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Recovered Toluene Inlet Temperature = 110 °C Outlet Temperature = 35 °C Max. Pressure = 207 kPa MOC = Carbon Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
E = Heat Exchangers		
E-371 A-B/C	1st Air Compressor Inter Stage Cooler	Duty = 3.3 GJ/hr Area = 370 m^2 Heat Transfer Coefficient = 220 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30 °C Outlet Temperature = 148 °C Max Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Air Inlet Temperature = 184 °C Outlet Temperature = 35 °C Max. Pressure = 280 kPa MOC = Carbon Steel
E-373 A/B	2nd Air Compressor Inter Stage Cooler	Duty = 1.9 GJ/hr Area = 31 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Boiler feed water Inlet Temperature = 30 °C Outlet Temperature = 148 °C Max Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Air Inlet Temperature = 172 °C Outlet Temperature = 93 °C Max. Pressure = 780 kPa MOC = Carbon Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
G = Compressors		
G-370	Three-Stage Air Compressor	Duty = 2,600 kW MOC = Carbon Steel Fluid = Air Volumetric Flow Rate = 19,000 liters/min. Temperature In = 24 °C Temperature Out = 260 °C Pressure In = 103 kPa Pressure Out = 1800 kPa
L = Pumps		
L-301 A/B	Pre-Catalytic Decomposition Pump	Power = 140 kW Volumetric Flow Rate = 4,600 liters/minute Fluid= Concentrated Hydrolysate Inlet Pressure = 62 kPa Outlet Pressure = 2100 kPa Temperature = 27 °C MOC = Cast Iron
L-323 A/B	1st Stage Furfural Pump	Power = 5.0 kW Volumetric Flow Rate = 2,700 liters/min. Fluid= Organic Acid Mix Inlet Pressure = 100 kPa Outlet Pressure = 170 kPa Temperature = 35 °C MOC = Copper Alloy
L-328 A/B	2nd Stage Furfural Pump	Power = 5.0 kW Volumetric Flow Rate = 2,700 liters/minute Fluid= Organic Acid Mix Inlet Pressure = 100 kPa Outlet Pressure = 170 kPa Temperature = 35 °C MOC = Copper Alloy
L-333 A/B	3rd Stage Furfural Pump	Power = 9.0 kW Volumetric Flow Rate = 2,700 liters/minute Fluid= Organic Acid Mix Inlet Pressure = 100 kPa Outlet Pressure = 228 kPa Temperature = 35 °C MOC = Copper Alloy

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
L = Pumps		
L-341 A/B	Furfural Extractant Pump	Power = 0.74 kW Volumetric Flow Rate = 9.5 liters/minute Fluid= Combined Furfural Extractant Inlet Pressure = 35 kPa Outlet Pressure = 220 kPa Temperature = 35 °C MOC = Cast Iron
L-353 A/B	Toluene Recovery Reflux Pump	Power = 0.74 kW Volumetric Flow Rate = 13 liters/minute Fluid= Toluene Recycle Inlet Pressure = 150 kPa Outlet Pressure = 250 kPa Temperature = 110 °C MOC = Cast Iron
L-361 A/B	Toluene Pump	Power = 0.74 kW Volumetric Flow Rate = 6.8 liters/minute Fluid= Toluene Inlet Pressure = 117 kPa Outlet Pressure = 207 kPa Temperature = 110 °C MOC = Cast Iron
M = Mixers		
M-322	1st Stage Furfural Mixer	Agitator Power = 39 kW Blade Length= 1.2 m Temperature= 35 °C Pressure= 170 kPa MOC = Stainless Steel
M-327	2nd Stage Furfural Mixer	Agitator Power = 39 kW Blade Length= 1.2 m Temperature= 35 °C Pressure= 170 kPa MOC = Stainless Steel
M-332	3rd Stage Furfural Mixer	Agitator Power = 39 kW Blade Length= 1.2 m Temperature= 35 °C Pressure= 170 kPa MOC = Stainless Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 3		
R = Reactors		
R-300 A-P/Q/R	Sn+2 Beta Zeolite Packed Bed Reactor	Diameter = 1.8 m Height = 15 m Temperature = 200 °C Pressure = 2,070 kPa Residence Time = 2hr Quantity of Catalyst = 4,100 kg MOC = Carbon Steel with Stainless Steel Cladding
Process Area 4		
D = Pressure Vessels		
D-400	1st Stage TOA Mixing Drum	Height = 8.0 m Diameter = 3.9 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-401	1st Stage TOA Settler	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-404	2nd Stage TOA Mixing Drum	Height = 8.0 m Diameter = 3.9 m Temperature = 35 °C Pressure = 300 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-405	2nd Stage TOA Settler	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 265 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
D = Pressure Vessels		
D-407	3rd Stage TOA Mixing Drum	Height = 8.0 m Diameter = 3.9 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-408	3rd Stage TOA Settler	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-411	4th Stage TOA Mixing Drum	Height = 8.0 m Diameter = 3.9 m Temperature = 35 °C Pressure = 300 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-412	4th Stage TOA Settler	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 265 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-413	5th Stage TOA Mixing Drum	Height = 8.0 m Diameter = 3.9 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-414	5th Stage TOA Settler	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
D = Pressure Vessels		
D-420	TOA Acid Extractant Hold Up Drum	Height = 4.4 m Diameter = 2.2 m Pressure = 110 kPa Temperature = 35 °C MOC = Carbon Steel with Stainless Steel Cladding
D-430	Volatile Acid Recovery Column	Rectifying Diameter = 0.61 m Stripping Diameter = 4.8 m Height = 14 m Number of Trays = 18 Temperature = 202°C Pressure = 410 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-432	Volatile Acid Column Reflux Drum	Height = 1.5 m Diameter = 0.76 m Pressure = 345 kPa Temperature = 99 °C MOC = Carbon Steel with Stainless Steel Cladding
D-440 A-D	Acid/l-Octanol Pressure Swing Adsorber	Total Height = 8.9 m Diameter = 3.5 m Packing height = 7.1 m Type of Molecular Sieves = 13x Temperature = 99 °C Minimum Pressure = 55 kPa Maximum Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
D = Pressure Vessels		
D-450	Formic/Acetic Column	Diameter = 1.6 m Height = 43 m Number of Trays = 70 Temperature = 121°C Pressure = 400 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-452	Formic/Acetic Acid Column Reflux Drum	Height = 1.9 m Diameter = 0.95 m Pressure = 340 kPa Temperature = 99°C MOC = Carbon Steel with Stainless Steel Cladding
E = Heat Exchangers		
E-431 A/B	Volatile Acid Column Condenser	Duty = 1.7 GJ/hr Area = 21 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 94 °C Max Pressure = 365 psi MOC = Carbon Steel **Tube Side** 1-Octonal Saturated Volatile Acids Inlet Temperature = 99 °C Outlet Temperature = 99 °C Max. Pressure = 365 kPa MOC = Copper

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	**Equipment Name**	**Equipment Specifications**
Process Area 4		
E = Heat Exchangers		
E-434 A/B	Volatile Acid Column Reboiler	Duty = 85 GJ/hr Area = 900 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Moderate Pressure Steam Inlet Temperature = 239 °C Outlet Temperature = 239 °C Max. Pressure = 3300 kPa MOC = Carbon Steel **Tube Side** Heavy Acid Extractant Inlet Temperature = 202 °C Outlet Temperature = 202°C Max Pressure = 410 kPa MOC = Nickel
E-435 A-C/D	Heavy Acid Extractant Cooler	Duty = 99 GJ/hr Area = 840 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30 °C Outlet Temperature = 148 °C Max. Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Liquid Acid Mixture Inlet Temperature = 202 °C Outlet Temperature = 35 °C Max Pressure = 370 kPa MOC = Nickel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
E = Heat Exchangers		
E-442 A/B	Volatile Acid Cooler	Duty = 0.99 GJ/hr Area = 200 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Tube Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 95 °C Max. Pressure = 310 kPa MOC = Carbon Steel **Shell Side** Acetic and Formic Acids Inlet Temperature = 99 °C Outlet Temperature = 35 °C Max Pressure = 55 kPa MOC = Teflon
E-443 A/B	Recovered l-Octanol Cooler	Duty = 0.26 GJ/hr Area = 52 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 95 °C Max. Pressure = 310 kPa MOC = Carbon Steel **Tube Side** 1-Octonal Inlet Temperature = 99 °C Outlet Temperature = 35 °C Max Pressure = 160 kPa MOC = Copper

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
E = Heat Exchangers		
E-451 A/B	Formic/Acetic Column Condenser	Duty = 3.9 GJ/hr Area = 61 m^2 Heat Transfer Coeff. = 350 W/(m^2-K) **Shell Side** Cooling Water Inlet Temperature = 30°C Outlet Temperature = 95 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Formic Acid Inlet Temperature = 99 °C Outlet Temperature = 99 °C Max. Pressure = 360 kPa MOC = Copper
E-454 A/B	Formic/Acetic Column Reboiler	Duty = 11 GJ/hr Area = 86 m^2 Heat Transfer Coeff. = 350 W/(m^2-°C) **Shell Side** Low-Pressure Steam Inlet Temperature = 148 °C Outlet Temperature = 142 °C Max. Pressure = 450 kPa MOC = Carbon Steel **Tube Side** Acetic Acid Inlet Temperature = 121 °C Outlet Temperature = 121 °C Max Pressure = 403 kPa MOC =Nickel
L = Pumps		
L-403 A/B	1st Stage TOA Pump	Power = 8.6 kW Volumetric Flow Rate = 2,900 liters/minute Fluid= Organic Acid Mix Inlet Pressure = 220 kPa Outlet Pressure = 350 kPa Temperature = 95 °C MOC = Copper Alloy

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
L = Pumps		
L-410 A/B	3rd Stage TOA Pump	Power = 8.6 kW Volumetric Flow Rate = 2,900 liters/minute Fluid= Organic Acid Mix Inlet Pressure = 220 kPa Outlet Pressure = 350 kPa Temperature = 95 °C MOC = Copper Alloy
L-421 A/B	TOA Hold Up Pump	Power = 18 kW Volumetric Flow Rate = 1,400 liters/minute Inlet Pressure = 89 kPa Fluid= TOA Acid Extractant Outlet Pressure = 450 kPa Temperature = 35 °C MOC = Cast Iron
L-433 A/B	Volatile Acid Column Reflux Pump	Power = 0.74 kW Volumetric Flow Rate = 7.8 liters/minute Fluid= Formic and Acetic Acid Inlet Pressure in = 338 kPa Outlet Pressure out = 796 kPa Temperature = 99 °C MOC = Copper Alloy
L-441 A/B	Pressure Swing Pump	Power = 0.74 kW Volumetric Flow Rate = 34 liters/minute Fluid= Acetic and Formic Acid Inlet Pressure = 40 kPa Outlet Pressure = 440 kPa Temperature = 35 °C MOC = Copper Alloy
L-453 A/B	Formic/Acetic Reflux Pump	Power = 0.74 kW Volumetric Flow Rate = 42 liters/minute Fluid= Formic Acid Inlet Pressure = 330 kPa Outlet Pressure = 470 kPa Temperature = 99 °C MOC = Copper Alloy

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 4		
M = Mixers		
M-402	1st Stage TOA Mixer	Duty = 54 kW Diameter = 1.4 m MOC = Stainless Steel
M-406	2nd Stage TOA Mixer	Duty = 54 kW Diameter = 1.4 m MOC = Stainless Steel
M-409	3rd Stage TOA Mixer	Duty = 54 kW Diameter = 1.4 m MOC = Stainless Steel
M-413	4th Stage TOA Mixer	Duty = 54 kW Diameter = 1.4 m MOC = Stainless Steel
M-415	5th Stage TOA Mixer	Duty = 54 kW Diameter = 1.4 m MOC = Stainless Steel
Process Area 5		
D = Pressure Vessels		
D-500	1st Stage TMA Mixing Drum	Height = 7.8 m Diameter = 3.9 m Temperature = 35 °C Pressure = 350 kPa MOC = Carbon Steel with Stainless Clad
D-501	1st Stage TMA Settler	Height = 2.1 m Diameter = 4.2 m Temperature = 35 °C Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-505	2nd Stage TMA Mixing Drum	Height = 7.8 m Diameter = 3.9 m Temperature = 35 °C Pressure = 280 kPa MOC = Carbon Steel with Stainless Clad
D-506	2nd Stage TMA Settler	Height = 2.1 m Diameter = 4.2 m Temperature = 35 °C Pressure = 250 kPa MOC = Carbon Steel with Stainless Steel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	**Equipment Name**	**Equipment Specifications**
Process Area 5		
D = Pressure Vessels		
D-510	3rd Stage TMA Mixing Drum	Height = 7.8 m Diameter = 3.9 m Temperature = 35 °C Pressure = 210 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-511	3rd Stage TMA Settler	Height = 2.1 m Diameter = 4.2 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-515	4th Stage TMA Mixing Drum	Height = 7.8 m Diameter = 3.9 m Temperature = 35 °C Pressure = 210 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-516	4th Stage TMA Settler	Height = 2.1 m Diameter = 4.2 m Temperature = 35 °C Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-520	TMA Acid Extractant Hold Up Drum	Height = 2.4 m Diameter = 4.8 m Pressure = 120 kPa Temperature = 35 °C MOC = Carbon Steel with Stainless Steel Cladding
D-530	TOA Acid Extactant Hold up Drum	Height = 4.4 m Diameter = 2.2 m Pressure = 100 kPa Temperature = 35 °C MOC = Carbon Steel with Stainless Steel Cladding

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
D = Pressure Vessels		
D-540	Back Extractant Solvent Recovery Column	Diameter = 2.0 m Height = 7.5 m Number of Trays = 10 Temperature = 224 °C Pressure = 218 kPa MOC = Carbon Steel with Stainless Steel Cladding
D-542	Back Extractant Recovery Reflux Drum	Height = 2.7 m Diameter = 1.3 m Pressure = 170 kPa Temperature = 54 °C MOC = Carbon Steel with Stainless Steel Cladding
D-550	TMA Acid Extractant Hold Up Drum	Height = 4.2 m Diameter = 2.1 m Pressure = 120 kPa Temperature = 54 °C MOC = Carbon Steel with Stainless Steel Cladding
D-560	Lactic/Levulinic Acid Column	Diameter = 1.1 m Height = 26 m Number of Trays = 40 Temperature = 395 °C Pressure = 395 kPa MOC = Carbon Steel
D-562	Lactic/Levulinic Reflux Drum	Height = 1.6 m Diameter = 0.79 m Pressure = 340 kPa Temperature = 232 °C MOC = Carbon Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
E = Heat Exchangers		
E-541 A/B	Back Extractant Recovery Condenser	Duty = 26 GJ/hr Area = 480 m^2 Heat Transfer Coefficient = 850 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 290 °C Max Pressure = 310 kPa MOC = Carbon Steel **Tube Side** TMA and Water Inlet Temperature = 54 °C Outlet Temperature = 54 °C Max. Pressure = 190 kPa MOC = Carbon Steel
E-544 A-C/D	Back Extractant Recovery Reboiler	Duty = 41 GJ/hr Area = 300 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Medium-High Pressure Steam Inlet Temperature = 238 °C Outlet Temperature = 232 °C Max. Pressure = 3310 kPa MOC = Carbon Steel **Tube Side** Lactic and Levulinic Acid Inlet Temperature = 224 °C Outlet Temperature = 224 °C Max Pressure = 218 kPa MOC = Nickel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
E = Heat Exchangers		
E-552 A/B	TMA/Water Cooler	Duty = 0.91 GJ/hr Area = 160 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Cooling Water Inlet Temperature = 30 °C Outlet Temperature = 49 °C Max. Pressure = 310 kPa MOC = Carbon Steel **Tube Side** Trimethylamine and Water Inlet Temperature = 54 °C Outlet Temperature = 35 °C Max Pressure = 260 kPa MOC = Carbon Steel
E-561 A/B	Lactic/Levulinic Condenser	Duty = 13 GJ/hr Area = 45 m^2 Heat Transfer Coefficient = 350 W/(m^2-°C) **Shell Side** Boiler Feed Water Inlet Temperature = 30 °C Outlet Temperature = 175 °C Max Pressure = 930 kPa MOC = Carbon Steel **Tube Side** Lactic Acid Inlet Temperature = 232 °C Outlet Temperature = 232 °C Max. Pressure = 360 kPa MOC = Nickel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
E = Heat Exchangers		
E-564 A/B	Lactic/Levulinic Reboiler	Duty = 0.84 GJ/hr Area = 3.7 m^2 Heat Transfer Coeff. = 350 W/(m^2-°C) **Shell Side** High Pressure Steam Inlet Temperature = 400 °C Outlet Temperature = 283 °C Max. Pressure = 4500 kPa MOC = Carbon Steel **Tube Side** Levulinic Acid Inlet Temperature = 278 °C Outlet Temperature = 278 °C Max Pressure = 395 kPa MOC = Nickel
L = Pumps		
L-513 A/B	3rd Stage TMA Pump	Power = 5.8 kW Volumetric Flow Rate = 3,300 liters/min. Fluid= Mixed Acids Inlet Pressure = 145 kPa Outlet Pressure = 210 kPa Temperature = 35 °C MOC = Cast Iron
L-518 A/B	4th Stage TMA Pump	Power = 5.8 kW Volumetric Flow Rate = 3,300 liters/minute Fluid= Mixed Acids Inlet Pressure = 140 kPa Outlet Pressure = 2,000 kPa Temperature = 35 °C MOC = Cast Iron
L-521 A/B	TMA Hold Up Drum Pump	Power = 27 kW Volumetric Flow Rate = 9,000 liters/minute Fluid= TMA Back Extractant Inlet Pressure = 120 kPa Outlet Pressure = 230 kPa Temperature = 35 °C MOC = Stainless Steel

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
L = Pumps		
L-531 A/B	TOA Hold Up Drum Pump	Power = 11 kW Volumetric Flow Rate = 3,900 liters/minute Fluid= TOA/ 1-Octanol Solution Inlet Pressure = 100 kPa Outlet Pressure = 210 kPa Temperature = 35 °C MOC = Cast Iron
L-543 A/B	Back Extractant Recovery Reflux Pump	Power = 3.7 kW Volumetric Flow Rate = 2,600 liters/minute Fluid= TMA/Water Recycle Inlet Pressure = 160 kPa Outlet Pressure = 240 kPa Temperature = 54 °C MOC = Stainless Steel
L-551 A/B	TMA Hold Up Drum Pump	Duty = 24 kW Volumetric Flow Rate = 6,400 liters/minute Fluid= TMA/ Water Solution Inlet Pressure = 120 kPa Outlet Pressure = 260 kPa Temperature = 54 °C MOC = Stainless Steel
L-563 A/B	Lactic/Levulinic Reflux Pump	Duty = 0.74 kW Volumetric Flow Rate = 140 liters/minute Fluid= Lactic Acid Recycle Inlet Pressure = 330 kPa Outlet Pressure = 500 kPa Temperature = 232 °C MOC = Cast Iron
L-565 A/B	Lactic/Levulinic Feed Pump	Duty = 0.74 kW Volumetric Flow Rate = 140 liters/minute Fluid= Lactic and Levulinic Acid Inlet Pressure = 180 kPa Outlet Pressure = 430 kPa Temperature = 224 °C MOC = Cast Iron

Table 39. Major Equipment for the Sn-Beta + CaSO4 Process (cont.)

Equipment ID	Equipment Name	Equipment Specifications
Process Area 5		
M = Mixers		
M-502	1st Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel
M-507	2nd Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel
M-512	3rd Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel
M-517	4th Stage TMA Mixer	Duty = 32 kW Diameter = 1.2 m MOC = Stainless Steel

APPENDIX I – CAPITAL COST SUMMARY TABLES

Table 40. Capital Cost Summary for Base Sn-Beta Process

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 1 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
D = Pressure Vessels										
D-100	Sulfuric Acid Mixing Vessel	1	Height = 2.0 m Diameter = 1.0 m MOC = Carbon Steel with Nickel Cladding	$4,000	$42,000	4.5	1.1	9	$370,000	$370,000
D-105	NaOH Mixing Vessel	1	Height = 2.1 m Diameter = 4.2 m MOC = Carbon Steel	$15,000	$120,000	1	1	4.1	$510,000	$510,000
D-110	Deacetylation Vessel	4	Height = 8.5 m Diameter = 4.3 m Pressure= 110 kPa MOC= Carbon Steel with Stainless Cladding	$94,000	$140,000	2.5	1.1	6.5	$900,000	$3,600,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 2 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
D = Pressure Vessels										
D-121	Acid Presteamer	1	Height = 5.8 m Diameter = 2.9 m Temperature = 238 °C Pressure = 3300 kPa MOC = Carbon Steel with Stainless Steel Cladding	$40,000	$62,000	2.5	1.1	7.5	$460,000	$460,000
D-132	Acid Hydrolysis Rotary Vacuum Knockout Drum	1	Diameter = 4.0 m Height = 7.9 m Pressure = 55 kPa MOC = Carbon Steel with Nickel Cladding	$100,000	$150,000	4.5	1.1	9	$1,400,000	$1,400,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 3 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
D = Pressure Vessels										
D-161	Gypsum Rotary Vacuum Knockout Drum	1	Diameter = 3.1 m Height = 6.1 m Pressure = 55 kPa MOC = Carbon Steel with Stainless Steel Cladding	$40,000	$62,000	2.5	1.1	6.5	$400,000	$400,000
G = Compressors										
G-135	Acid Hydrolysis Rotary Vacuum Air Compressor	1	Duty = 13 kW MOC = Carbon Steel	$10,000	$15,000	2.5	1	2.5	$39,000	$39,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 4 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
G = Compressors										
G-135 - M	Acid Hydrolysis Rotary Vacuum Air Compressor Motor	1	Duty = 13 kW MOC = Carbon Steel	$15,000	$23,000	1.5	1	1.5	$35,000	$35,000
G-164	Gypsum Rotary Vacuum Air Compressor	1	Duty = 4.4 kW MOC = Carbon Steel	$8,000	$12,000	2.5	1	2.5	$31,000	$31,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 5 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
G = Compressors										
G-164 - M	Gypsum Rotary Vacuum Air Compressor Motor	1	Duty = 4.4 kW MOC = Carbon Steel	$500	$770	1.5	1	1.5	$1,200	$1,200
H = Filters										
H-131	Acid Rotary Vacuum Filter	7	Duty = 1.5 kW Volume = 15 m^3 Disks = 6 Disk Size = 6 ft MOC = Nickel Alloy	-	$250,000	5	1	5	$1,300,000	$9,100,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 6 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
H = Filters										
H-161	Gypsum Rotary Vacuum Filter	1	Duty = 0.74 kW Volume = 5.1 m^3 Disks = 2 Disk Size = 4 ft MOC = Carbon Steel	-	$100,000	2.4	1	2.4	$240,000	$240,000
J = Conveyors										
J-108	NaOH Feed Belt Conveyor	1	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$68,000
J-108 - Motor	NaOH Feed Belt Conveyor Motor	2	Power = 0.76 kW	$200	$310	2	1	2	$620	$1,200

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 7 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-111	Deacetylation Screw Conveyor	4	Length = 15 m Diameter = 0.45 m MOC = Stainless Steel	$10,000	$15,000	2.4	1	2.4	$37,000	$150,000
J-111 - Motor	Deacetylation Screw Conveyor Motor	8	Power = 3.7 kW	$400	$620	2	1	2	$1,200	$9,900
J-112	Pre-Deacetylation Stover Belt Conveyor	4	Length= 15 m Width= 0.45 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$270,000
J-112-M	Pre-Deacetylation Stover Belt Conveyor Motor	8	Power = 2.2 kW	$300	$460	2	1	2	$920	$7,400

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 8 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-114	Post-Deacetylation Corn Stover Belt	4	Length= 15 m Width= 0.45 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$270,000
J-114-M	Post-Deacetylation Corn Stover Belt Motor	8	Power = 2.2 kW	$300	$460	2	1	2	$920	$7,400
J-122	Pre-Hydrolysis Stover Belt Conveyor	1	Length= 15 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$68,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 9 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-122-M	Pre-Hydrolysis Stover Belt Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500
J-134	Acid Filtration Corn Stover Belt Conveyor	7	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$470,000
J-134-M	Acid Filtration Corn Stover Belt Conveyor Motor	14	Power = 1.5 kW	$250	$390	2	1	2	$770	$11,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 10 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-152	Calcium Carbonate Feed Conveyor	1	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$100,000	$150,000	2.2	1	2.2	$340,000	$340,000
J-152-M	Calcium Carbonate Feed Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500
J-153	Calcium Carbonate Screw Conveyor	1	Length= 15 m Width= 0.61 m MOC = Stainless Steel	$3,500	$5,400	2	1	2	$11,000	$11,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 11 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-153-M	Calcium Carbonate Screw Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500
J-162	Gypsum Belt Conveyor	1	Length= 15 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$68,000
J-162-M	Gypsum Belt Conveyor Motor	2	Power = 1.5 kW	$200	$310	2	1	2	$620	$1,200

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 12 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
L = Pumps										
L-107	NaOH Feed Pump	2	Power = 12 kW Inlet Pressure = 83 kPa MOC = Stainless Steel	$10,000	$15,000	1	1	3.5	$54,000	$110,000
L-113	Pre-Deacetylation Corn Stover Pump	4	Power = 0.74 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$2,000	$3,100	1	1	3.5	$11,000	$43,000
L-116	Black Liquor Pump	2	Power = 3.7 kW Inlet Pressure = 69 kPa MOC = Cast Iron	$5,000	$7,700	1	1	3.5	$27,000	$54,000
L-133	Acid Hydrolysis Rotary Vac. Slurry Pump	2	Power = 19 kW Inlet Pressure = 34 kPa MOC = Cast Iron	$12,000	$18,000	1	1	7.2	$130,000	$270,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 13 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
L = Pumps										
L-151	Acid Neutralization Slurry Pump	3	Power = 11 kW Inlet Pressure = 240 kPa MOC = Cast Iron	$15,000	$23,000	1	1	3.5	$81,000	$240,000
L-165	Gypsum Rotary Vacuum Slurry Pump	2	Power = 46 kW Inlet Pressure = 35 kPa MOC = Cast Iron	$18,000	$28,000	1	1	3.5	$97,000	$190,000
M = Mixers										
M-101	Sulfuric Acid Mixer	1	Agitator Power = 2.1 kW MOC= Stainless Steel	$20,000	$31,000	2.5	1	2.5	$77,000	$77,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 14 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
M = Mixers										
M-106	NaOH Mixer	1	Agitator Power = 13 kW MOC= Carbon Steel	$55,000	$85,000	2	1	2	$170,000	$170,000
M-115	Deacetylation Mixer	4	Agitator Power = 96 kW MOC= Carbon Steel	$110,000	$170,000	2	1	2	$340,000	$1,400,000
R = Reactors										
R-120	Acid Hydrolysis Reactor	1	Pressure= 560 kPa Duty = 29 kW MOC = Stainless Steel	$45,000	$69,000	2.5	1	2.5	$170,000	$170,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 15 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
R = Reactors										
R-150	Acid Hydrolysis Neutralization Reactor	1	Height = 5.8 m Diameter = 11 m Pressure = 240 kPa MOC = Carbon Steel with Nickel Cladding	$45,000	$70,000	8.5	1	8.5	$590,000	$590,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
D = Pressure Vessels										
D-221	Enzymatic Filtration Knockout Drum	1	Diameter = 3.1 m Height = 6.7 m Pressure= 55 kPa MOC = Carbon Steel with Stainless Steel Cladding	$43,000	$67,000	2.5	1.1	6.5	$440,000	$440,000
D-230	Hydrolysate Mixing Drum	1	Diameter = 2.5 m Height = 4.9 m Pressure = 220 kPa MOC = Carbon Steel	$17,000	$27,000	1	1.1	1.1	$29,000	$29,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 17 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
E = Heat Exchangers										
E-212	Batch Hydrolysis Recycle Heat Exchanger	8	Transfer Area = 1.7 m^2 Shell/Tube MOC = cs/cs Max Pressure = 45 psia	$2,000	$3,100	1	1	3.1	$9,500	$76,000
F-205	Pre-Batch Hold Up	1	Diameter = 9.1 m Height = 18 m Pressure =100kPa MOC = Car.Steel	$24,000	$42,000	5	1	5	$180,000	$180,000
F-213	Post-Batch Hold Up	1	Diameter = 9.1 m Height = 18 m MOC = Carbon Steel	$24,000	$42,000	1.9	1	1.9	$70,000	$70,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 18 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
G = Compressors										
G-223	Enzymatic Filtration Air Compressor	1	Duty = 13 kW MOC = Carbon Steel	$10,000	$15,000	2.5	1	2.5	$39,000	$39,000
G-223 - M	Enzymatic Filtration Air Compressor Motor	1	Duty = 13 kW MOC = Carbon Steel	$15,000	$23,000	1.5	1	1.5	$35,000	$35,000
H = Filters										
H-220	Enzymatic Rotary Vacuum Filter	3	Duty = 1.5 kW Volume = 500 m^2 Disks = 6 Disk Size = 6 ft MOC = Carbon Steel with Stainless Steel Cladding	-	$250,000	2.4	1	2.4	$600,000	$1,800,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 19 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
J = Conveyors										
J-202	Pre-Enzymatic Screw Conveyor	8	Length = 15 m Diameter =0.45 m MOC = Stainless Steel	$10,000	$15,000	2.4	1	2.4	$37,000	$300,000
J-202-M	Pre-Enzymatic Screw Conveyor Motor	16	Power = 2.2 kW	$300	$460	2	1	2	$920	$15,000
J-204	Corn Stover Belt Conveyor	1	Length= 15 m Width= 0.61 m MOC = Elastomer (rubber) material	$100,000	$150,000	2.2	1	2.2	$340,000	$340,000
J-204-M	Corn Stover Belt Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 20 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
J = Conveyors										
J-224	Lignin Rich Belt Conveyor	3	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$200,000
J-224-M	Lignin Rich Belt Conveyor Motor	6	Power = 1.5 kW	$250	$390	2	1	2	$770	$4,600
L = Pumps										
L-201	Enzymatic Slurry Pump	2	Power = 33 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$15,000	$23,000	1	1	3.5	$81,000	$160,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 21 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
L = Pumps										
L-211	Batch Hydrolysis Recycle Pump	8	Power = 1.5 kW Inlet Pressure = 340 kPa MOC = Cast Iron	$4,500	$6,900	1	1	3.5	$24,000	$190,000
L-222	Enzymatic Filtration Pump	2	Power = 4.5 kW Inlet Pressure = 35 kPa MOC = Cast Iron	$7,200	$11,000	1	1	3.5	$39,000	$78,000
L-225	Enzymatic Filtration Fee d Slurry Pump	3	Power = 6.7 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$11,000	$17,000	1	1	3.5	$59,000	$180,000
M = Mixers										
M-214	Batch Hydrolysis Mixer	7	Agitator Power = 430 kW MOC= Car. Steel	$300,000	$460,000	2	1	2	$920,000	$6,500,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 22 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
R = Reactors										
R-200	Continuous, High-Solids Enzymatic Reactor	9	Height = 17 m Diameter = 4.3 m Pressure = 300 kPa MOC = Carbon Steel with Stainless Steel Cladding	$90,000	$140,000	5	1	5	$690,000	$6,200,000
R-210	Batch Enzymatic Reactor	7	Height = 16 m Diameter = 8.2 m Pressure = 340 kPa MOC = Carbon Steel	$110,000	$170,000	1	1.1	4.2	$710,000	$5,000,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-310	Furfural Flash Drum	1	Height = 3.7 m Diameter = 7.4 m Pressure = 172 kPa MOC = Carbon Steel with Stainless Steel Cladding	$19,000	$30,000	2.5	1.1	6.5	$190,000	$190,000
D-320	1st Stage Furfural Extraction Mixing Drum	1	Height = 7.4 m Diameter = 7.3 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding	$55,000	$85,000	2.5	1.1	6.5	$550,000	$550,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 24 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
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ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-321	1st Stage Furfural Extraction Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$11,000	$17,000	2.5	1.1	6.5	$110,000	$110,000
D-325	2nd Stage Furfural Extraction Mixing Drum	1	Height = 7.4 m Diameter = 7.3 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding	$55,000	$85,000	2.5	1.1	6.5	$550,000	$550,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 25 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-326	2nd Stage Furfural Extraction Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$11,000	$17,000	2.5	1.1	6.5	$110,000	$110,000
D-330	3rd Stage Furfural Extraction Mixing Drum	1	Height = 7.4 m Diameter = 7.3 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding	$55,000	$85,000	2.5	1.1	6.5	$550,000	$550,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-331	3rd Stage Furfural Extraction Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$11,000	$17,000	2.5	1.1	6.5	$110,000	$110,000
D-340	Furfural Extraction Hold Up Drum	1	Height = 0.79 m Diameter = 0.39 m Pressure = 60 kPa MOC = Carbon Steel with Stainless Steel Cladding	$630	$970	2.5	1.1	5.5	$5,300	$5,300

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 27 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-350	Toluene Recovery Column	1	Diameter= 0.23 m Height = 12 m Number of Trays = 29 Pressure = 217 kPa MOC = Carbon Steel with Stainless Steel Cladding	$19,000	$29,000	2.5	1.1	6.5	$190,000	$190,000
D-350-T	Toluene Recovery Column Trays	29	Diameter= 0.23 m Number of Trays = 29 MOC = Stainless Steel	$55	$85	2.2	1.1	2.3	$200	$5,700

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-352	Toluene Recovery Column Reflux Drum	1	Height = 0.28 m Length= 0.85 m Pressure = 159 kPa MOC = Carbon Steel with Stainless Steel Cladding	$1,000	$1,500	2.5	1	2.5	$3,900	$3,900
D-360	Toluene Recycle Drum	1	Height = 0.94 m Diameter = 0.3 m Pressure = 117 kPa MOC = Carbon Steel with Stainless Steel Cladding	$1,000	$1,500	2.5	1.1	5.5	$8,500	$8,500

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 29 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
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ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-372	1st Compressor Inter stage Knockout Drum	1	Height = 3.6 m Diameter = 7.3 m Pressure = 270 kPa MOC = Carbon Steel	$68,000	$110,000	1	1.1	4.2	$440,000	$440,000
D-374	2nd Compressor Inter stage Knockout Drum	1	Height = 2.8 m Diameter = 5.5 m Pressure = 760 kPa MOC = Carbon Steel	$32,000	$49,000	1	1.2	4.4	$220,000	$220,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 30 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
E = Heat Exchangers										
E-302	Catalytic Decomposition Preheater	2	Area = 710 m^2 Shell/Tube MOC = cs/cs Max Pressure = 3300 kPa	$61,000	$94,000	1	1.1	3.2	$300,000	$600,000
E-311	Pre-Furfural Extraction Cooler	5	Area = 710 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 1800 kPa	$250,000	$380,000	1	1.1	3.2	$1,200,000	$6,000,000
E-351	Toluene Recovery Condenser	2	Area = 3.9 m^2 Shell/Tube MOC = cs/cs Max Pressure = 310 kPa	$3,800	$5,900	1	1	3.1	$18,000	$36,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 31 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
E = Heat Exchangers										
E-354	Toluene Recovery Reboiler	2	Area = 2.4 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 1700 kPa	\$3,300	\$5,100	1	1	3.1	\$16,000	\$32,000
E-362	Toluene Cooler	2	Transfer Area = 9.3 m^2 Shell/Tube MOC = cs/cs Max Pressure = 310 kPa	\$4,300	\$6,600	1	1	3.1	\$21,000	\$41,000
E-371	First-interstage cooler	2	Area = 470 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 280 kPa	\$32,000	\$49,000	1	1	3.1	\$150,000	\$310,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 32 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
E = Heat Exchangers										
E-373	Second-interstage cooler	2	Area = 79 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 780 kPa	$9,700	$15,000	1	1	3.1	$47,000	$93,000
G = Compressors										
G-370	Regeneration Compressor	1	Duty = 6,900 kW MOC = Carbon Steel	-	$2,400,000	1.5	1	1.5	$3,600,000	$3,600,000
L = Pumps										
L-301	Pre-Catalytic Decomposition Pump	2	Power = 140 kW Inlet Pressure = 62 kPa MOC = Cast Iron	$26,000	$40,000	1	1	3.5	$140,000	$280,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 33 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
L = Pumps										
L-323	1st Stage Furfural Pump	2	Power = 4.8 kW Inlet Pressure = 100 kPa MOC = Copper Alloy	$5,800	$8,900	1.3	1	4.6	$41,000	$82,000
L-328	2nd Stage Furfural Pump	2	Power = 4.8 kW Inlet Pressure = 100 kPa MOC = Copper Alloy	$5,800	$8,900	1.3	1	4.6	$41,000	$82,000
L-333	3rd Stage Furfural Pump	2	Power = 8.7 kW Inlet Pressure = 100 kPa MOC = Copper Alloy	$7,700	$12,000	1.3	1	4.6	$54,000	$110,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
L = Pumps										
L-341	Furfural Extractant Pump	2	Power = 0.74 kW Inlet Pressure = 35 kPa MOC = Cast Iron	$380	$580	1	1	3.5	$2,000	$4,100
L-353	Toluene Recovery Reflux Pump	2	Power = 0.74 kW Inlet Pressure = 150 kPa MOC = Cast Iron	$380	$580	1	1	3.5	$2,000	$4,100
L-361	Toluene Pump	2	Power = 0.74 kW Inlet Pressure = 117 kPa MOC = Cast Iron	$280	$430	1	1	3.5	$1,500	$3,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 35 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
M = Mixers										
M-322	1st Stage Furfural Mixer	1	Agitator Power = 39 kW MOC= Stainless Steel	$64,000	$99,000	2.5	1	2.5	$250,000	$250,000
M-327	2nd Stage Furfural Mixer	1	Agitator Power = 39 kW MOC= Stainless Steel	$64,000	$99,000	2.5	1	2.5	$250,000	$250,000
M-332	3rd Stage TOA Mixer	1	Agitator Power = 39 kW MOC= Stainless Steel	$64,000	$99,000	2.5	1	2.5	$250,000	$250,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 36 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
R = Reactors										
R-300	Sn+2 Beta Zeolite Packed Bed Reactor	18	Diameter = 1.8 m Height = 15 m Temperature = 200 °C Pressure = 2,070 kPa MOC = Carbon Steel with Stainless Steel Cladding	$25,000	$39,000	5	2	10	$390,000	$7,100,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-400	1st Stage TOA Mixing Drum	1	Height = 8.3 m Diameter = 4.1 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$70,000	$110,000	2.5	1.1	6.5	$700,000	$1,400,000
D-401	1st Stage TOA Settler	1	Height = 4.6 m Diameter = 2.3 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Cladding	$16,000	$25,000	2.5	1.1	5.5	$140,000	$280,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 46 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-405	2nd Stage TOA Mixing Drum	1	Height = 8.3 m Diameter = 4.1 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$70,000	$110,000	2.5	1.1	6.5	$700,000	$1,400,000
D-406	2nd Stage TOA Settler	1	Height = 4.6 m Diameter = 2.3 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Cladding	$16,000	$25,000	2.5	1.1	5.5	$140,000	$280,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 39 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-420	TOA Acid Extractant Hold Up Drum	1	Height = 4.4 m Diameter = 2.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$13,000	$23,000	2.5	1.1	5.5	$130,000	$130,000
D-430-R	Volatile Acid Recovery Column Rectifying Section	1	Rectifying Diameter = 0.58m Height = 7 m Pressure = 410 kPa MOC = Carbon Steel with Stainless Clad	$15,000	$23,000	2.5	1.1	6.5	$150,000	$150,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 40 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-430-S	Volatile Acid Recovery Column Stripping Section	1	Stripping Diameter = 3.9 m Height = 6 m Pressure = 410 kPa MOC = Carbon Steel with Stainless Clad	$62,000	$95,000	2.5	1.1	6.5	$620,000	$620,000
D-430-TR	Volatile Acid Recovery Column Trays for Rectifying Section	20	Rectifying Diameter = 0.58m Number of Trays = 20 MOC = Stainless Steel	$250	$390	2.2	1.1	2.3	$900	$10,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 41 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-430-TS	Volatile Acid Recovery Column Trays for Stripping Section	15	Stripping Diameter = 3.9 m Number of Trays = 15 MOC = Stainless Steel	$7,000	$11,000	2.2	1.2	2.6	$28,000	$280,000
D-432	Volatile Acid Column Reflux Drum	1	Height = 1.5 m Diameter = 0.73 m Pressure = 345 kPa MOC = Carbon Steel with Stainless Steel Cladding	$2,400	$3,700	2.5	1.1	5.5	$20,000	$20,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 42 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-440	Acid/l-octanol Pressure Swing Adsorber	4	Total Height = 8.9 m Diameter = 3.5 m Maximum Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$53,000	$81,000	2.5	1.1	6.5	$530,000	$2,100,000
D-450	Formic/ Acetic Distillation Column	1	Diameter = 1.3 m Height = 43 m # of Trays = 70 Pressure = 400 kPa MOC = Carbon Steel with Stainless Clad	$77,000	$120,000	2.5	1.1	6.5	$770,000	$770,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 43 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-450-T	Formic/ Acetic Distillation Column Trays	70	Diameter = 1.3 m Number of Trays = 70 Pressure = 400 kPa MOC = Stainless Steel	$750	$1,200	2.2	1	2.2	$2,500	$180,000
D-452	Formic/ Acetic Acid Column Reflux Drum	1	Height = 2.2 m Diameter = 1.1 m Pressure = 340 kPa MOC = Carbon Steel with Stainless Steel Cladding	$4,500	$7,000	2.5	1.1	5.5	$38,000	$38,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 44 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
E = Heat Exchangers										
E-431	Volatile Acid Column Condenser	2	Area = 19 m^2 Shell/Tube MOC = cs/Cu Max. Pressure = 365 kPa	$4,600	$7,000	1.3	1	3.5	$25,000	$49,000
E-434	Volatile Acid Column Reboiler	3	Area = 590 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 3300 kPa	$46,000	$70,000	2.8	1.1	5.8	$400,000	$1,210,000
E-435	Heavy Acid Extractant Cooler	4	Area = 740 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 450 kPa	$43,000	$66,000	2.8	1	5.6	$370,000	$1,480,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 45 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
E = Heat Exchangers										
E-442	Volatile Acid Cooler	2	Area = 81 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 45 psia	$9,300	$14,000	1	1	3.1	$44,000	$89,000
E-443	Recovered l-ocatanol Cooler	2	Area = 35 m^2 Shell/Tube MOC = cs/Cu Max. Pressure = 310 kPa	$7,900	$12,000	1	1	3.1	$38,000	$76,000
E-451	Formic/ Acetic Column Condenser	2	Area = 91 m^2 Shell/Tube MOC = cs/Cu Max. Pressure = 360 kPa	$13,000	$21,000	1	1	3.1	$64,000	$130,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 46 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
E = Heat Exchangers										
E-454	Formic/Acetic Column Reboiler	2	Area = 27 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 450 kPa	\$6,600	\$10,200	2.8	1	5.6	\$57,000	\$110,000
L = Pumps										
L-403	1st Stage TOA Pump	2	Power = 8.3 kW Inlet Pressure = 220 kPa MOC = Copper Alloy	\$7,000	\$11,000	1.3	1	4.6	\$49,000	\$99,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
L = Pumps										
L-421	TOA Hold Up Pump	2	Power = 13 kW Inlet Pressure = 89 kPa MOC = Cast Iron	$9,200	$14,000	1	1	3.5	$50,000	$100,000
L-433	Volatile Acid Column Reflux Pump	2	Power = 0.74 kW Inlet Pressure in = 338 kPa MOC = Copper Alloy	$450	$700	1.3	1	4.6	$3,200	$10,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 48 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
L = Pumps										
L-441	Pressure Swing Pump	2	Power = 0.74 kW Inlet Pressure = 40 kPa MOC = Copper Alloy	$1,800	$2,700	1.3	1	4.6	$13,000	$25,000
L-453	Formic/Acetic Reflux Pump	2	Power = 0.74 kW Inlet Pressure = 330 kPa MOC = Copper Alloy	$1,000	$1,500	1.3	1	4.6	$7,000	$14,000
M = Mixers										
M-402	1st Stage TOA Mixer	1	Duty = 54 kW MOC = Stainless Steel	$79,000	$120,000	2.5	1	2.5	$300,000	$300,000
M-407	2nd Stage TOA Mixer	1	Duty = 54 kW MOC = Stainless Steel	$79,000	$120,000	2.5	1	2.5	$300,000	$300,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 49 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
M = Mixers										
M-412	3rd Stage TOA Mixer	1	Duty = 54 kW MOC = Stainless Steel	$79,000	$120,000	2.5	1	2.5	$300,000	$300,000
Process Area 5										
D = Pressure Vessels										
D-500	1st Stage TMA Mixing Drum	1	Height = 6.9 m Diameter = 3.5 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$57,000	$88,000	2.5	1.1	6.5	$570,000	$2,300,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 50 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-501	1st Stage TMA Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	5.5	$120,000	$500,000
D-505	2nd Stage TMA Mixing Drum	1	Height = 6.9 m Diameter = 3.5 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$26,000	$40,000	2.5	1.1	6.5	$100,000	$100,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 51 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-506	2nd Stage TMA Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	5.5	$120,000	$500,000
D-510	3rd Stage TMA Mixing Drum	1	Height = 6.9 m Diameter = 3.5 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$26,000	$40,000	2.5	1.1	6.5	$100,000	$100,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 52 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-511	3rd Stage TMA Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	5.5	$120,000	$500,000
D-515	4th Stage TMA Mixing Drum	1	Height = 6.9 m Diameter = 3.5 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$26,000	$40,000	2.5	1.1	6.5	$100,000	$100,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 53 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-516	4th Stage TMA Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	5.5	$120,000	$500,000
D-520	TOA Acid Extractant Hold Up Drum	1	Height = 2.3 m Diameter = 4.7 m Pressure = 120 kPa MOC = Carbon Steel with Stainless Steel Cladding	$26,000	$40,000	2.5	1	2.5	$100,000	$100,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 54 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-530	TOA Acid Extactant Hold up Drum	1	Height = 9.6 m Diameter = 5.9 m Pressure = 100 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	5.5	$120,000	$120,000
D-540	Back Extractant Solvent Recovery Column	1	Diameter = 1.3 m Height = 7.5 m # of Trays = 10 Pressure = 218 kPa MOC = Carbon Steel with Stainless Cladding	$17,000	$25,000	2.5	1.1	6.5	$170,000	$170,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 55 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-540-T	Back Extractant Solvent Recovery Column Trays	10	Diameter = 1.3 m Number of Trays = 10 Pressure = 218 kPa MOC = Stainless Steel	$2,400	$3,700	2.2	1.2	2.6	$10,000	$100,000
D-542	Back Extractant Recovery Reflux Drum	1	Height = 2.1 m Diameter = 1.0 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding	$38,000	$59,000	2.5	1.1	5.5	$320,000	$320,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 56 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-550	TOA Acid Extractant Hold Up Drum	1	Height = 3.2 m Diameter = 1.6 m Pressure = 120 kPa MOC = Carbon Steel with Stainless Steel Cladding	$15,000	$23,000	2.5	1.1	5.5	$120,000	$120,000
D-560	Lactic/ Levulinic Column	1	Diameter = 0.83 m Height = 26 m # of Trays = 40 Pressure = 395 kPa MOC = Carbon Steel	$37,000	$57,000	1	1.1	6.5	$370,000	$370,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 57 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-560-T	Lactic/ Levulinic Column Trays	40	Diameter = 0.83m Number of Trays = 40 Pressure = 395 kPa MOC = Stainless Steel	$930	$1,430	2.2	1.1	2.3	$3,300	$130,000
D-562	Lactic/ Levulinic Reflux Drum	1	Height = 0.98 m Diameter = 0.49 m Pressure = 340 kPa Temperature = 232 °C MOC = Carbon Steel	$930	$1,430	1	1.1	5.5	$7,900	$10,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 58 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
E = Heat Exchangers										
E-541	Back Extractant Recovery Condenser	2	Area = 270 m^2 Shell/Tube MOC = cs/cs Max Pressure = 310 kPa	$29,000	$45,000	1	1	3.1	$140,000	$280,000
E-544	Back Extractant Recovery Reboiler	4	Area = 120 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 3310 kPa	$16,000	$25,000	2.8	1.1	6	$150,000	$300,000
E-552	Lactic/ Levulinic Condenser	2	Area = 160 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 310 kPa	$3,000	$4,600	2.8	1	5.6	$30,000	$50,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 59 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
E = Heat Exchangers										
E-564	Lactic/Levulinic Reboiler	2	Area = 16 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 4500 kPa	$4,100	$6,300	2.8	1.1	6	$38,000	$76,000
L = Pumps										
L-513	3rd Stage TMA Pump	2	Power = 4.2 kW Inlet Pressure = 145 kPa MOC = Cast Iron	$7,800	$12,000	1	1	3.5	$42,000	$84,000
L-518	4th Stage TMA Pump	2	Power = 4.2 kW Inlet Pressure = 140 kPa MOC = Cast Iron	$2,000	$3,200	1	1	3.5	$11,000	$22,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 60 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
L = Pumps										
L-521	TMA Hold Up Drum Pump	2	Power = 13 kW Inlet Pressure = 120 kPa MOC = Stainless Steel	$8,300	$13,000	1.9	1	4.9	$62,000	$120,000
L-531	TOA Hold Up Drum Pump	2	Power = 8.0 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$9,300	$14,000	1	1	3.5	$50,000	$100,000
L-543	Back Extractant Recovery Reflux Pump	2	Power = 1.7 kW Inlet Pressure = 160 kPa MOC = Stainless Steel	$3,000	$4,600	1.9	1	4.9	$23,000	$46,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 61 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
L = Pumps										
L-551	TMA Hold Up Drum Pump	2	Duty = 11 kW Inlet Pressure = 120 kPa MOC = Stainless Steel	$10,000	$15,000	1.9	1	4.9	$75,000	$150,000
L-563	Lactic/Levulinic Reflux Pump	2	Duty = 0.74 kW Inlet Pressure = 330 kPa MOC = Cast Iron	$2,600	$3,900	1	1	3.5	$14,000	$28,000
L-565	Lactic/Levulinic Feed Pump	2	Duty = 0.74 kW Inlet Pressure = 180 kPa MOC = Cast Iron	$1,200	$1,800	1	1	3.5	$6,000	$13,000
M = Mixers										
M-502	1st Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$43,000	$66,000	2.5	1	2.5	$160,000	$160,000

Table 40. Capital Cost Summary for Base Sn-Beta Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 62 of 62
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
M = Mixers										
M-507	2nd Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$43,000	$66,000	2.5	1	2.5	$160,000	$160,000
M-512	3rd Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$43,000	$66,000	2.5	1	2.5	$160,000	$160,000
M-517	4th Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$43,000	$66,000	2.5	1	2.5	$160,000	$160,000

Total Bare Module Cost	$81,000,000
Contingency and Fee (18%)	$15,000,000
Total Module Cost	$95,000,000
Auxiliary Factor (30%)	$29,000,000
Fixed Capital Investment	$120,000,000
Working Capital	$19,000,000
Chemicals and Catalyst	$2,100,000
TCI	$140,000,000

Table 40. Capital Cost Summary for Sn-Beta + CaSO4 Process

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 1 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
D = Pressure Vessels										
D-100	Sulfuric Acid Mixing Vessel	1	Height = 2.0 m Diameter = 1.0 m MOC = Carbon Steel with Nickel Cladding	\$4,000	\$42,000	4.5	1.1	9	\$370,000	\$370,000
D-105	NaOH Mixing Vessel	1	Height = 2.1 m Diameter = 4.2 m MOC = Carbon Steel	\$15,000	\$120,000	1	1	4.1	\$510,000	\$510,000
D-110	Deacetylation Vessel	4	Height = 8.5 m Diameter = 4.3 m Pressure= 110 kPa MOC= Carbon Steel with Stainless Steel Cladding	\$94,000	\$140,000	2.5	1.1	6.5	\$900,000	\$3,600,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 2 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
D = Pressure Vessels										
D-121	Acid Presteamer	1	Height = 5.8 m Diameter = 2.9 m Temperature = 238 °C Pressure = 3300 kPa MOC = Carbon Steel with Stainless Steel Cladding	$40,000	$62,000	2.5	1.1	7.5	$460,000	$460,000
D-132	Acid Hydrolysis Rotary Vacuum Knockout Drum	1	Diameter = 4.0 m Height = 7.9 m Pressure = 55 kPa MOC = Carbon Steel with Nickel Cladding	$100,000	$150,000	4.5	1.1	9	$1,400,000	$1,400,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 3 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
D = Pressure Vessels										
D-161	Gypsum Rotary Vacuum Knockout Drum	1	Diameter = 3.1 m Height = 6.1 m Pressure = 55 kPa MOC = Carbon Steel with Stainless Steel Cladding	$40,000	$62,000	2.5	1.1	6.5	$400,000	$400,000
G = Compressors										
G-135	Acid Hydrolysis Rotary Vacuum Air Compressor	1	Duty = 13 kW MOC = Carbon Steel	$10,000	$15,000	2.5	1	2.5	$39,000	$39,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 4 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
G = Compressors										
G-135 - M	Acid Hydrolysis Rotary Vacuum Air Compressor Motor	1	Duty = 13 kW MOC = Carbon Steel	$15,000	$23,000	1.5	1	1.5	$35,000	$35,000
G-164	Gypsum Rotary Vacuum Air Compressor	1	Duty = 4.4 kW MOC = Carbon Steel	$8,000	$12,000	2.5	1	2.5	$31,000	$31,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 5 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
G = Compressors										
G-164 - M	Gypsum Rotary Vacuum Air Compressor Motor	1	Duty = 4.4 kW MOC = Carbon Steel	$500	$770	1.5	1	1.5	$1,200	$1,200
H = Filters										
H-131	Acid Rotary Vacuum Filter	7	Duty = 1.5 kW Volume = 15 m^3 Disks = 6 Disk Size = 6 ft MOC = Nickel Alloy	-	$250,000	5	1	5	$1,300,000	$9,100,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 6 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
H = Filters										
H-161	Gypsum Rotary Vacuum Filter	1	Duty = 0.74 kW Volume = 5.1 m^3 Disks = 2 Disk Size = 4 ft MOC = Carbon Steel	-	$100,000	2.4	1	2.4	$240,000	$240,000
J = Conveyors										
J-108	NaOH Feed Belt Conveyor	1	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$68,000
J-108 - Motor	NaOH Feed Belt Conveyor Motor	2	Power = 0.76 kW	$200	$310	2	1	2	$620	$1,200

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 7 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-111	Deacetylation Screw Conveyor	4	Length = 15 m Diameter = 0.45 m MOC = Stainless Steel	$10,000	$15,000	2.4	1	2.4	$37,000	$150,000
J-111 - Motor	Deacetylation Screw Conv. Motor	8	Power = 3.7 kW	$400	$620	2	1	2	$1,200	$9,900
J-112	Pre-Deacetylation Stover Belt Conveyor	4	Length= 15 m Width= 0.45 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$270,000
J-112-M	Pre-Deacetylation Stover Belt Conveyor Motor	8	Power = 2.2 kW	$300	$460	2	1	2	$920	$7,400

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 8 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-114	Post-Deacetylation Corn Stover Belt	4	Length= 15 m Width= 0.45 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$270,000
J-114-M	Post-Deacetylation Corn Stover Belt Motor	8	Power = 2.2 kW	$300	$460	2	1	2	$920	$7,400
J-122	Pre-Hydrolysis Stover Belt Conveyor	1	Length= 15 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$68,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 9 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-122-M	Pre-Hydrolysis Stover Belt Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500
J-134	Acid Filtration Corn Stover Belt Conveyor	7	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$470,000
J-134-M	Acid Filtration Corn Stover Belt Conveyor Motor	14	Power = 1.5 kW	$250	$390	2	1	2	$770	$11,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 10 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-152	Calcium Carbonate Feed Conveyor	1	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$100,000	$150,000	2.2	1	2.2	$340,000	$340,000
J-152-M	Calcium Carbonate Feed Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500
J-153	Calcium Carbonate Screw Conveyor	1	Length= 15 m Width= 0.61 m MOC = Stainless Steel	$3,500	$5,400	2	1	2	$11,000	$11,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 11 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
J = Conveyors										
J-153-M	Calcium Carbonate Screw Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500
J-162	Gypsum Belt Conveyor	1	Length= 15 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$68,000
J-162-M	Gypsum Belt Conveyor Motor	2	Power = 1.5 kW	$200	$310	2	1	2	$620	$1,200

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
L = Pumps										
L-107	NaOH Feed Pump	2	Power = 12 kW Inlet Pressure = 83 kPa MOC = Stainless Steel	$10,000	$15,000	1	1	3.5	$54,000	$110,000
L-113	Pre-Deacetylation Corn Stover Pump	4	Power = 0.74 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$2,000	$3,100	1	1	3.5	$11,000	$43,000
L-116	Black Liquor Pump	2	Power = 3.7 kW Inlet Pressure = 69 kPa MOC = Cast Iron	$5,000	$7,700	1	1	3.5	$27,000	$54,000
L-133	Acid Hydrolysis Rotary Vac. Slurry Pump	2	Power = 19 kW Inlet Pressure = 34 kPa MOC = Cast Iron	$12,000	$18,000	1	1	7.2	$130,000	$270,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (contin

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 13 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
L = Pumps										
L-151	Acid Neutralization Slurry Pump	3	Power = 11 kW Inlet Pressure = 240 kPa MOC = Cast Iron	$15,000	$23,000	1	1	3.5	$81,000	$240,000
L-165	Gypsum Rotary Vacuum Slurry Pump	2	Power = 46 kW Inlet Pressure = 35 kPa MOC = Cast Iron	$18,000	$28,000	1	1	3.5	$97,000	$190,000
M = Mixers										
M-101	Sulfuric Acid Mixer	1	Agitator Power = 2.1 kW MOC= Stainless Steel	$20,000	$31,000	2.5	1	2.5	$77,000	$77,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 14 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
M = Mixers										
M-106	NaOH Mixer	1	Agitator Power = 13 kW MOC= Carbon Steel	$55,000	$85,000	2	1	2	$170,000	$170,000
M-115	Deacetylation Mixer	4	Agitator Power = 96 kW MOC= Carbon Steel	$110,000	$170,000	2	1	2	$340,000	$1,400,000
R = Reactors										
R-120	Acid Hydrolysis Reactor	1	Pressure= 560 kPa Duty = 29 kW MOC = Stainless Steel	$45,000	$69,000	2.5	1	2.5	$170,000	$170,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 15 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 1										
R = Reactors										
R-150	Acid Hydrolysis Neutralization Reactor	1	Height = 5.8 m Diameter = 11 m Pressure = 240 kPa MOC = Carbon Steel with Nickel Cladding	$45,000	$70,000	8.5	1	8.5	$590,000	$590,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 16 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
D = Pressure Vessels										
D-221	Enzymatic Filtration Knockout Drum	1	Diameter = 3.1 m Height = 6.7 m Pressure= 55 kPa MOC = Carbon Steel with Stainless Steel Cladding	\$43,000	\$67,000	2.5	1.1	6.5	\$440,000	\$440,000
D-230	Hydrolysate Mixing Drum	1	Diameter = 2.5 m Height = 4.9 m Pressure = 220 kPa MOC = Carbon Steel	\$17,000	\$27,000	1	1.1	1.1	\$29,000	\$29,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 17 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
E = Heat Exchangers										
E-212	Batch Hydrolysis Recycle Heat Exchanger	8	Transfer Area = 1.7 m^2 Shell/Tube MOC = cs/cs Max Pressure = 45 psia	$2,000	$3,100	1	1	3.1	$9,500	$76,000
F-205	Pre-Batch Hold Up	1	Diameter = 9.1 m Height = 18 m Pressure =100kPa MOC = Car.Steel	$24,000	$42,000	5	1	5	$180,000	$180,000
F-213	Post-Batch Hold Up	1	Diameter = 9.1 m Height = 18 m MOC = Carbon Steel	$24,000	$42,000	1.9	1	1.9	$70,000	$70,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 18 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
G = Compressors										
G-223	Enzymatic Filtration Air Compressor	1	Duty = 13 kW MOC = Carbon Steel	$10,000	$15,000	2.5	1	2.5	$39,000	$39,000
G-223 - M	Enzymatic Filtration Air Compressor Motor	1	Duty = 13 kW MOC = Carbon Steel	$15,000	$23,000	1.5	1	1.5	$35,000	$35,000
H = Filters										
H-220	Enzymatic Rotary Vacuum Filter	3	Duty = 1.5 kW Volume = 500 m^2 Disks = 6 Disk Size = 6 ft MOC = Carbon Steel with Stainless Steel Cladding	-	$250,000	2.4	1	2.4	$600,000	$1,800,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 19 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
J = Conveyors										
J-202	Pre-Enzymatic Screw Conveyor	8	Length = 15 m Diameter =0.45 m MOC = Stainless Steel	$10,000	$15,000	2.4	1	2.4	$37,000	$300,000
J-202-M	Pre-Enzymatic Screw Conveyor Motor	16	Power = 2.2 kW	$300	$460	2	1	2	$920	$15,000
J-204	Corn Stover Belt Conveyor	1	Length= 15 m Width= 0.61 m MOC = Elastomer (rubber) material	$100,000	$150,000	2.2	1	2.2	$340,000	$340,000
J-204-M	Corn Stover Belt Conveyor Motor	2	Power = 1.5 kW	$250	$390	2	1	2	$770	$1,500

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 20 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
J = Conveyors										
J-224	Lignin Rich Belt Conveyor	3	Length= 150 m Width= 0.61 m MOC = Elastomer (rubber) material	$20,000	$31,000	2.2	1	2.2	$68,000	$200,000
J-224-M	Lignin Rich Belt Conveyor Motor	6	Power = 1.5 kW	$250	$390	2	1	2	$770	$4,600
L = Pumps										
L-201	Enzymatic Slurry Pump	2	Power = 33 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$15,000	$23,000	1	1	3.5	$81,000	$160,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 21 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
L = Pumps										
L-211	Batch Hydrolysis Recycle Pump	8	Power = 1.5 kW Inlet Pressure = 340 kPa MOC = Cast Iron	$4,500	$6,900	1	1	3.5	$24,000	$190,000
L-222	Enzymatic Filtration Pump	2	Power = 4.5 kW Inlet Pressure = 35 kPa MOC = Cast Iron	$7,200	$11,000	1	1	3.5	$39,000	$78,000
L-225	Enzymatic Filtration Feed Slurry Pump	3	Power = 6.7 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$11,000	$17,000	1	1	3.5	$59,000	$180,000
M = Mixers										
M-214	Batch Hydrolysis Mixer	7	Agitator Power = 430 kW MOC= Car. Steel	$300,000	$460,000	2	1	2	$920,000	$6,500,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 22 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 2										
R = Reactors										
R-200	Continuous, High-Solids Enzymatic Reactor	9	Height = 17 m Diameter = 4.3 m Pressure = 300 kPa MOC = Carbon Steel with Stainless Steel Cladding	$90,000	$140,000	5	1	5	$690,000	$6,200,000
R-210	Batch Enzymatic Reactor	7	Height = 16 m Diameter = 8.2 m Pressure = 340 kPa MOC = Carbon Steel	$110,000	$170,000	1	1.1	4.2	$710,000	$5,000,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 23 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-310	Furfural Flash Drum	1	Height = 3.7 m Diameter = 7.4 m Pressure = 172 kPa MOC = Carbon Steel with Stainless Steel Cladding	$19,000	$30,000	2.5	1.1	6.5	$190,000	$190,000
D-320	1st Stage Furfural Extraction Mixing Drum	1	Height = 7.4 m Diameter = 7.3 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Clad	$55,000	$85,000	2.5	1.1	6.5	$550,000	$550,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 24 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-321	1st Stage Furfural Extraction Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$11,000	$17,000	2.5	1.1	6.5	$110,000	$110,000
D-325	2nd Stage Furfural Extraction Mixing Drum	1	Height = 7.4 m Diameter = 7.3 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding	$55,000	$85,000	2.5	1.1	6.5	$550,000	$550,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 25 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-326	2nd Stage Furfural Extraction Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$11,000	$17,000	2.5	1.1	6.5	$110,000	$110,000
D-330	3rd Stage Furfural Extraction Mixing Drum	1	Height = 7.4 m Diameter = 7.3 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding	$55,000	$85,000	2.5	1.1	6.5	$550,000	$550,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 26 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-331	3rd Stage Furfural Extraction Settler	1	Height = 1.6 m Diameter = 3.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$11,000	$17,000	2.5	1.1	6.5	$110,000	$110,000
D-340	Furfural Extraction Hold Up Drum	1	Height = 0.79 m Diameter = 0.39 m Pressure = 60 kPa MOC = Carbon Steel with Stainless Steel Cladding	$630	$970	2.5	1.1	5.5	$5,300	$5,300

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 27 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-350	Toluene Recovery Column	1	Diameter= 0.23 m Height = 12 m Number of Trays = 29 Pressure = 217 kPa MOC = Carbon Steel with Stainless Steel Cladding	$19,000	$29,000	2.5	1.1	6.5	$190,000	$190,000
D-350-T	Toluene Recovery Column Trays	29	Diameter= 0.23 m Number of Trays = 29 MOC = Stainless Steel	$55	$85	2.2	1.1	2.3	$200	$5,700

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 28 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-352	Toluene Recovery Column Reflux Drum	1	Height = 0.28 m Length= 0.85 m Pressure = 159 kPa MOC = Carbon Steel with Stainless Steel Cladding	$1,000	$1,500	2.5	1	2.5	$3,900	$3,900
D-360	Toluene Recycle Drum	1	Height = 0.94 m Diameter = 0.3 m Pressure = 117 kPa MOC = Carbon Steel with Stainless Steel Cladding	$1,000	$1,500	2.5	1.1	5.5	$8,500	$8,500

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 29 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
D = Pressure Vessels										
D-372	1st Compressor Inter stage Knockout Drum	1	Height = 2.7 m Diameter = 5.3 m Pressure = 270 kPa MOC = Carbon Steel	$38,000	$60,000	1	1.1	4.2	$250,000	$250,000
D-374	2nd Compressor Inter stage Knockout Drum	1	Height = 2.0 m Diameter = 4.0 m Pressure = 760 kPa MOC = Carbon Steel	$18,000	$28,000	1	1.2	4.4	$120,000	$120,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 30 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
E = Heat Exchangers										
E-302	Catalytic Decomposition Preheater	2	Area = 710 m^2 Shell/Tube MOC = cs/cs Max Pressure = 3300 kPa	$61,000	$94,000	1	1.1	3.2	$300,000	$600,000
E-311	Pre-Furfural Extraction Cooler	5	Area = 710 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 1800 kPa	$250,000	$380,000	1	1.1	3.2	$1,200,000	$6,000,000
E-351	Toluene Recovery Condenser	2	Area = 3.9 m^2 Shell/Tube MOC = cs/cs Max Pressure = 310 kPa	$3,800	$5,900	1	1	3.1	$18,000	$36,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 31 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
E = Heat Exchangers										
E-354	Toluene Recovery Reboiler	2	Area = 2.4 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 1700 kPa	$3,300	$5,100	1	1	3.1	$16,000	$32,000
E-362	Toluene Cooler	2	Transfer Area = 9.3 m^2 Shell/Tube MOC = cs/cs Max Pressure = 310 kPa	$4,300	$6,600	1	1	3.1	$21,000	$41,000
E-371	First-interstage cooler	2	Area = 370 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 280 kPa	$27,000	$41,000	1	1	3.1	$130,000	$260,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 32 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
E = Heat Exchangers										
E-373	Second-interstage cooler	2	Area = 31 m^2 Shell/Tube MOC = cs/cs Max. Pressure = 780 kPa	$5,500	$8,500	1	1	3.1	$26,000	$53,000
G = Compressors										
G-370	Regeneration Compressor	1	Duty = 2,600 kW MOC = Carbon Steel	-	$1,400,000	1.5	1	1.5	$2,000,000	$2,000,000
L = Pumps										
L-301	Pre-Catalytic Decomposition Pump	2	Power = 140 kW Inlet Pressure = 62 kPa MOC = Cast Iron	$26,000	$40,000	1	1	3.5	$140,000	$280,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 33 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
L = Pumps										
L-323	1st Stage Furfural Pump	2	Power = 5.0 kW Inlet Pressure = 100 kPa MOC = Copper Alloy	$5,900	$9,100	1.3	1	4.6	$42,000	$84,000
L-328	2nd Stage Furfural Pump	2	Power = 5.0 kW Inlet Pressure = 100 kPa MOC = Copper Alloy	$5,900	$9,100	1.3	1	4.6	$42,000	$84,000
L-333	3rd Stage Furfural Pump	2	Power = 9.0 kW Inlet Pressure = 100 kPa MOC = Copper Alloy	$7,900	$12,000	1.3	1	4.6	$56,000	$110,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 34 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
L = Pumps										
L-341	Furfural Extractant Pump	2	Power = 0.74 kW Inlet Pressure = 35 kPa MOC = Cast Iron	$380	$580	1	1	3.5	$2,000	$4,100
L-353	Toluene Recovery Reflux Pump	2	Power = 0.74 kW Inlet Pressure = 150 kPa MOC = Cast Iron	$380	$580	1	1	3.5	$2,000	$4,100
L-361	Toluene Pump	2	Power = 0.74 kW Inlet Pressure = 117 kPa MOC = Cast Iron	$280	$430	1	1	3.5	$1,500	$3,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 35 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
M = Mixers										
M-322	1st Stage Furfural Mixer	1	Agitator Power = 39 kW MOC= Stainless Steel	$65,000	$100,000	2.5	1	2.5	$250,000	$250,000
M-327	2nd Stage Furfural Mixer	1	Agitator Power = 39 kW MOC= Stainless Steel	$65,000	$100,000	2.5	1	2.5	$250,000	$250,000
M-332	3rd Stage TOA Mixer	1	Agitator Power = 39 kW MOC= Stainless Steel	$65,000	$100,000	2.5	1	2.5	$250,000	$250,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 3										
R = Reactors										
R-300	Sn+2 Beta Zeolite Packed Bed Reactor	18	Diameter = 1.8 m Height = 15 m Temperature = 200 °C Pressure = 2,070 kPa MOC = Carbon Steel with Stainless Steel Cladding	$25,000	$39,000	5	2	10	$390,000	$7,100,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 37 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-400	1st Stage TOA Mixing Drum	1	Height = 8.0 m Diameter = 3.9 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$66,000	$100,000	2.5	1.1	6.5	$660,000	$3,300,000
D-401	1st Stage TOA Settler	1	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Clad	$18,000	$27,000	2.5	1.1	5.5	$150,000	$760,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 38 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-405	2nd Stage TOA Mixing Drum	1	Height = 8.0 m Diameter = 3.9 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$66,000	$100,000	2.5	1.1	6.5	$660,000	$3,300,000
D-406	2nd Stage TOA Settler	1	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Clad	$18,000	$27,000	2.5	1.1	5.5	$150,000	$760,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 39 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-407	3rd Stage TOA Mixing Drum	1	Height = 8.0 m Diameter = 3.9 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$66,000	$100,000	2.5	1.1	6.5	$660,000	$3,300,000
D-408	3rd Stage TOA Settler	1	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Clad	$18,000	$27,000	2.5	1.1	5.5	$150,000	$760,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 40 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-411	4th Stage TOA Mixing Drum	1	Height = 8.0 m Diameter = 3.9 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$66,000	$100,000	2.5	1.1	6.5	$660,000	$3,300,000
D-412	4th Stage TOA Settler	1	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Clad	$18,000	$27,000	2.5	1.1	5.5	$150,000	$760,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 41 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-413	5th Stage TOA Mixing Drum	1	Height = 8.0 m Diameter = 3.9 m Pressure = 180 kPa MOC = Carbon Steel with Stainless Steel Cladding	$66,000	$100,000	2.5	1.1	6.5	$660,000	$3,300,000
D-414	5th Stage TOA Settler	1	Height = 2.4 m Diameter = 4.8 m Temperature = 35 °C Pressure = 140 kPa MOC = Carbon Steel with Stainless Clad	$18,000	$27,000	2.5	1.1	5.5	$150,000	$760,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 42 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-420	TOA Acid Extractant Hold Up Drum	1	Height = 4.4 m Diameter = 2.2 m Pressure = 110 kPa MOC = Carbon Steel with Stainless Steel Cladding	$7,700	$23,000	2.5	1.1	5.5	$130,000	$130,000
D-430-R	Volatile Acid Recovery Column Rectifying Section	1	Rectifying Diameter = 0.61m Height = 7 m Pressure = 410 kPa MOC = Carbon Steel with Stainless Clad	$16,000	$25,000	2.5	1.1	6.5	$160,000	$160,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 43 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-430-S	Volatile Acid Recovery Column Stripping Section	1	Stripping Diameter = 4.8 m Height = 6 m Pressure = 410 kPa MOC = Carbon Steel with Stainless Clad	$80,000	$123,000	2.5	1.1	6.5	$800,000	$800,000
D-430-TR	Volatile Acid Recovery Column Trays for Rectifying Section	20	Rectifying Diameter = 0.61m Number of Trays = 20 MOC = Stainless Steel	$260	$400	2.2	1.1	2.3	$930	$10,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 44 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-430-TS	Volatile Acid Recovery Column Trays for Stripping Section	15	Stripping Diameter = 4.8 m Number of Trays = 15 MOC = Stainless Steel	$8,000	$12,000	2.2	1.2	2.6	$32,000	$320,000
D-432	Volatile Acid Column Reflux Drum	1	Height = 1.5 m Diameter = 0.76m Pressure = 345 kPa MOC = Carbon Steel with Stainless Steel Cladding	$2,600	$4,000	2.5	1.1	5.5	$22,000	$20,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 45 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-440	Acid/l-octanol Pressure Swing Adsorber	4	Total Height = 8.9 m Diameter = 3.5 m Max. Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$52,000	$80,000	2.5	1.1	6.5	$520,000	$2,100,000
D-450	Formic/ Acetic Distillation Column	1	Diameter = 1.6 m Height = 43 m # of Trays = 70 Pressure = 400 kPa MOC = Carbon Steel with Stainles Clad	$60,000	$92,000	2.5	1.1	6.5	$600,000	$600,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 46 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
D = Pressure Vessels										
D-450-T	Formic/ Acetic Distillation Column Trays	70	Diameter = 1.6 m Number of Trays = 70 Pressure = 400 kPa MOC = Stainless Steel	$700	$1,000	2.2	1	2.2	$2,200.0	$160,000
D-452	Formic/ Acetic Acid Column Reflux Drum	1	Height = 1.9 m Diameter = 0.95m Pressure = 340 kPa MOC = Carbon Steel with Stainless Steel Cladding	$3,500	$5,400	2.5	1.1	5.5	$29,000	$29,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 47 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
E = Heat Exchangers										
E-431	Volatile Acid Column Condenser	2	Area = 21 m^2 Shell/Tube MOC = cs/Cu Max. Pressure = 365 kPa	$4,900	$7,500	1.3	1	3.5	$26,000	$53,000
E-434	Volatile Acid Column Reboiler	3	Area = 900 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 3300 kPa	$59,000	$90,000	2.8	1.1	5.8	$520,000	$1,550,000
E-435	Heavy Acid Extractant Cooler	4	Area = 840 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 450 kPa	$46,000	$71,000	2.8	1	5.6	$400,000	$2,000,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 48 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
E = Heat Exchangers										
E-442	Volatile Acid Cooler	2	Area = 200 m^2 Shell/Tube MOC = cs/Cu Max Pressure = 45 psia	$16,100	$25,000	1	1	3.1	$77,000	$153,000
E-443	Recovered l-ocatanol Cooler	2	Area = 52 m^2 Shell/Tube MOC = cs/Cu Max. Pressure = 310 kPa	$10,000	$15,000	1	1	3.1	$48,000	$96,000
E-451	Formic/ Acetic Column Condenser	2	Area = 61 m^2 Shell/Tube MOC = cs/Cu Max. Pressure = 360 kPa	$10,000	$16,000	1	1	3.1	$49,000	$100,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 49 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
E = Heat Exchangers										
E-454	Formic/Acetic Column Reboiler	2	Area = 86 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 450 kPa	$13,000	$21,000	2.8	1	5.6	$110,000	$230,000
L = Pumps										
L-403	1st Stage TOA Pump	2	Power = 8.6 kW Inlet Pressure = 220 kPa MOC = Copper Alloy	$7,100	$11,000	1.3	1	4.6	$51,000	$100,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 50 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
L = Pumps										
L-410	3rd Stage TOA Pump	2	Power = 8.6 kW Inlet Pressure = 220 kPa MOC = Copper Alloy	$7,100	$11,000	1.3	1	4.6	$51,000	$100,000
L-421	TOA Hold Up Pump	2	Power = 18 kW Inlet Pressure = 89 kPa MOC = Cast Iron	$11,400	$17,000	1	1	3.5	$61,000	$120,000
L-433	Volatile Acid Column Reflux Pump	2	Power = 0.74 kW Inlet Pressure in = 338 kPa MOC = Copper Alloy	$450	$700	1.3	1	4.6	$3,200	$10,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
L = Pumps										
L-441	Pressure Swing Pump	2	Power = 0.74 kW Inlet Pressure = 40 kPa MOC = Copper Alloy	$1,800	$2,700	1.3	1	4.6	$13,000	$25,000
L-453	Formic/Acetic Reflux Pump	2	Power = 0.74 kW Inlet Pressure = 330 kPa MOC = Copper Alloy	$2,100	$3,200	1.3	1	4.6	$15,000	$30,000
M = Mixers										
M-402	1st Stage TOA Mixer	1	Duty = 54 kW MOC = Stainless Steel	$73,000	$110,000	2.5	1	2.5	$280,000	$280,000
M-407	2nd Stage TOA Mixer	1	Duty = 54 kW MOC = Stainless Steel	$73,000	$110,000	2.5	1	2.5	$280,000	$280,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 52 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 4										
M = Mixers										
M-412	3rd Stage TOA Mixer	1	Duty = 54 kW MOC = Stainless Steel	$73,000	$110,000	2.5	1	2.5	$280,000	$280,000
Process Area 5										
D = Pressure Vessels										
D-500	1st Stage TMA Mixing Drum	1	Height = 7.8 m Diameter = 3.9 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$71,000	$110,000	2.5	1.1	6.5	$710,000	$2,800,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 53 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-501	1st Stage TMA Settler	1	Height = 2.1 m Diameter = 4.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$24,000	$36,000	2.5	1.1	5.5	$200,000	$800,000
D-505	2nd Stage TMA Mixing Drum	1	Height = 7.8 m Diameter = 3.9 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$71,000	$110,000	2.5	1.1	6.5	$710,000	$2,800,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 54 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-506	2nd Stage TMA Settler	1	Height = 2.1 m Diameter = 4.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$24,000	$36,000	2.5	1.1	5.5	$200,000	$800,000
D-510	3rd Stage TMA Mixing Drum	1	Height = 7.8 m Diameter = 3.9 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$71,000	$110,000	2.5	1.1	6.5	$710,000	$2,800,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 55 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-511	3rd Stage TMA Settler	1	Height = 2.1 m Diameter = 4.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$24,000	$36,000	2.5	1.1	5.5	$200,000	$800,000
D-515	4th Stage TMA Mixing Drum	1	Height = 7.8 m Diameter = 3.9 m Pressure = 350 kPa MOC = Carbon Steel with Stainless Steel Cladding	$71,000	$110,000	2.5	1.1	6.5	$710,000	$2,800,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-516	4th Stage TMA Settler	1	Height = 2.1 m Diameter = 4.2 m Pressure = 320 kPa MOC = Carbon Steel with Stainless Steel Cladding	$24,000	$36,000	2.5	1.1	5.5	$200,000	$800,000
D-520	TOA Acid Extractant Hold Up Drum	1	Height = 2.4 m Diameter = 4.8 m Pressure = 120 kPa MOC = Carbon Steel with Stainless Steel Cladding	$29,000	$44,000	2.5	1	2.5	$110,000	$110,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 57 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-530	TOA Acid Extactant Hold up Drum	1	Height = 4.4 m Diameter = 2.2 m Pressure = 100 kPa MOC = Carbon Steel with Stainless Steel Cladding	$9,000	$14,000	2.5	1.1	5.5	$80,000	$80,000
D-540	Back Extractant Solvent Recovery Column	1	Diameter = 2.0 m Height = 7.5 m # of Trays = 10 Pressure = 218 kPa MOC = Carbon Steel with Stainless Clad	$29,000	$44,000	2.5	1.1	6.5	$290,000	$290,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 58 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-540-T	Back Extractant Solvent Recovery Column Trays	10	Diameter = 2.0 m Number of Trays = 10 Pressure = 218 kPa MOC = Stainless Steel	$3,200	$4,900	2.2	1.2	2.6	$13,000	$130,000
D-542	Back Extractant Recovery Reflux Drum	1	Height = 2.7 m Diameter = 1.3 m Pressure = 170 kPa MOC = Carbon Steel with Stainless Steel Cladding	$59,000	$92,000	2.5	1.1	5.5	$500,000	$500,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 59 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-550	TOA Acid Extractant Hold Up Drum	1	Height = 4.2 m Diameter = 2.1 m Pressure = 120 kPa MOC = Carbon Steel with Stainless Steel Cladding	$24,000	$36,000	2.5	1.1	5.5	$200,000	$200,000
D-560	Lactic/ Levulinic Column	1	Diameter = 1.1 m Height = 26 m # of Trays = 40 Pressure = 395 kPa MOC = Carbon Steel	$15,000	$23,000	1	1.1	6.5	$150,000	$150,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 60 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
D = Pressure Vessels										
D-560-T	Lactic/ Levulinic Column Trays	40	Diameter = 1.1 m Number of Trays = 40 Pressure = 395 kPa MOC = Stainless Steel	$600	$920	2.2	1.1	2.3	$2,100	$90,000
D-562	Lactic/ Levulinic Reflux Drum	1	Height = 1.6 m Diameter = 0.79m Pressure = 340 kPa Temperature = 232 °C MOC = Carbon Steel	$2,270	$3,500	1	1.1	5.5	$19,200	$20,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
E = Heat Exchangers										
E-541	Back Extractant Recovery Condenser	2	Area = 480 m^2 Shell/Tube MOC = cs/cs Max Pressure = 310 kPa	$46,000	$70,000	1	1	3.1	$220,000	$440,000
E-544	Back Extractant Recovery Reboiler	2	Area = 300 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 3310 kPa	$28,000	$43,000	2.8	1.1	6	$260,000	$510,000
E-552	Lactic/ Levulinic Condenser	2	Area = 160 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 310 kPa	$7,300	$11,000	2.8	1	5.6	$60,000	$130,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 62 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
E = Heat Exchangers										
E-564	Lactic/Levulinic Reboiler	2	Area = 3.7 m^2 Shell/Tube MOC = cs/Ni Max. Pressure = 4500 kPa	$1,700	$2,600	2.8	1.1	6	$16,000	$31,000
L = Pumps										
L-513	3rd Stage TMA Pump	2	Power = 5.8 kW Inlet Pressure = 145 kPa MOC = Cast Iron	$9,500	$15,000	1	1	3.5	$51,000	$102,000
L-518	4th Stage TMA Pump	2	Power = 5.8 kW Inlet Pressure = 140 kPa MOC = Cast Iron	$2,500	$3,800	1	1	3.5	$13,000	$27,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 63 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
L = Pumps										
L-521	TMA Hold Up Drum Pump	2	Power = 27 kW Inlet Pressure = 120 kPa MOC = Stainless Steel	$13,100	$20,000	1.9	1	4.9	$99,000	$200,000
L-531	TOA Hold Up Drum Pump	2	Power = 11 kW Inlet Pressure = 100 kPa MOC = Cast Iron	$11,300	$17,000	1	1	3.5	$61,000	$122,000
L-543	Back Extractant Recovery Reflux Pump	2	Power = 3.7 kW Inlet Pressure = 160 kPa MOC = Stainless Steel	$4,700	$7,200	1.9	1	4.9	$35,000	$71,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 64 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
L = Pumps										
L-551	TMA Hold Up Drum Pump	2	Duty = 24 kW Inlet Pressure = 120 kPa MOC = Stainless Steel	$15,500	$24,000	1.9	1	4.9	$120,000	$230,000
L-563	Lactic/Levulinic Reflux Pump	2	Duty = 0.74 kW Inlet Pressure = 330 kPa MOC = Cast Iron	$4,400	$6,800	1	1	3.5	$24,000	$48,000
L-565	Lactic/Levulinic Feed Pump	2	Duty = 0.74 kW Inlet Pressure = 180 kPa MOC = Cast Iron	$2,900	$4,500	1	1	3.5	$16,000	$32,000
M = Mixers										
M-502	1st Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$53,000	$81,000	2.5	1	2.5	$200,000	$200,000

Table 41. Capital Cost Summary for Sn-Beta + CaSO4 Process (continued)

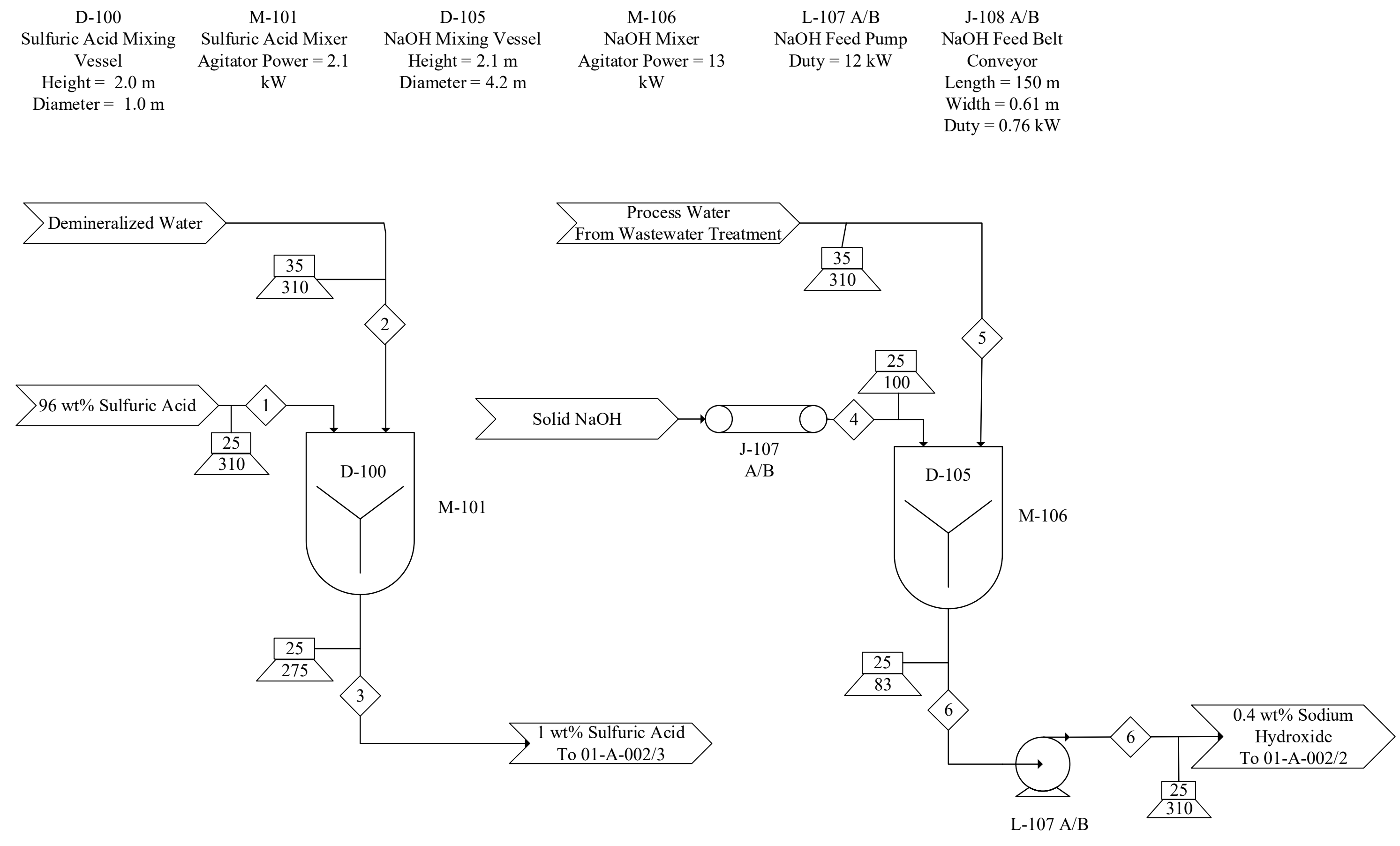
JOB TITLE: Scoping Study for Processing Corn Stover into Lactic and Levulinic Acid REVISION# 0 PAGE 65 of 65
LOCATION: North Dakota **DEVELOPED BY:** Andrew Kohler
DATE BASIS FOR ESTIMATE: December 2018

ID#	Equipment Description	# of Units	Capacity/Size Specification	PEC (2004)	PEC (2018)	Material Factor	Pressure/ Other Factor	Actual BMF	BMC	Total BMC
Process Area 5										
M = Mixers										
M-507	2nd Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$53,000	$81,000	2.5	1	2.5	$200,000	$200,000
M-512	3rd Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$53,000	$81,000	2.5	1	2.5	$200,000	$200,000
M-517	4th Stage TMA Mixer	1	Duty = 32 kW MOC = Stainless Steel	$53,000	$81,000	2.5	1	2.5	$200,000	$200,000

Total Bare Module Cost	$86,000,000
Contingency and Fee (18%)	$15,000,000
Total Module Cost	$100,000,000
Auxiliary Factor (30%)	$30,000,000
Fixed Capital Investment	$130,000,000
Working Capital	$20,000,000
Chemicals and Catalyst	$2,400,000
TCI	$150,000,000

APPENDIX J – PROCESS DRAWINGS

Figure 44. Base Sn-Beta Process Drawings



Components/Stream	1	2	3	4	5	6
Sodium Hydroxide	0	0	0	450	0	450
Sulfuric Acid	180	0	180	0	0	0
Water	7.7	18,000	18,000	0	110,000	110,000
Total	**190**	**18,000**	**18,000**	**450**	**110,000**	**110,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

◇ Stream Identification

Input/Output

■ Temperature (°C), gas

☐ Temperature (°C), liquid

▲ Pressure (kPa), gas

△ Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

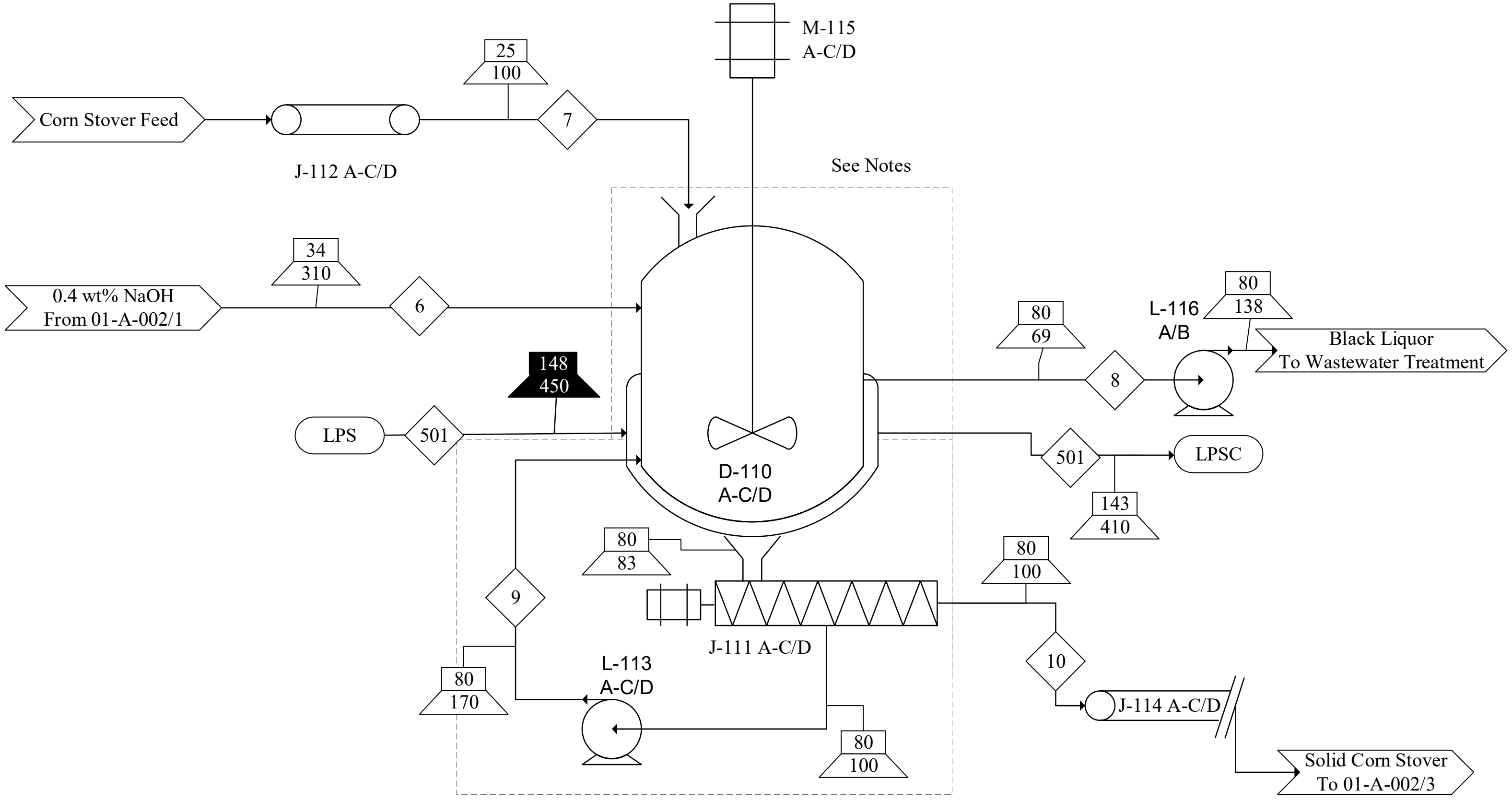
Process Flow Diagram

Title

Process Area 01: Corn Stover Preteatment

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	01-A-002/1	0	373

D-110 A-C/D	J-111 A-C/D	J-112 A-C/D	L-113 A-C/D	J-114 A-C/D	M-115 A-C/D	L-116 A/B
Deacetylation Vessel Height = 8.5 m Diameter = 4.3 m	Deacetylation Screw Conveyor Duty = 3.7 kW Length = 15 m Width = 0.45 m	Pre-Deacetylation Corn Stover Belt Conveyor Duty = 2.2 kW Length = 15 m Width = 0.45 m	Pre-Deacetylation Corn Stover Pump Duty = 0.74 kW	Post-Deacetylation Corn Stover Belt Conveyor Duty = 2.2 kW Length = 15 m Width = 0.45 m	Deacetylation Mixer Agitation Power= 96 kW	Black Liquor Pump Duty = 3.7 kW



Components/Stream	6	7	8	9	10	501
Acetate	0	0	430	0	0	0
Ash	0	0	1,000	0	0	0
Corn Stover	0	27,000	0	0	21,000	0
Extractables	0	0	4,000	0	0	0
Lignin	0	0	860	0	0	0
Sodium Hydroxide	450	0	450	0	0	0
Water	110,000	0	110,000	4,200	0	10,000
Xylan	0	0	100	0	0	0
Total	**110,000**	**27,000**	**120,000**	**4,200**	**21,000**	**10,000**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. D-100 is a semi-batch process, see supplemental PFD 01-A-003 for more information.

5. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

- ◇# Stream Identification
- Input/Output
- ■# Temperature (°C), gas
- □# Temperature (°C), liquid
- ▲# Pressure (kPa), gas
- △# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 04-A-002
BFD 01-A-001	PFD 05-A-002
PFD 02-A-002	Supplemental
PFD 03-A-002	PFD 01-A-003

Drawing Type
Process Flow Diagram

Title

Process Area 01: Corn Stover Preteatment

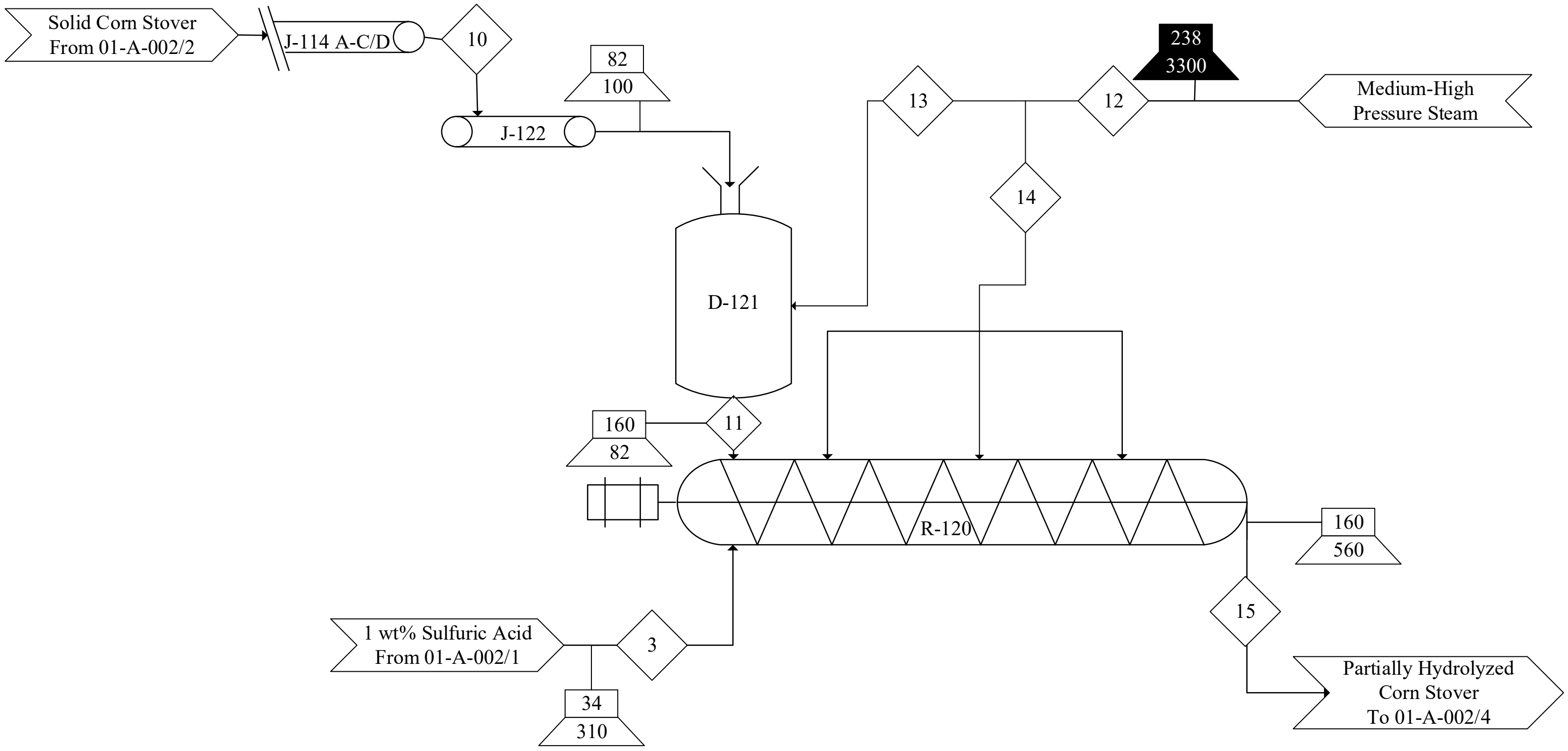
By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	01-A-002/2	0	374

D-121
Acid Presteamer
Height = 5.8 m
Diameter = 2.9 m

R-120
Acid Hydrolysis
Reactor
Length = 10 m
Diameter = 1.0 m
Power = 29 kW

J-122
Pre-Hydrolysis
Corn Stover Belt
Conveyor
Power = 1.5 kW
Length = 15 m
Width = 6.1 m

J-114 A-C/D
Post-Deacetylation
Corn Stover Belt
Conveyor
Duty = 2.2 kW
Length = 15 m
Width = 6.1 m



Components/Stream	3	10	11	12	13	14	15
Acetic Acid	0	0	0	0	0	0	59
Corn Stover	0	21,000	21,000	0	0	0	14,000
Furfural	0	0	0	0	0	0	190
Glucose	0	0	0	0	0	0	1,100
Glucose Oligomer	0	0	0	0	0	0	280
HMF	0	0	0	0	0	0	260
Sulfuric Acid	180	0	0	0	0	0	180
Water	18,000	0	1,000	40,000	1,000	39,000	59,000
Xylose	0	0	0	0	0	0	5,500
Xylose Oligomer	0	0	0	0	0	0	82

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. High pressure steam is directly injected into the reactor to keep it isothermal.
5. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

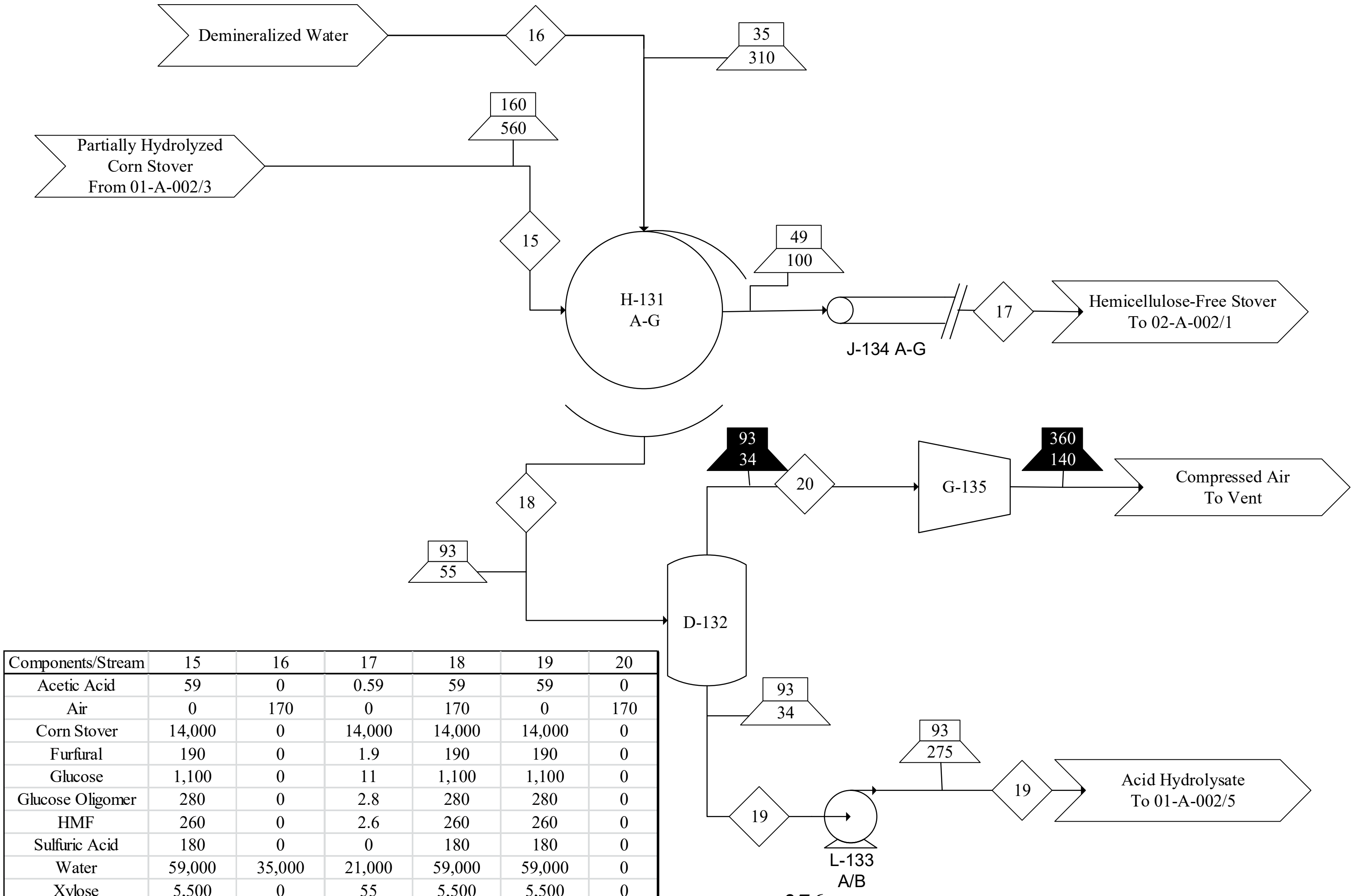
Drawing Type
Process Flow Diagram

Title

Process Area 01: Corn Stover Preteatment

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	01-A-002/3	0	375

H-131, A-G	D-132	L-133 A/B	J-134 A-G	G-135
Acid Rotary Vacuum Filter	Acid Filtration Knockout Drum	Acid Filtration Pump	Acid Filtration Corn Stover Belt Conveyor	Air Compressor
Duty = 1.5 kW	Diameter = 4.0 m	Duty = 19 kW	Power = 1.5 kW	Duty = 13 kW
Volume = 15 m^3	Height =7.9 m		Length = 150 m	
			Width =0.61 m	



Components/Stream	15	16	17	18	19	20
Acetic Acid	59	0	0.59	59	59	0
Air	0	170	0	170	0	170
Corn Stover	14,000	0	14,000	14,000	14,000	0
Furfural	190	0	1.9	190	190	0
Glucose	1,100	0	11	1,100	1,100	0
Glucose Oligomer	280	0	2.8	280	280	0
HMF	260	0	2.6	260	260	0
Sulfuric Acid	180	0	0	180	180	0
Water	59,000	35,000	21,000	59,000	59,000	0
Xylose	5,500	0	55	5,500	5,500	0
Xylose Oligomer	82	0	82	82	82	0
Total	**82,000**	**35,000**	**35,000**	**82,000**	**82,000**	**170**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. H-121 is run at vacuum conditions to remove liquids from the solids.
5. Process water washes the stover to ensure high product recovery.
6. Pumps have an efficiency of 65%
7. Compressors have an efficiency of 65%.

Special Symbols / Nomenclature

Symbol	Meaning
◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

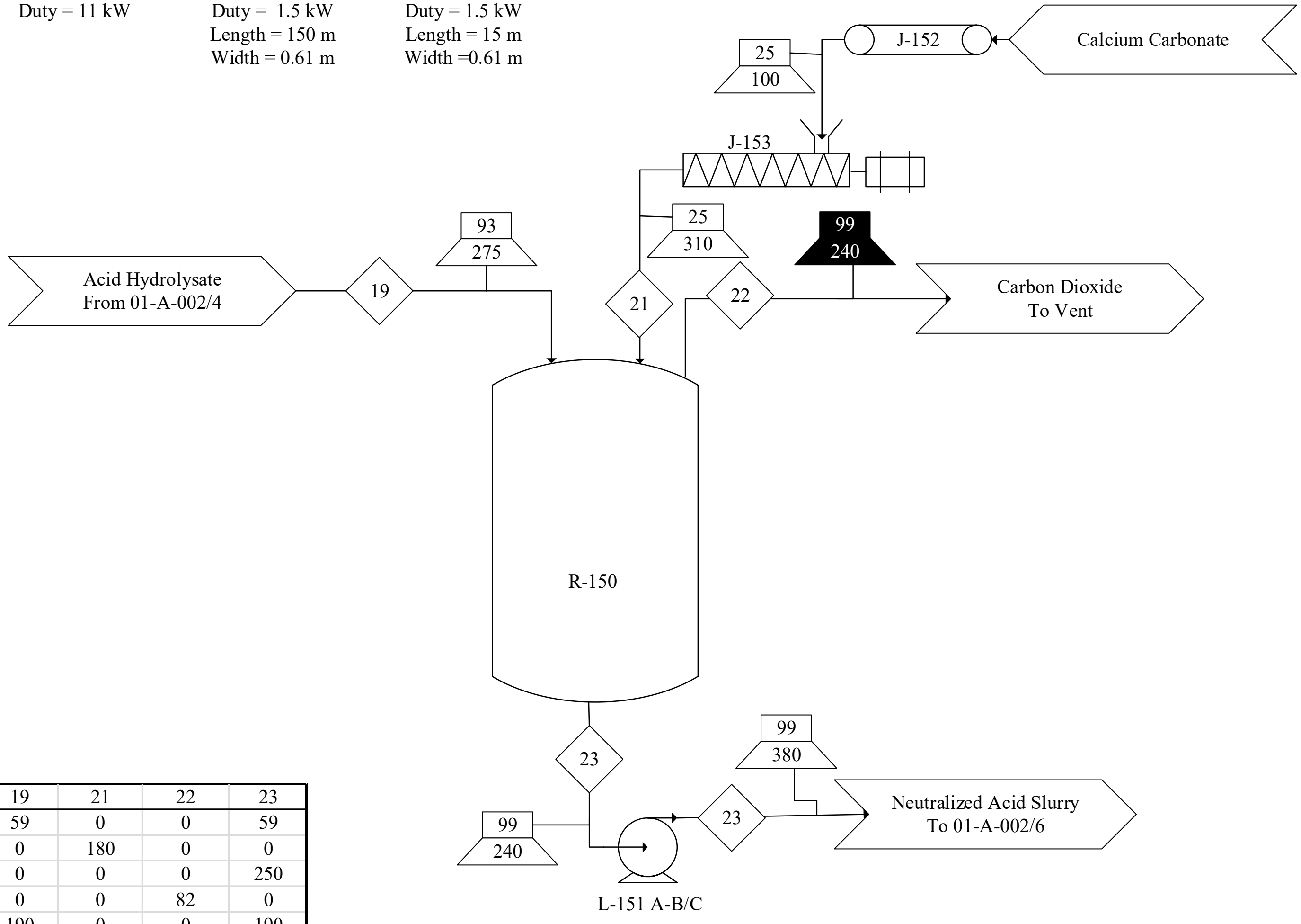
Drawing Type
Process Flow Diagram

Title

Process Area 01: Corn Stover Pretreatment

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	01-A-002/4	0	376

R-150	L-151 A-B/C	J-152	J-153
Neutralization Reactor	Acid Phase Separation Slurry Pump	Calcium Carbonate Feed Conveyor	Calcium Carbonate Screw Conveyor
Diameter =5.8 m	Duty = 11 kW	Duty = 1.5 kW	Duty = 1.5 kW
Height = 11 m		Length = 150 m	Length = 15 m
		Width = 0.61 m	Width =0.61 m



Components/Stream	19	21	22	23
Acetic Acid	59	0	0	59
Calcium Carbonate	0	180	0	0
Calcium Sulfate	0	0	0	250
Carbon Dioxide	0	0	82	0
Furfural	190	0	0	190
Glucose	1,100	0	0	1,100
Glucose Oligomer	280	0	0	280
HMF	260	0	0	260
Sulfuric Acid	180	0	0	0
Water	59,000	0	0	59,000
Xylose	5,500	0	0	5,500
Xylose Oligomer	82	0	0	82
Total	**82,000**	**180**	**42**	**82,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ # Stream Identification

Input/Output

■ # Temperature (°C), gas

☐ # Temperature (°C), liquid

Pressure (kPa), gas

Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001 PFD 04-A-002
PFD 02-A-002 PFD 05-A-002
PFD 03-A-002

Drawing Type

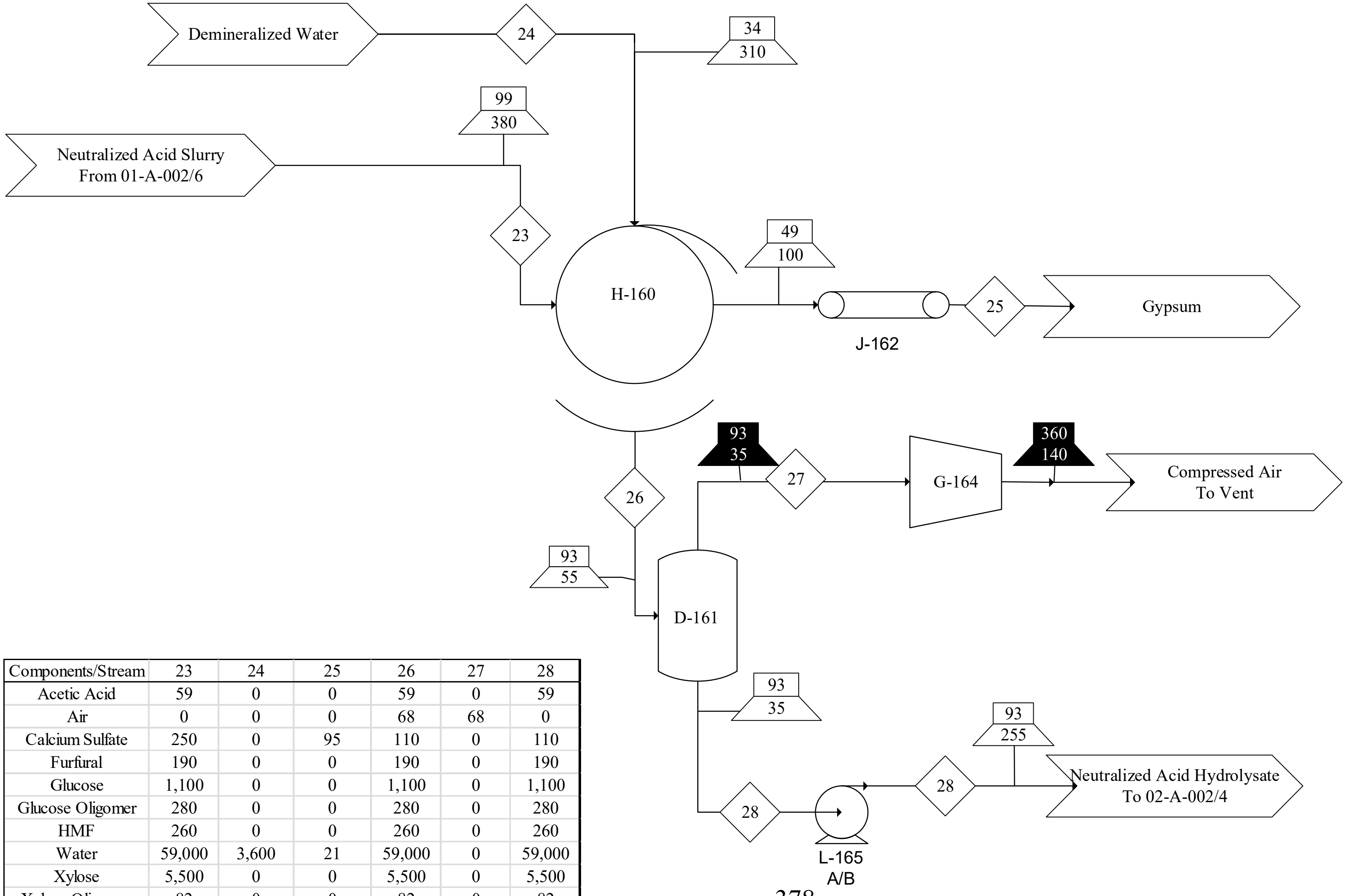
Process Flow Diagram

Title

Process Area 01: Corn Stover Pretreatment

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	01-A-002/5	0	377

H-160	D-161	J-162	L-165 A/B	G-164
Gypsum Rotary Vacuum Filtration	Gypsum Filtration Knockout Drum	Gypsum Filtration Corn Stover Belt Conveyor	Gypsum Filtration Pump	Gypsum Filtration Air Compressor
Duty = 0.74 kW	Diameter = 3.1 m	Power = 1.5 kW	Duty = 46 kW	Duty = 4.4 kW
Volume = 5.1 m^3	Height = 6.1 m	Length = 150 m		
		Width = 0.61 m		



Components/Stream	23	24	25	26	27	28
Acetic Acid	59	0	0	59	0	59
Air	0	0	0	68	68	0
Calcium Sulfate	250	0	95	110	0	110
Furfural	190	0	0	190	0	190
Glucose	1,100	0	0	1,100	0	1,100
Glucose Oligomer	280	0	0	280	0	280
HMF	260	0	0	260	0	260
Water	59,000	3,600	21	59,000	0	59,000
Xylose	5,500	0	0	5,500	0	5,500
Xylose Oligomer	82	0	0	82	0	82
Total	**82,000**	**3,600**	**120**	**82,000**	**68**	**82,000**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
□ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

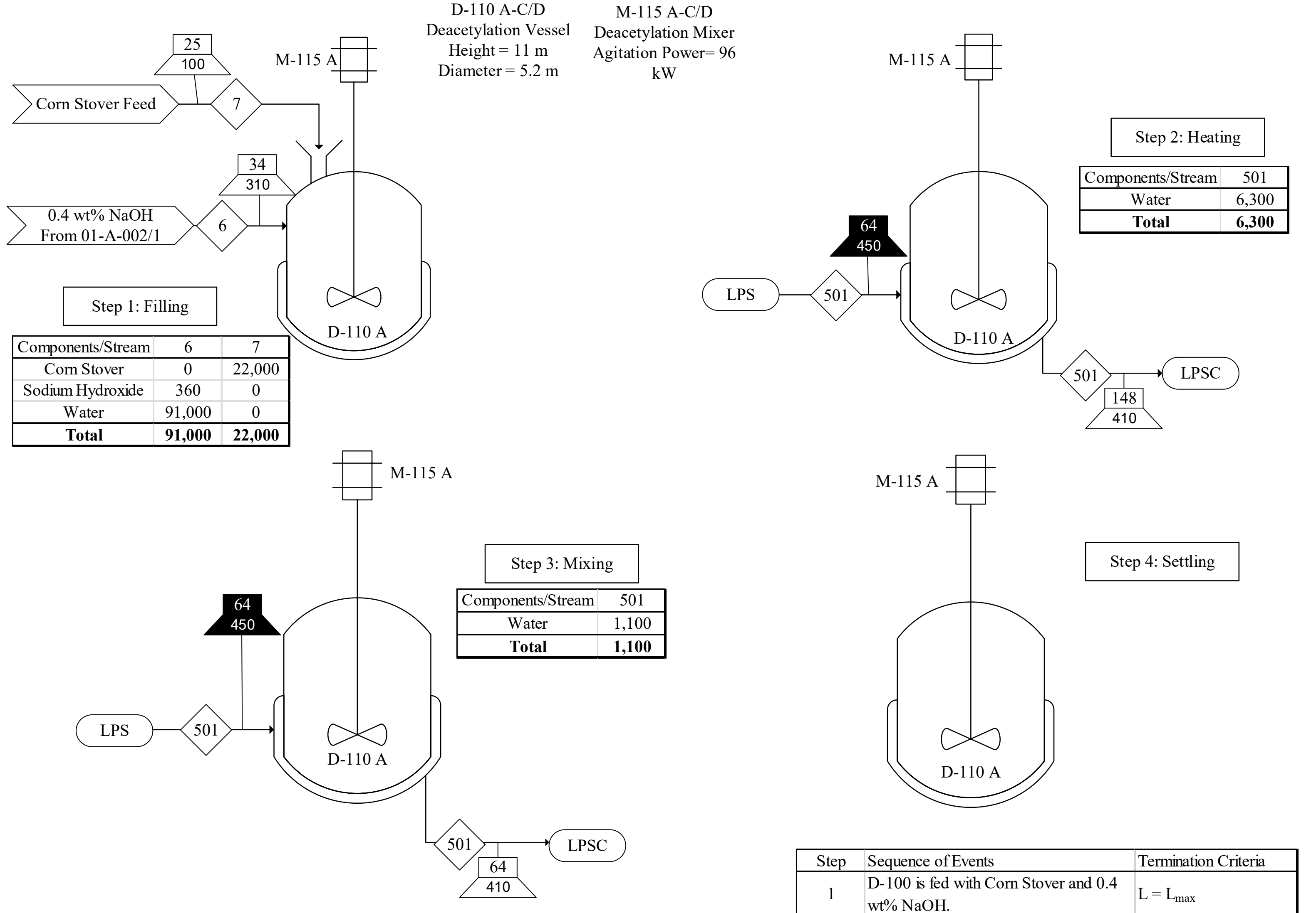
Drawing Type

Process Flow Diagram

Title

Process Area 01: Corn Stover Pretreatment

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	01-A-002/6	0	378



Step 1: Filling

Components/Stream	6	7
Corn Stover	0	22,000
Sodium Hydroxide	360	0
Water	91,000	0
Total	**91,000**	**22,000**

Step 2: Heating

Components/Stream	501
Water	6,300
Total	**6,300**

Step 3: Mixing

Components/Stream	501
Water	1,100
Total	**1,100**

Step	Sequence of Events	Termination Criteria
1	D-100 is fed with Corn Stover and 0.4 wt% NaOH.	$L = L_{max}$
2	D-100 is heated with steam until it reaches 180° F	$T = T_{operating}$
3	D-100 is held at the operating temperature and is continually stired	$C_{Ash\ (aq)} > C_{Ash\ (aq),\ max}$
4	Stirring is stopped and the solids are allowed to settle to the bottom	$\rho_{(aq)} < \rho_{(aq),\ min}$

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 10 minutes
5. Step 2 is 10 minutes
6. Step 3 is 60 minutes
7. Step 4 is 5 minute
8. Three vessels are used in parallel for continual deacetylation.
9. Each vessel cycle are staggered by 0.8 hours.

Special Symbols / Nomenclature

◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

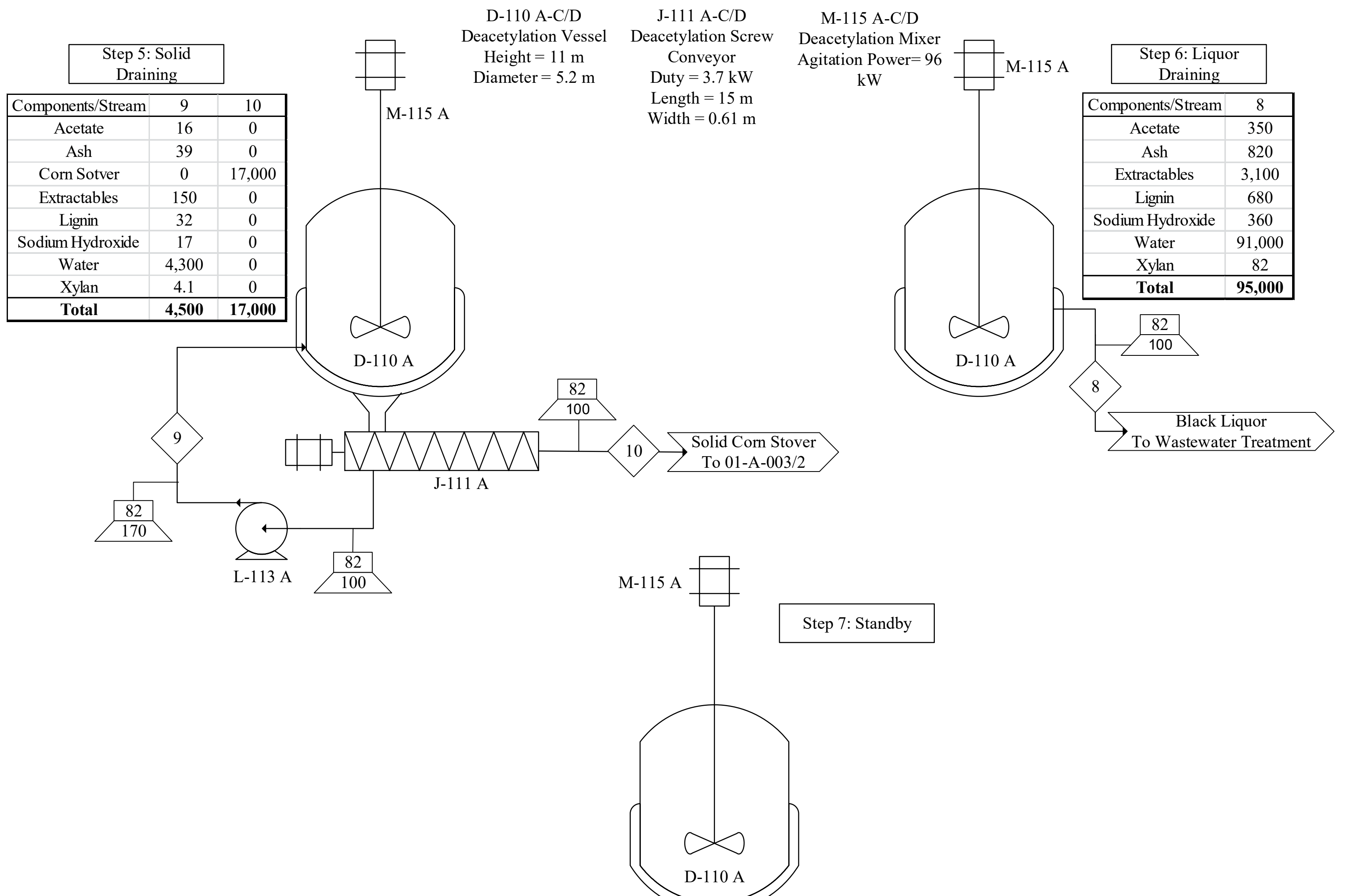
Drawing Type

Supplemental Process Flow Diagram

Title

Deacetylation Semi-Batch Steps

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	01-A-003/1	0	379



Step 5: Solid Draining

Components/Stream	9	10
Acetate	16	0
Ash	39	0
Corn Sotver	0	17,000
Extractables	150	0
Lignin	32	0
Sodium Hydroxide	17	0
Water	4,300	0
Xylan	4.1	0
Total	**4,500**	**17,000**

Step 6: Liquor Draining

Components/Stream	8
Acetate	350
Ash	820
Extractables	3,100
Lignin	680
Sodium Hydroxide	360
Water	91,000
Xylan	82
Total	**95,000**

Step	Sequence of Events	Termination Criteria
5	Solids are screw conveyed from the bottom of the tank. Recovered liquids are recycled back to the vessel.	$L_{solids} < L_{solids, min}$
6	The remaining liquids within the vessel are drained.	$L_{liquids} < L_{liquids, min}$
7	D-100 is on standby.	$C_{Ash (aq)} > 1/2\ C_{Ash (aq), max}$ of vessel in Step 3

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 5 is 48 minutes
5. Step 6 is 10 minutes
6. Step 7 is 1 minute
7. Three vessels are used in parallel for continual deacetylation.

Special Symbols / Nomenclature

◇ #	Stream Identification
⇨	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

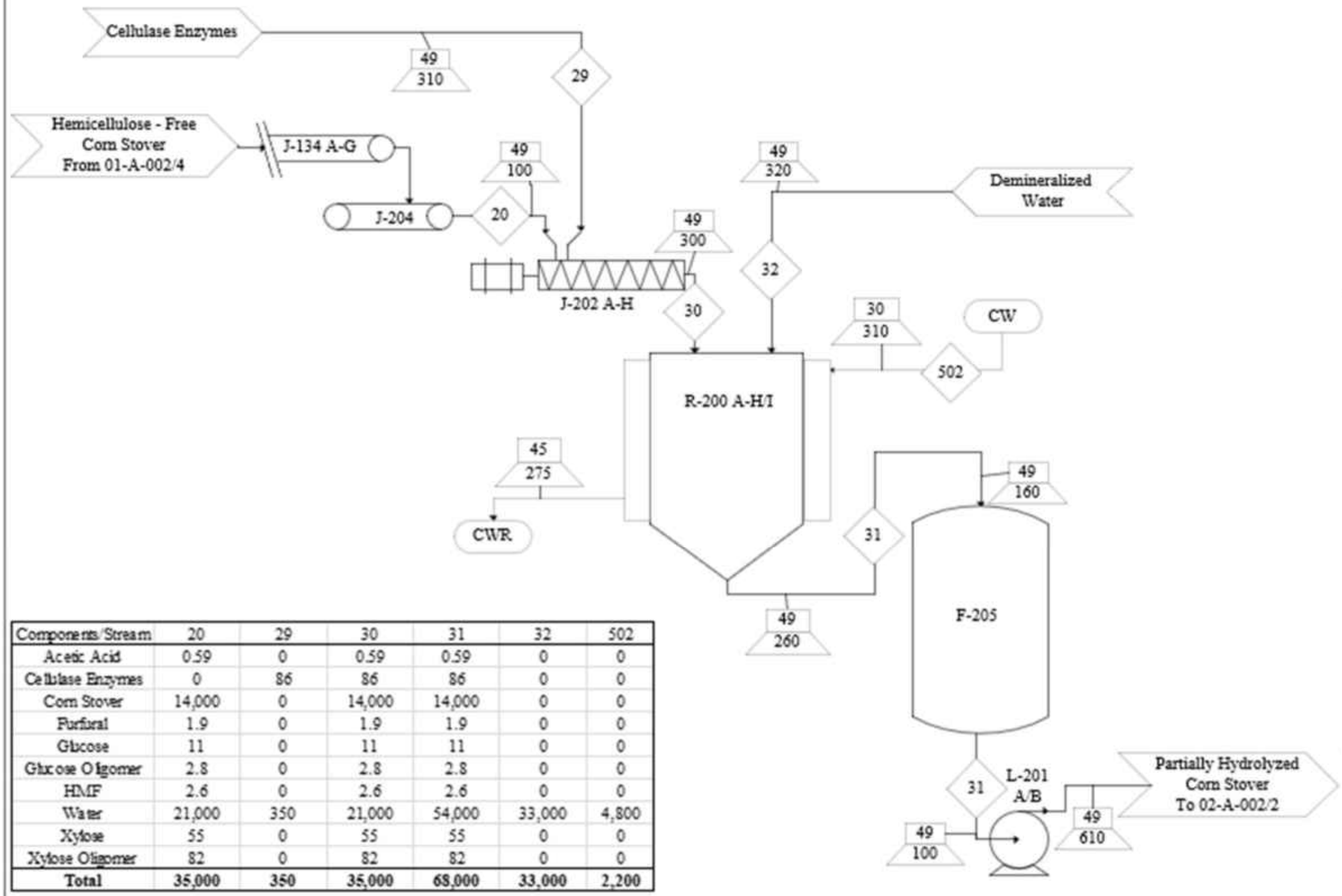
Supplemental Process Flow Diagram

Title

Deacetylation Semi-Batch Steps

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	01-A-003/2	0	380

R-200 A-H/I	L-201 A/B	J-202 A-H	J-204	F-205	J-134 A-G
Continuous, High-Solids Enzymatic Reactor	Enzymatic Slurry Pump	Pre-Enzymatic Screw Conveyor	Corn Stover Belt Conveyor	Pre-Batch Hold Up	Acid Filtration Corn Stover Belt Conveyor
Height = 17 m	Duty = 33 kW	Duty = 2.2 kW	Duty = 1.5 kW	Height: 18 m	Power = 1.5 kW
Diameter= 4.3 m		Length = 15 m	Length = 15 m	Diameter: 9.1 m	Length = 150 m
		Width = 0.45 m	Width = 0.61 m		Width = 0.61 m



Components/Stream	20	29	30	31	32	502
Acetic Acid	0.59	0	0.59	0.59	0	0
Cellulase Enzymes	0	86	86	86	0	0
Corn Stover	14,000	0	14,000	14,000	0	0
Furfural	1.9	0	1.9	1.9	0	0
Glucose	11	0	11	11	0	0
Glucose Oligomer	2.8	0	2.8	2.8	0	0
HMF	2.6	0	2.6	2.6	0	0
Water	21,000	350	21,000	54,000	33,000	4,800
Xylose	55	0	55	55	0	0
Xylose Oligomer	82	0	82	82	0	0
Total	**35,000**	**350**	**35,000**	**68,000**	**33,000**	**2,200**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Pumps have an efficiency of 65%

5. Due to high solids concentration enzymatic reaction starts in a flow reactor. When the stream reaches the bottom of the reactor (24 hrs) it is a pump-able concertation.

6. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

◇ #	Stream Identification
	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 02-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 03-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 02: Glucose Recovery

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	02-A-002/1	0	381

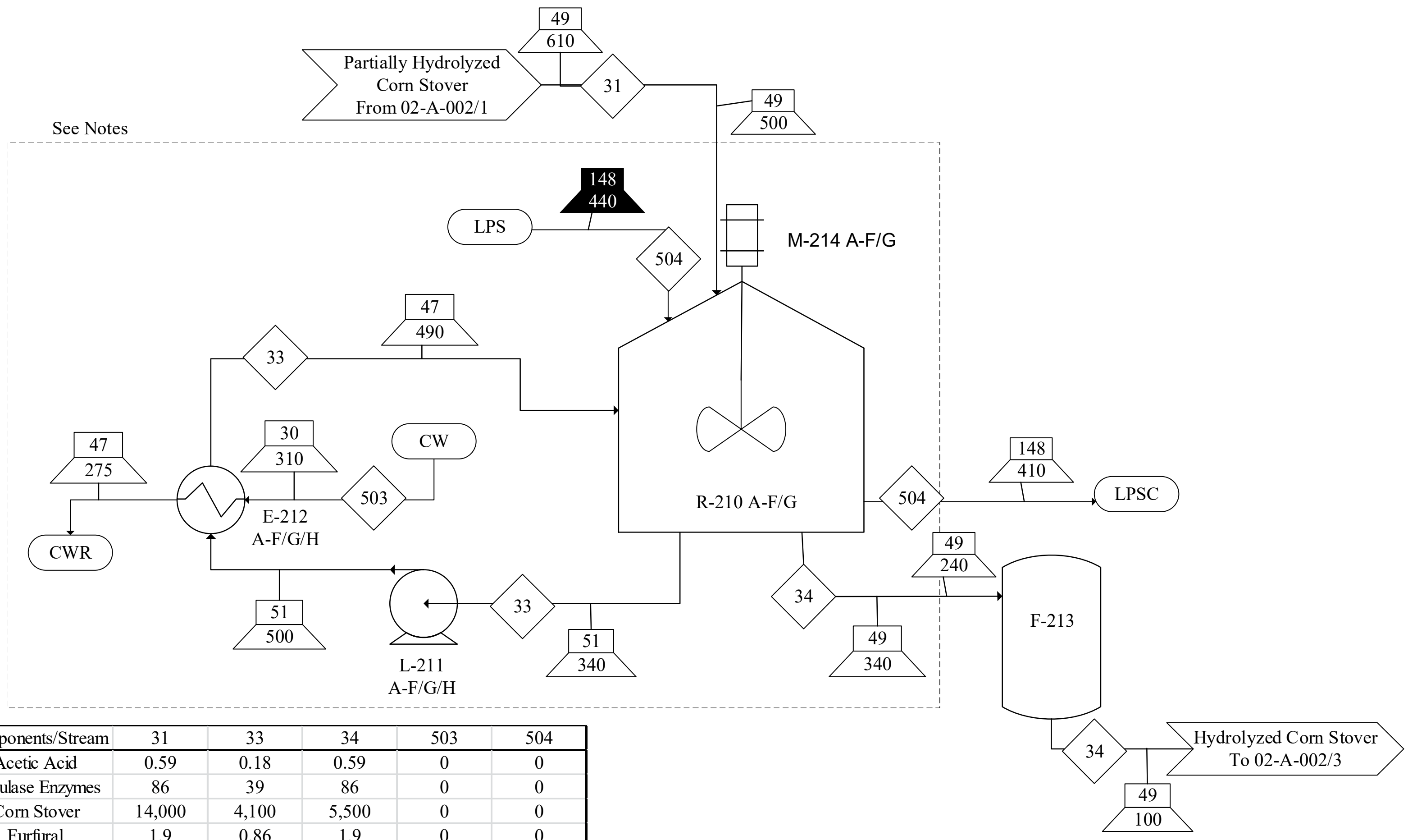
R-210 A-F/G
Batch Enzymatic Reactor
Height = 16 m
Diameter= 8.2 m

L-211 A-F/G/H
Batch Hydrolysis Recycle Pump
Duty = 1.5 kW

E-212 A-F/G/H
Batch Hydrolysis Recycle Heat Exchanger
Duty = 0.057 GJ/hr
Area = 1.7 m^2

F-213
Post-Batch Hold Up
Height: 18 m
Diameter: 9.1 m

M-214 A-F/G
Batch Hydrolysis Mixer
Agitation Power= 430 kW



Components/Stream	31	33	34	503	504
Acetic Acid	0.59	0.18	0.59	0	0
Cellulase Enzymes	86	39	86	0	0
Corn Stover	14,000	4,100	5,500	0	0
Furfural	1.9	0.86	1.9	0	0
Glucose	11	820	8,100	0	0
Glucose Oligomer	2.8	33	350	0	0
HMF	2.6	1	2.6	0	0
Water	54,000	18,000	54,000	9,500	250
Xylose	55	18	55	0	0
Xylose Oligomer	82	0	82	0	0
Total	**68,000**	**23,000**	**68,000**	**9,500**	**250**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. R-210 is a semi-batch process, more information can be seen on Supplemental PFD 02-A-003.
6. Pressure drops in stream 38 due to the 54 foot height increase to get to the top of the reactor.

Special Symbols / Nomenclature

◇# Stream Identification
▷ Input/Output
■# Temperature (°F), gas
☐# Temperature (°F), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 04-A-002
BFD 02-A-001	PFD 05-A-002
PFD 01-A-002	Supplemental
PFD 03-A-002	PFD 02-A-003

Drawing Type

Process Flow Diagram

Title

Process Area 02: Glucose Recovery

By/Date	DWG #/ Sheet #	Rev	
AJK 7/05/20	02-A-002/2	0	382

H-220, A-C
Enzymatic Rotary
Vacuum Filtration
Duty = 1.5 kW
Volume = 50 m^3

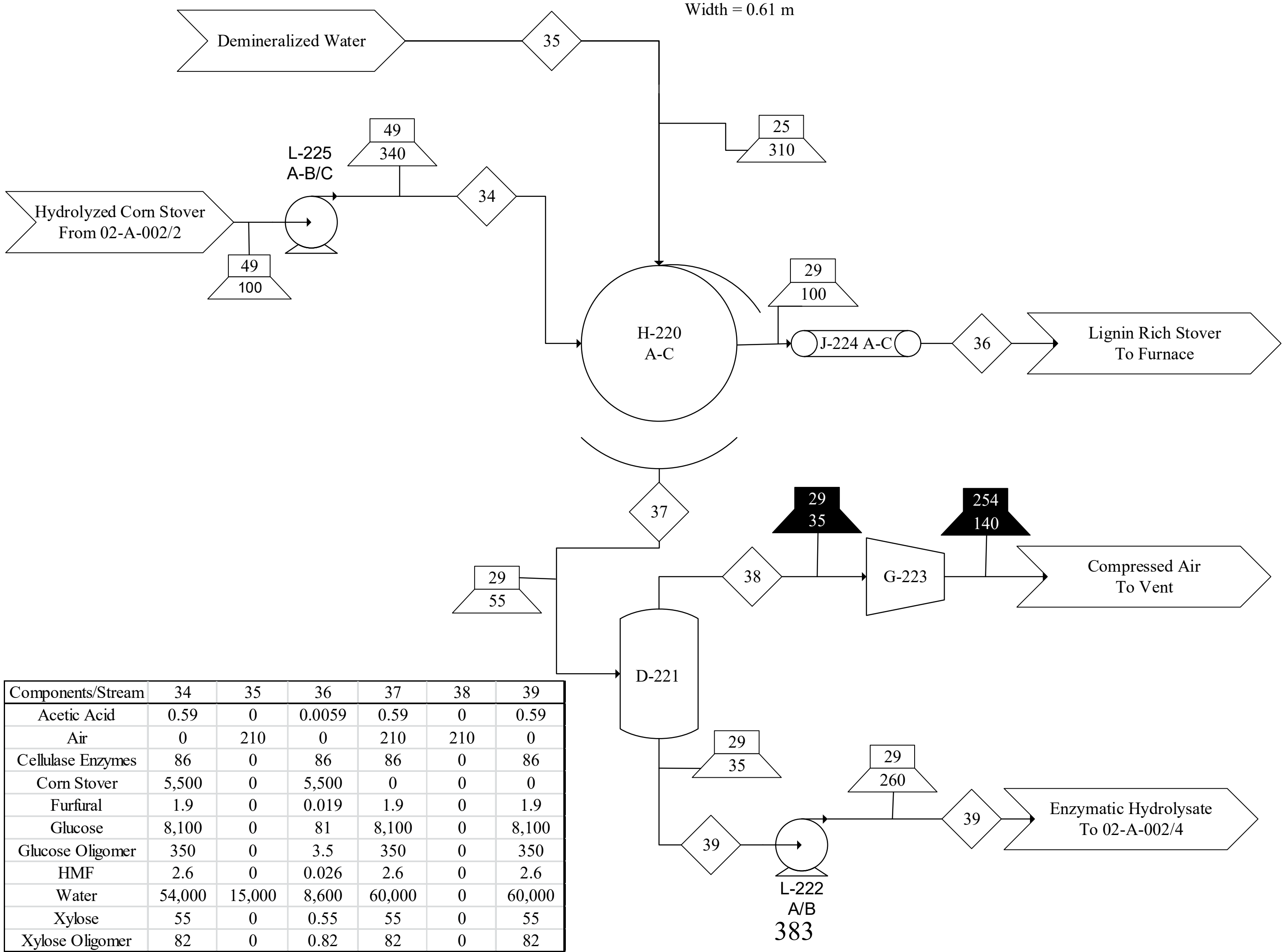
D-221
Enzymatic Filtration
Knockout Drum
Diameter = 3.1 m
Height = 6.7 m

L-222 A/B
Enzymatic Filtration
Pump
Duty = 4.5 kW

G-223
Enzymatic Filtration
Air Compressor
Duty 13 kW

J-224 A-C
Lignin Rich Belt
Conveyor
Power = 1.5 kW
Length = 150 m
Width = 0.61 m

L-225 A-B/C
Enzymatic Filtration
Feed Slurry Pump
Duty =6.7 kW



Components/Stream	34	35	36	37	38	39
Acetic Acid	0.59	0	0.0059	0.59	0	0.59
Air	0	210	0	210	210	0
Cellulase Enzymes	86	0	86	86	0	86
Corn Stover	5,500	0	5,500	0	0	0
Furfural	1.9	0	0.019	1.9	0	1.9
Glucose	8,100	0	81	8,100	0	8,100
Glucose Oligomer	350	0	3.5	350	0	350
HMF	2.6	0	0.026	2.6	0	2.6
Water	54,000	15,000	8,600	60,000	0	60,000
Xylose	55	0	0.55	55	0	55
Xylose Oligomer	82	0	0.82	82	0	82
Total	**68,000**	**15,000**	**14,000**	**68,000**	**210**	**68,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Compressors have an efficiency of 65%.

Special Symbols / Nomenclature

◇ # Stream Identification
Input/Output
■ # Temperature (°C), gas
□ # Temperature (°C), liquid
■ # Pressure (kPa), gas
□ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 02-A-001
PFD 01-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

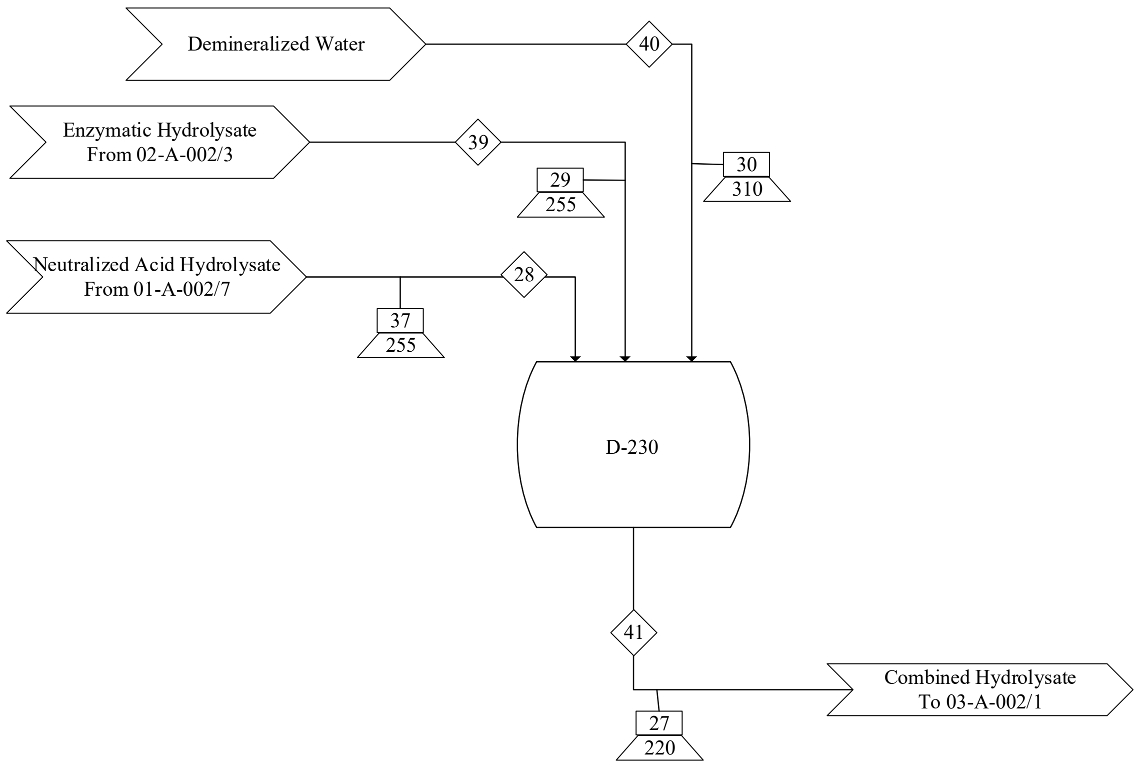
Process Flow Diagram

Title

Process Area 02: Glucose Recovery

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	02-A-002/3	0	383

D-230
Hydrolysate Mixing
Drum
Diameter = 2.5 m
Height = 4.9 m



Components/Stream	28	39	40	41
Acetic Acid	59	0.59	0	59
Calcium Sulfate	110	0	0	110
Furfural	190	1.9	0	190
Glucose	1,100	8,100	0	9,200
Glucose Oligomer	280	350	0	630
HMF	260	2.6	0	260
Water	59,000	60,000	23,000	140,000
Xylose	5,500	55	0	5,500
Xylose Oligomer	82	82	0	180
Total	**82,000**	**68,000**	**23,000**	**150,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 02-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 03-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 02: Glucose Recovery

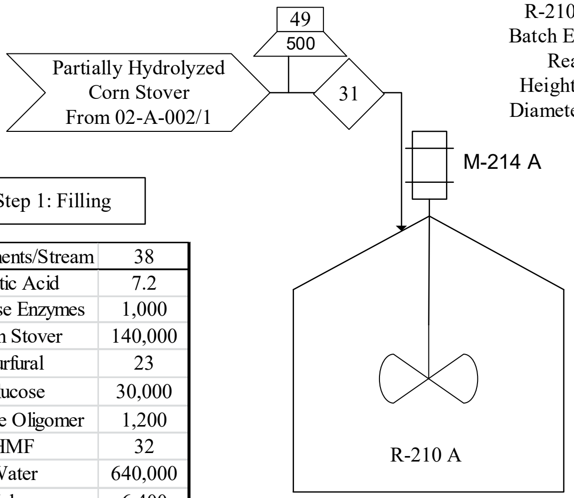
By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	02-A-002/4	0	384

R-210 A-F/G
Batch Enzymatic Reactor
Height = 16 m
Diameter= 8.2 m

L-211 A-F/G/H
Batch Hydrolysis Recycle Pump
Duty = 1.5 kW

E-212 A-F/G/H
Batch Hydrolysis Recycle Heat Exchanger
Duty = 0.056 GJ/hr
Area = 1.7 m^2

M-214 A-F/G
Batch Hydrolysis Mixer
Agitation Power= 430 kW

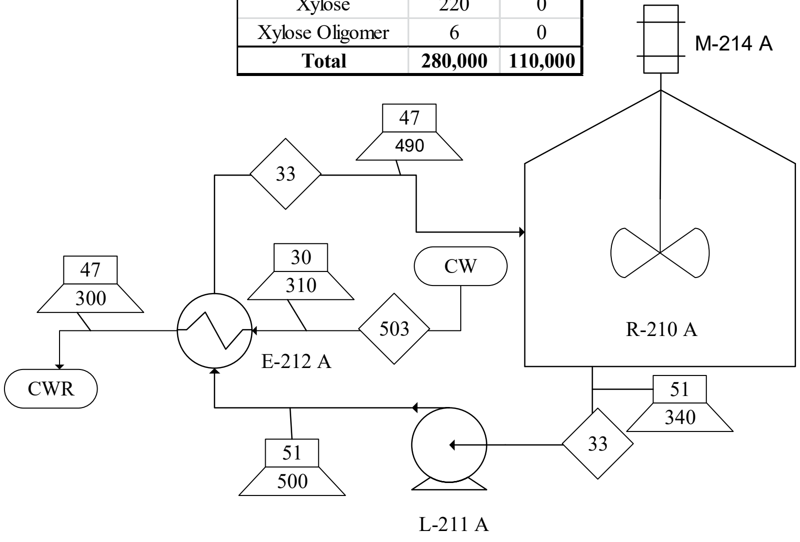


Step 1: Filling

Components/Stream	38
Acetic Acid	7.2
Cellulase Enzymes	1,000
Corn Stover	140,000
Furfural	23
Glucose	30,000
Glucose Oligomer	1,200
HMF	32
Water	640,000
Xylose	6,400
Xylose Oligomer	9.5
Total	**820,000**

Step 2: Hydrolysis

Components/Stream	33	503
Acetic Acid	5.5	0
Cellulase Enzymes	350	0
Corn Stover	50,000	0
Furfural	7.7	0
Glucose	10,000	0
Glucose Oligomer	400	0
HMF	11	0
Water	220,000	110,000
Xylose	220	0
Xylose Oligomer	6	0
Total	**280,000**	**110,000**



Step	Sequence of Events	Termination Criteria
1	R-210 is filled with partially hydrolyzed corn stover from the continuous reactor (R-200).	$L = L_{max}$
2	R-210 is stirred and its contents recycled to maintain vessel temperature.	$C_{Glucose\ (aq)} > C_{Gkucose\ (aq),\ min}$
3	R-210 is drained of its contents.	$L_{liquids} < L_{liquids,\ min}$
4	R-210 is filled with low pressure steam to steralize the reactor.	$P_{vessel} = P_{vessel,\ max}$
5	R-210 is on standby.	Vessel ahead is 11 hours into reaction

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 1 hour
5. Step 2 is 60 hours
6. Step 3 is 1 hour
7. Step 4 is 1 hour
8. Step 5 is 3 hours
9. Six reactors are used in parallel for continual hydrolysis.

Special Symbols / Nomenclature

◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

Supplemental Process Flow Diagram

Title

Enzymatic Hydrolysis Semi-Batch Steps

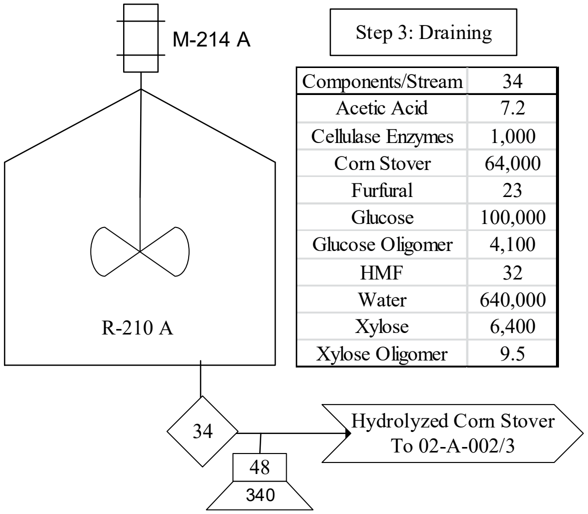
By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	02-A-003/1	0	385

R-210 A-F/G
Batch Enzymatic Reactor
Height = 16 m
Diameter= 8.2 m

E-212 A-F/G/H
Batch Hydrolysis Recycle Heat Exchanger
Duty = 0.057 GJ/hr
Area = 1.7 m^2

M-214 A-F/G
Batch Hydrolysis Mixer
Agitation Power= 430 kW

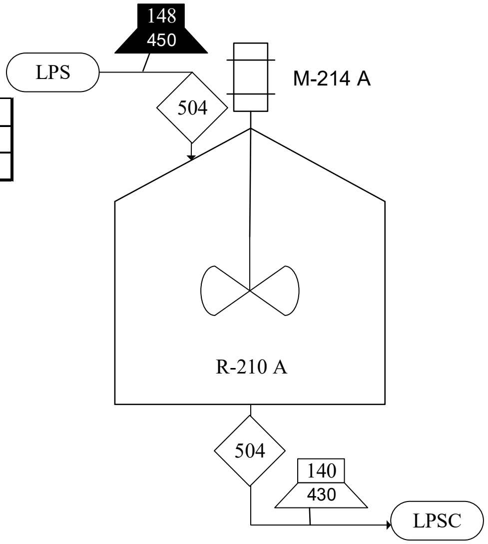
Step 3: Draining



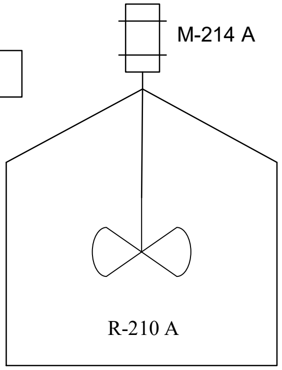
Components/Stream	34
Acetic Acid	7.2
Cellulase Enzymes	1,000
Corn Stover	64,000
Furfural	23
Glucose	100,000
Glucose Oligomer	4,100
HMF	32
Water	640,000
Xylose	6,400
Xylose Oligomer	9.5

Step 4: Sterilization

Components/Stream	504
Water	2,700
Total	**2,700**



Step 5: Standby



Step	Sequence of Events	Termination Criteria
1	R-210 is filled with partially hydrolyzed corn stover from the continuous reactor (R-200).	$L = L_{max}$
2	R-210 is stirred and its contents recycled to maintain vessel temperature.	$C_{Glucose\ (aq)} > C_{Gkucose\ (aq),\ min}$
3	R-210 is drained of its contents.	$L_{liquids} < L_{liquids,\ min}$
4	R-210 is filled with low pressure steam to steralize the reactor.	$P_{vessel} = P_{vessel,\ max}$
5	R-210 is on standby.	Vessel ahead is 11 hours into reaction

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 1 hour
5. Step 2 is 60 hours
6. Step 3 is 1 hour
7. Step 4 is 1 hour
8. Step 5 is 3 hours
9. Six reactors are used in parallel for continual hydrolysis.

Special Symbols / Nomenclature

◇ #	Stream Identification
⇨	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

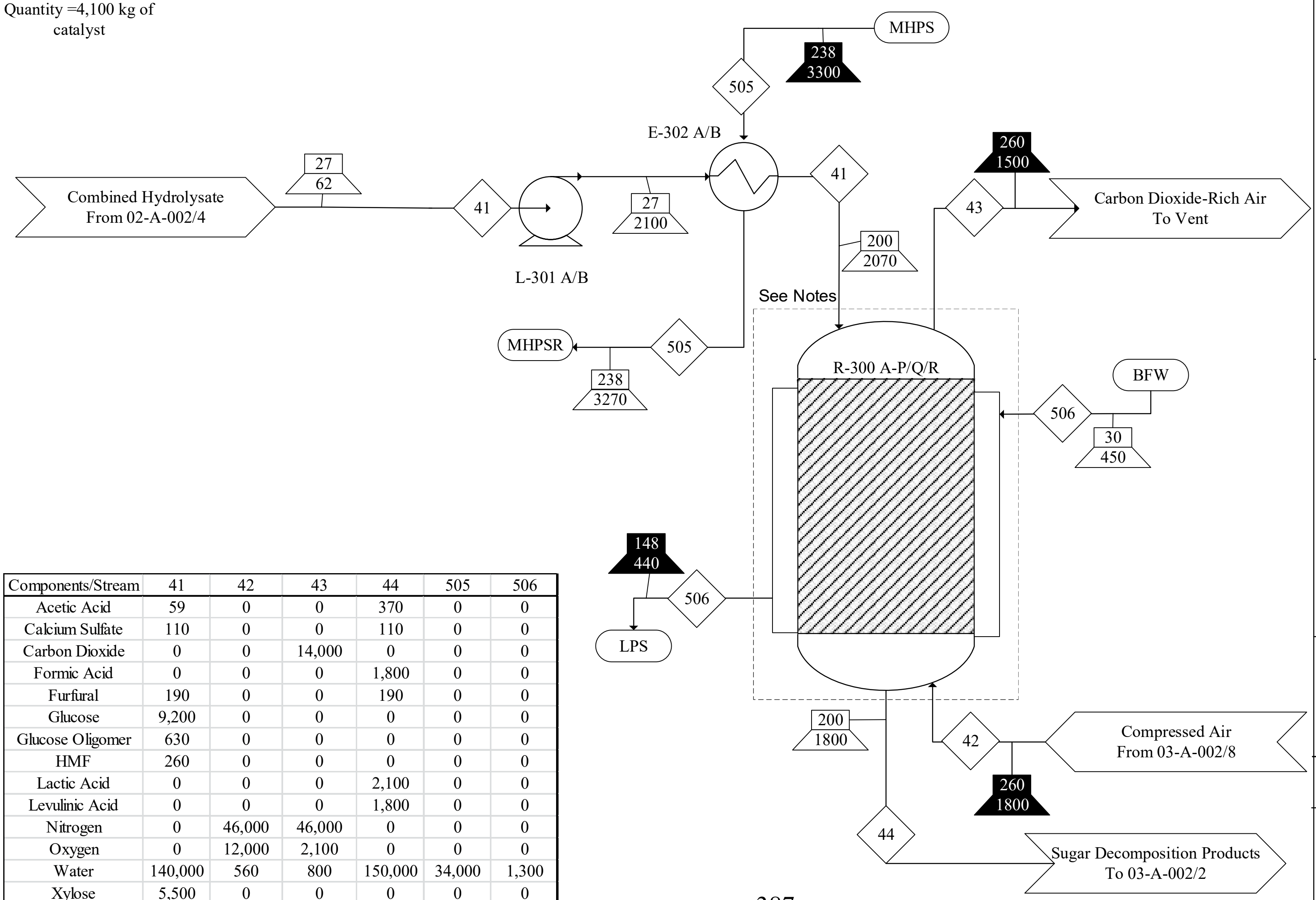
Supplemental Process Flow Diagram

Title

Enzymatic Hydrolysis Semi-Batch Steps

By/Date	DWG # / Sheet #	Rev	Page
AJK 7/05/20	02-A-003/2	0	386

R-300 A-P/Q/R	L-301 A/B	E-302 A/B
Sn2+ Beta Zeolite Packed Bed Reactor	Pre-Catalytic Decomposition Pump	Catalytic Decomposition Preheater
Diameter = 1.8 m	Duty = 140 kW	Duty = 94 GJ/hr
Height = 15 m		Area = 710 m^2
Quantity =4,100 kg of catalyst		



Components/Stream	41	42	43	44	505	506
Acetic Acid	59	0	0	370	0	0
Calcium Sulfate	110	0	0	110	0	0
Carbon Dioxide	0	0	14,000	0	0	0
Formic Acid	0	0	0	1,800	0	0
Furfural	190	0	0	190	0	0
Glucose	9,200	0	0	0	0	0
Glucose Oligomer	630	0	0	0	0	0
HMF	260	0	0	0	0	0
Lactic Acid	0	0	0	2,100	0	0
Levulinic Acid	0	0	0	1,800	0	0
Nitrogen	0	46,000	46,000	0	0	0
Oxygen	0	12,000	2,100	0	0	0
Water	140,000	560	800	150,000	34,000	1,300
Xylose	5,500	0	0	0	0	0
Xylose Oligomer	82	0	0	0	0	0
Total	**150,000**	**58,000**	**62,000**	**150,000**	**34,000**	**1,300**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Due to coke formation, two extra reactors are present and the catalyst will need to be calcined every two weeks.
6. Coke formation makes the reactor a semi-batch process, see drawing 03-A-003 for more details.

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 04-A-002
BFD 03-A-001	PFD 05-A-002
PFD 01-A-002	Supplemental
PFD 02-A-002	PFD 03-A-003

Drawing Type

Process Flow Diagram

Title

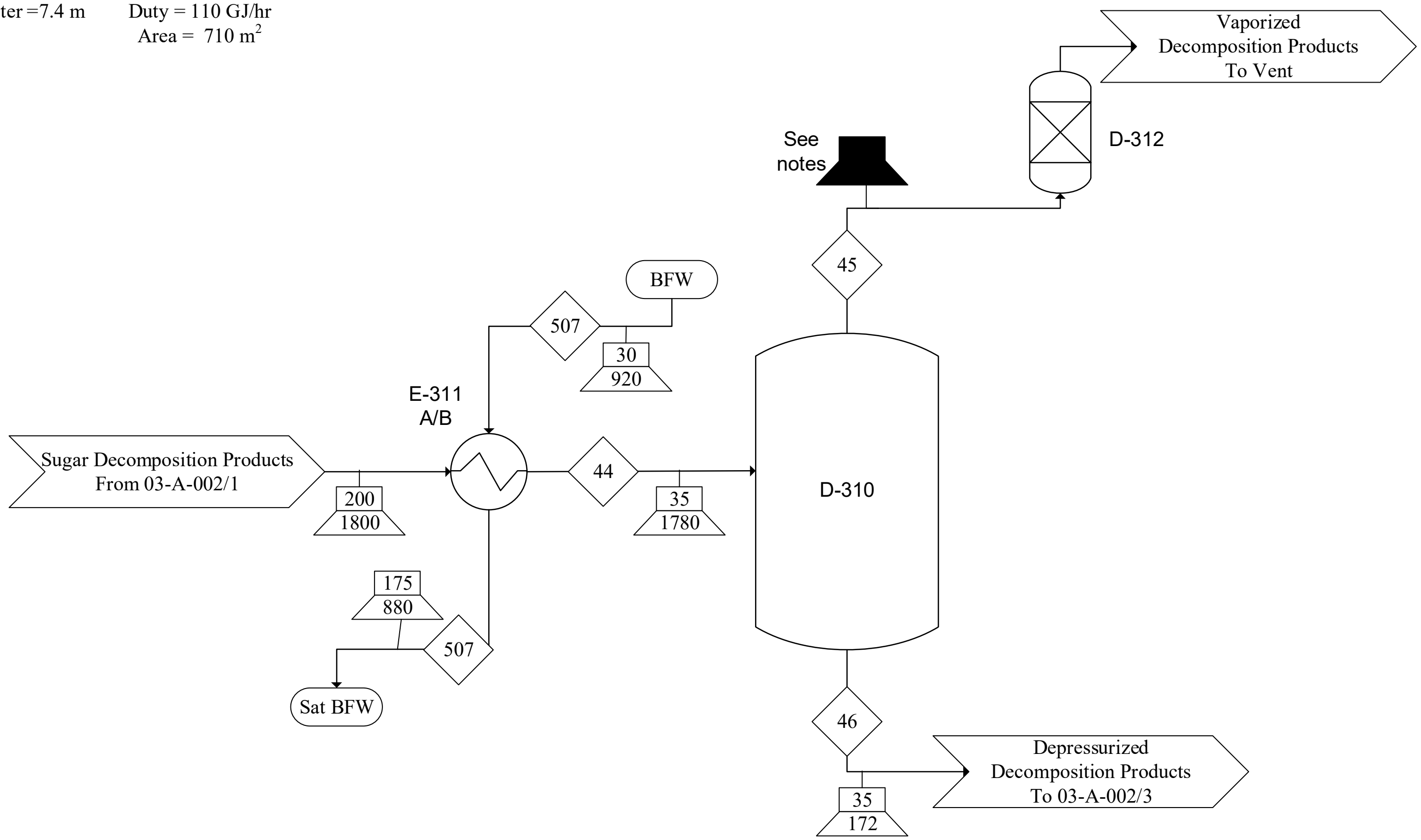
Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	03-A-002/1	0	387

D-310
Furfural Flash Drum
Height = 3.7 m
Diameter =7.4 m

E-311 A-D/E/F
Pre-Furfural
Extraction Cooler
Duty = 110 GJ/hr
Area = 710 m^2

D-312
Chemical Filter
See Notes



Components/Stream	44	45	46	507
Acetic Acid	370	0	370	0
Calcium Sulfate	110	0	110	0
Formic Acid	1,800	Trace	1,800	0
Furfural	190	0	190	0
Lactic Acid	2,100	0	2,100	0
Levulinic Acid	1,800	0	1,800	0
Water	150,000	Trace	150,000	180,000
Total	**150,000**	**Trace**	**150,000**	**180,000**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Stream 73 normally has no flow. Since flow is dependent on process upsets, temperature and pressure are not specified.

5. A chemical filter is added on stream 73 to remove formic acid, allowing water vapor to vent to the atmosphere. Since the flow of stream 73 is particularily small, D-312 will also be small. Design of D-312 will be specified in the future, more detailed designs.

Special Symbols / Nomenclature

◇ # Stream Identification
〉 Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
▲ # Pressure (kPa), gas
△ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 02-A-002

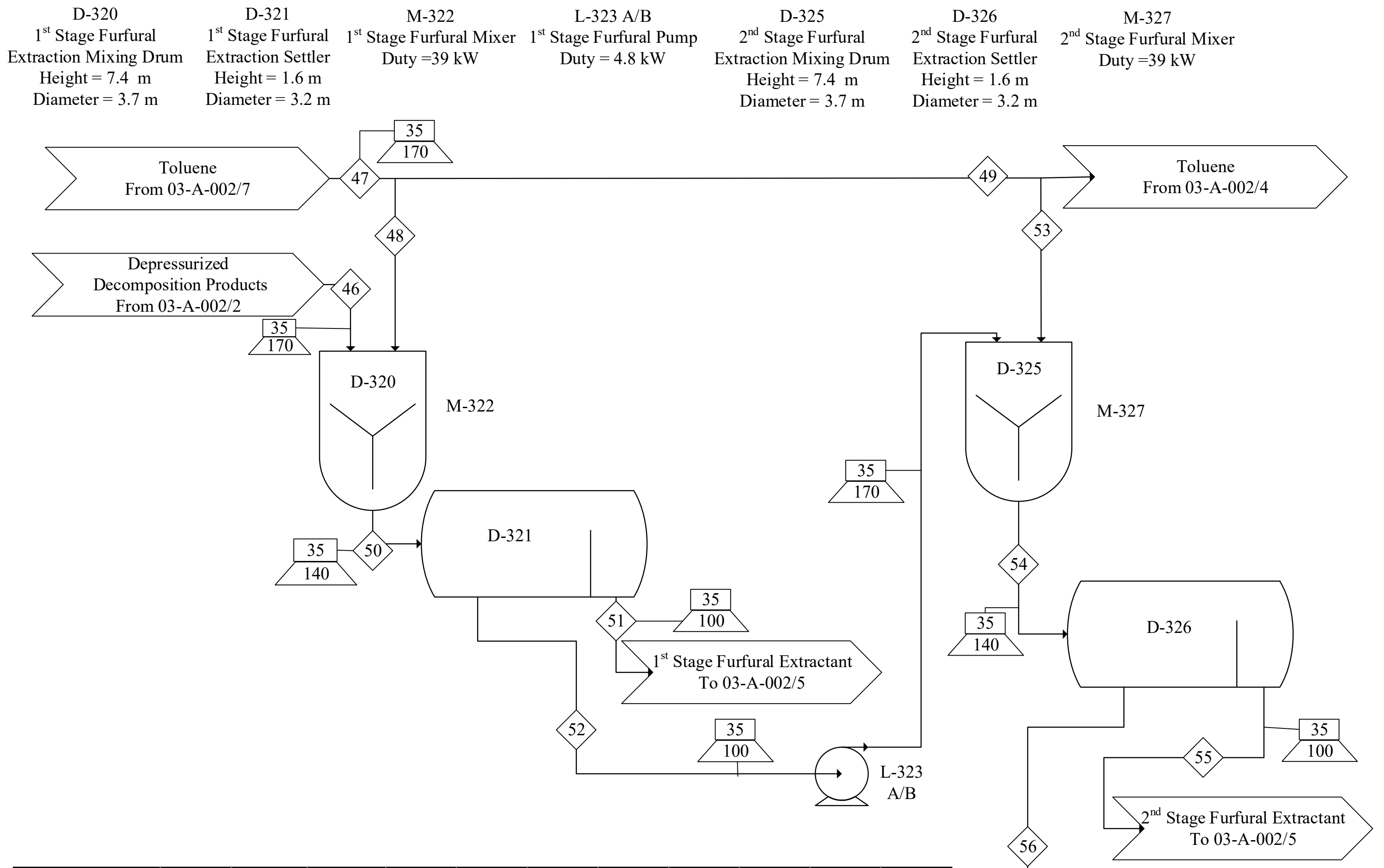
Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	03-A-002/2	0	388



Components/Stream	46	47	48	49	50	51	52	53	54	55	56
Acetic Acid	370	0	0	0	370	0	370	0	370	0	370
Calcium Sulfate	110	0	0	0	110	0	110	0	110	0	110
Formic Acid	1,800	0	0	0	1,800	0	1,800	0	1,800	0	1,800
Furfural	190	0.25	0.21	Trace	190	160	31	0	31	25	6
Lactic Acid	2,100	0	0	0	2,100	0	2,100	0	2,100	0	2,100
Levulinic Acid	1,800	0	0	0	1,800	0	1,800	0	1,800	0	1,800
Toluene	0	330	270	55	270	270	0	43	43	43	0
Water	150,000	0	0	0	150,000	Trace	150,000	0	150,000	Trace	150,000
Total	**150,000**	**330**	**270**	**120**	**150,000**	**430**	**150,000**	**43**	**150,000**	**68**	**150,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ # Stream Identification

Input/Output

■ # Temperature (°C), gas

☐ # Temperature (°C), liquid

■ # Pressure (kPa), gas

☐ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

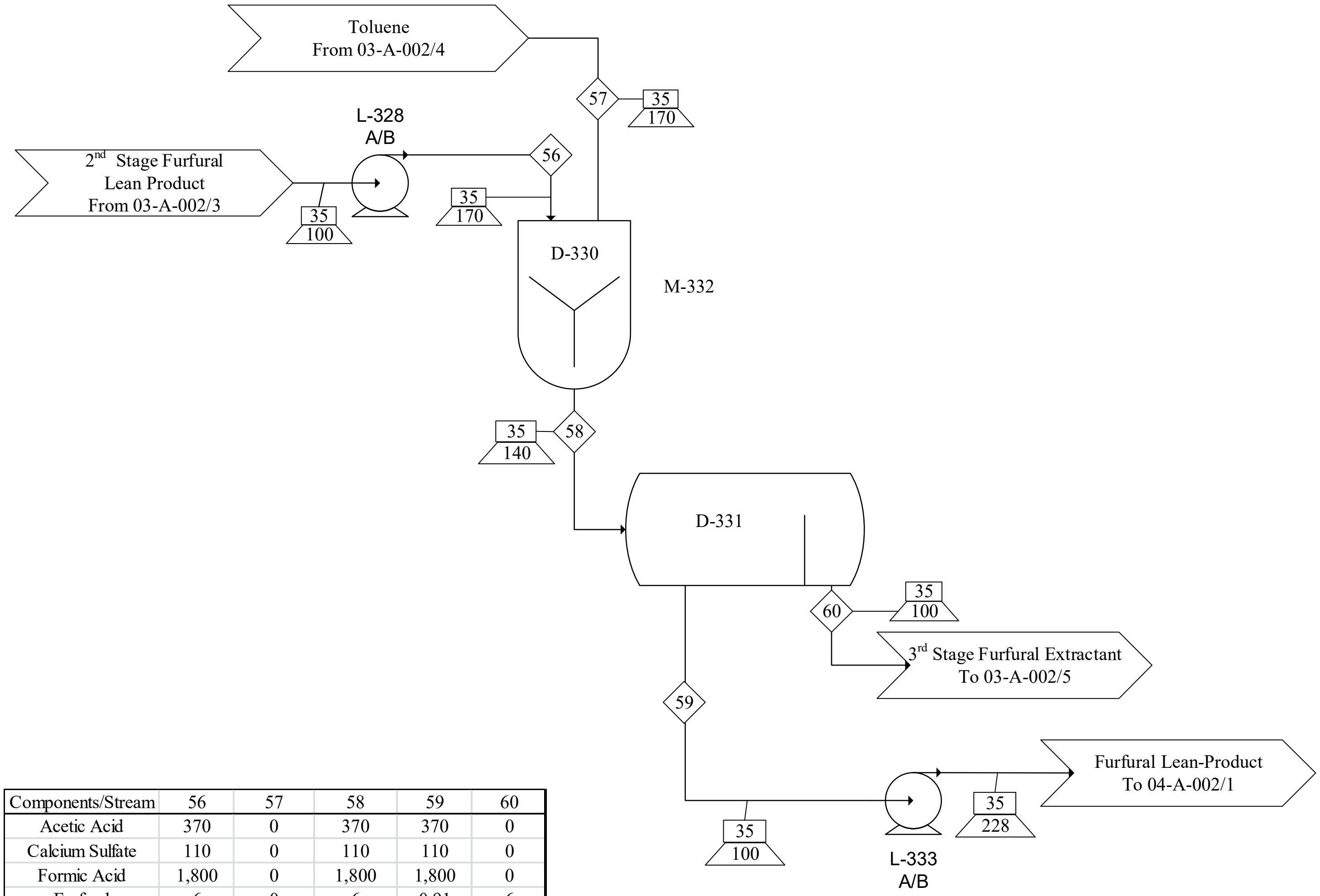
Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	03-A-002/3	0	389

D-330	D-331	M-332	L-328 A/B	L-333 A/B
3rd Stage Furfural Extraction Mixing Drum	3rd Stage Furfural Extraction Settler	3rd Stage Furfural Mixer	2nd Stage Furfural Pump	3rd Stage Furfural Pump
Height = 7.4 m	Height = 1.6 m	Duty = 39 kW	Duty = 4.8 kW	Duty = 14 kW
Diameter = 3.7 m	Diameter = 3.2 m			



Components/Stream	56	57	58	59	60
Acetic Acid	370	0	370	370	0
Calcium Sulfate	110	0	110	110	0
Formic Acid	1,800	0	1,800	1,800	0
Furfural	6	0	6	0.91	6
Lactic Acid	2,100	0	2,100	2,100	0
Levulinic Acid	1,800	0	1,800	1,800	0
Toluene	0	12	12	0	12
Water	150,000	0	150,000	150,000	0
Total	**150,000**	**12**	**150,000**	**150,000**	**18**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ Stream Identification

〉 Input/Output

■ Temperature (°C), gas

☐ Temperature (°C), liquid

▲ Pressure (kPa), gas

△ Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	
BFD 03-A-001	PFD 04-A-002
PFD 01-A-002	PFD 05-A-002
PFD 02-A-002	

Drawing Type

Process Flow Diagram

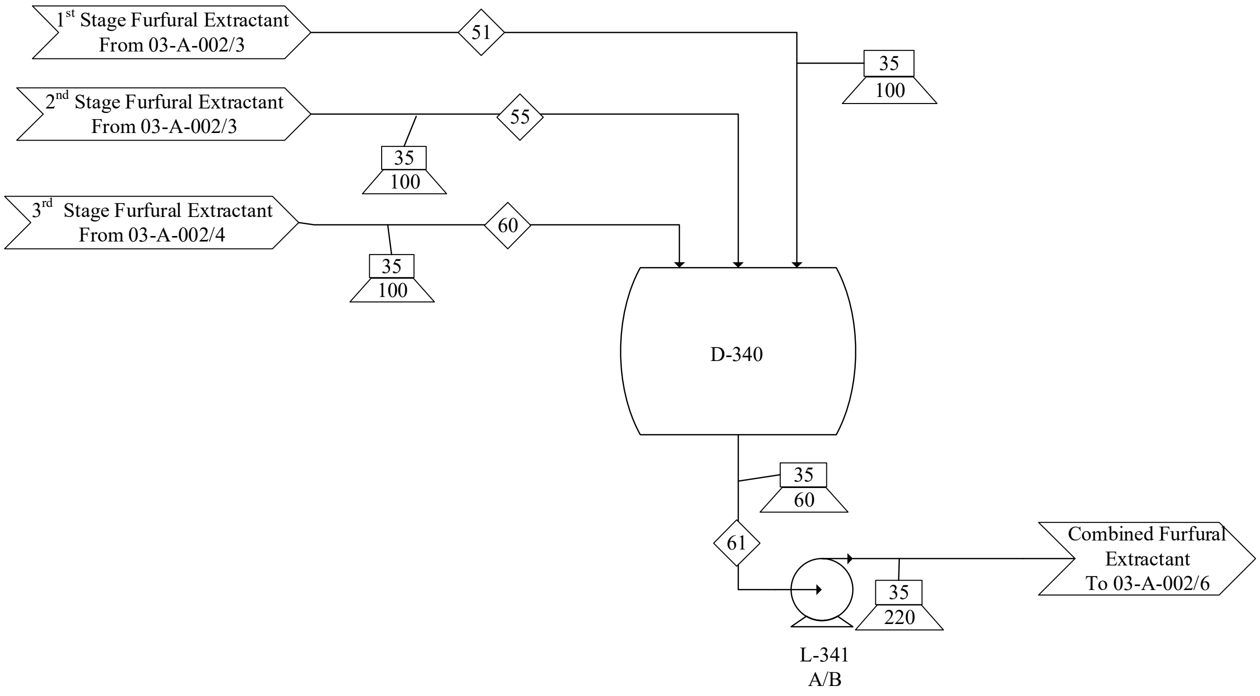
Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	03-A-002/4	0	390

D-340
Furfural Extractant
Holdup Drum
Height = 0.79 m
Diameter = 0.39 m

L-341 A/B
Furfural Extractant
Pump
Duty = 0.74 kW



Components/Stream	51	55	60	61
Furfural	160	25	6	190
Toluene	270	43	18	330
Total	**430**	**68**	**18**	**510**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

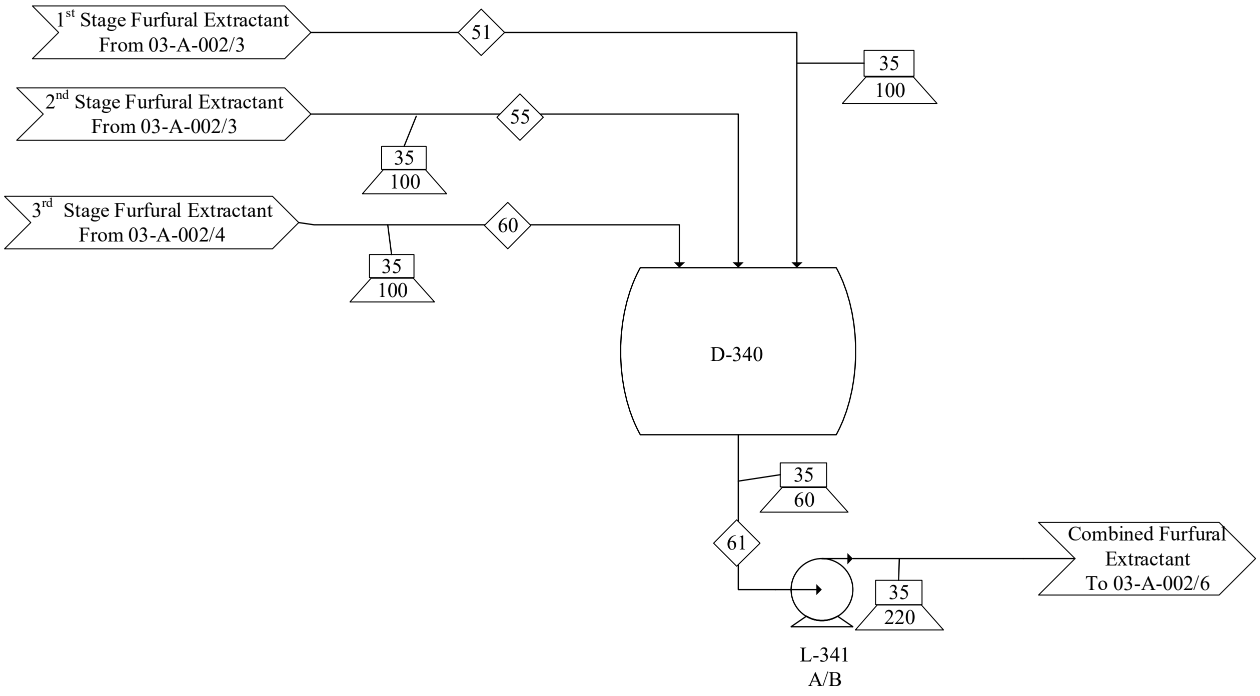
Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/5	0	391

D-340
Furfural Extractant
Holdup Drum
Height = 0.79 m
Diameter = 0.39 m

L-341 A/B
Furfural Extractant
Pump
Duty = 0.74 kW



Components/Stream	51	55	60	61
Furfural	160	25	6	190
Toluene	270	43	18	330
Total	**430**	**68**	**18**	**510**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ # Stream Identification
Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
▲ # Pressure (kPa), gas
△ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/5	0	392

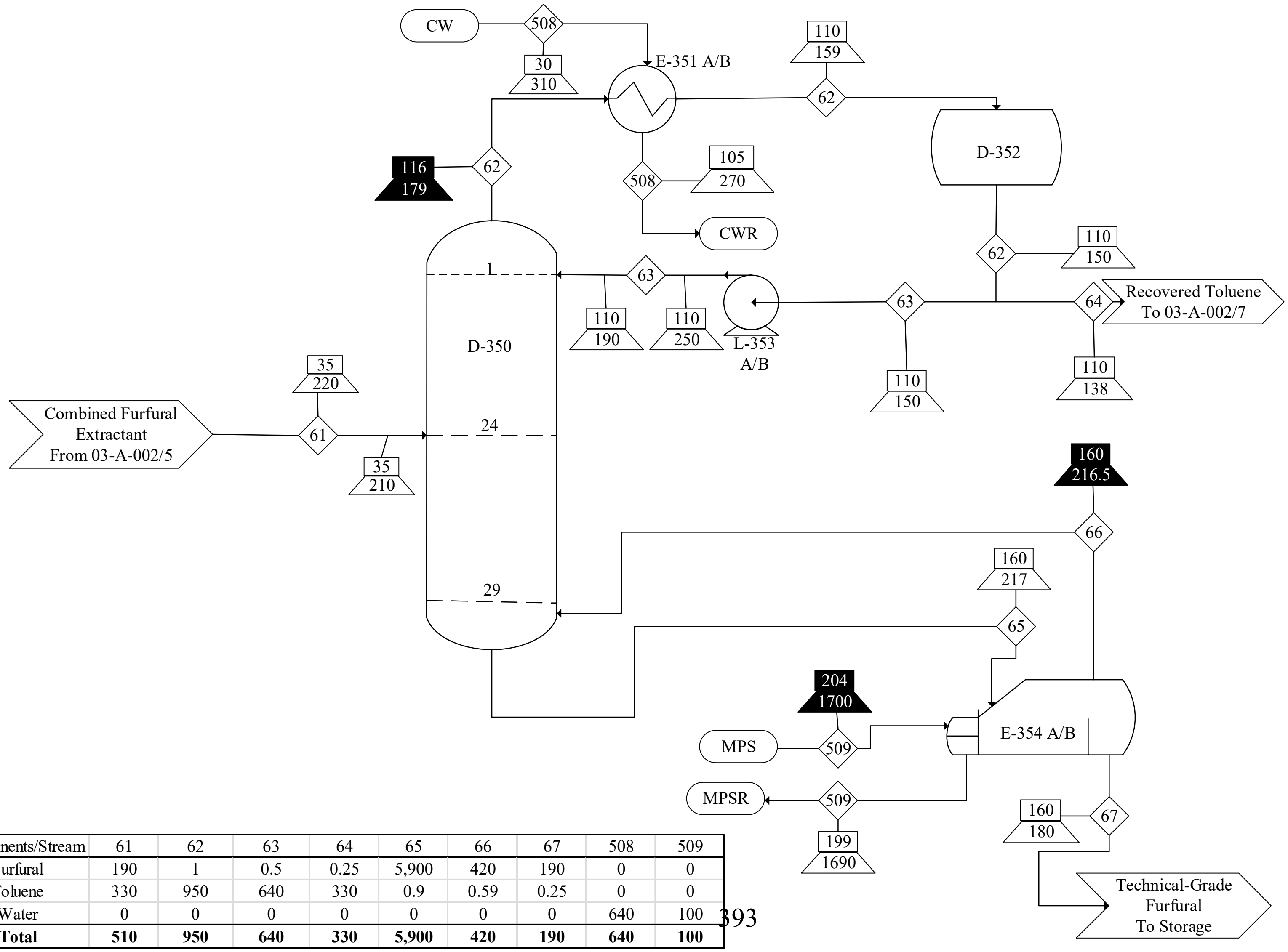
D-350
Toluene Recovery Column
Diameter: 0.23 m
Height: 12 m

E-351 A/B
Toluene Recovery Condenser
Duty: 0.20 GJ/hr
Area: 3.9 m^2

D-352
Toluene Recovery Reflux Drum
Height: 0.28 m
Length: 0.85 m

L-353 A/B
Toluene Recovery Reflux Pump
Duty: 0.74 kW

E-354 A/B
Toluene Recovery Reboiler
Duty: 0.30 GJ/hr
Area: 2.4 m^2



Components/Stream	61	62	63	64	65	66	67	508	509
Furfural	190	1	0.5	0.25	5,900	420	190	0	0
Toluene	330	950	640	330	0.9	0.59	0.25	0	0
Water	0	0	0	0	0	0	0	640	100
Total	**510**	**950**	**640**	**330**	**5,900**	**420**	**190**	**640**	**100**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Trays have and efficiency of 70%
6. Pressures are different in streams 91 and 92 due to control valve on stream 91 being after the pump.

Special Symbols / Nomenclature

#	Stream Identification
#	Input/Output
#	Temperature (°C), gas
#	Temperature (°C), liquid
#	Pressure (kPa), gas
#	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 02-A-002

Drawing Type

Process Flow Diagram

Title

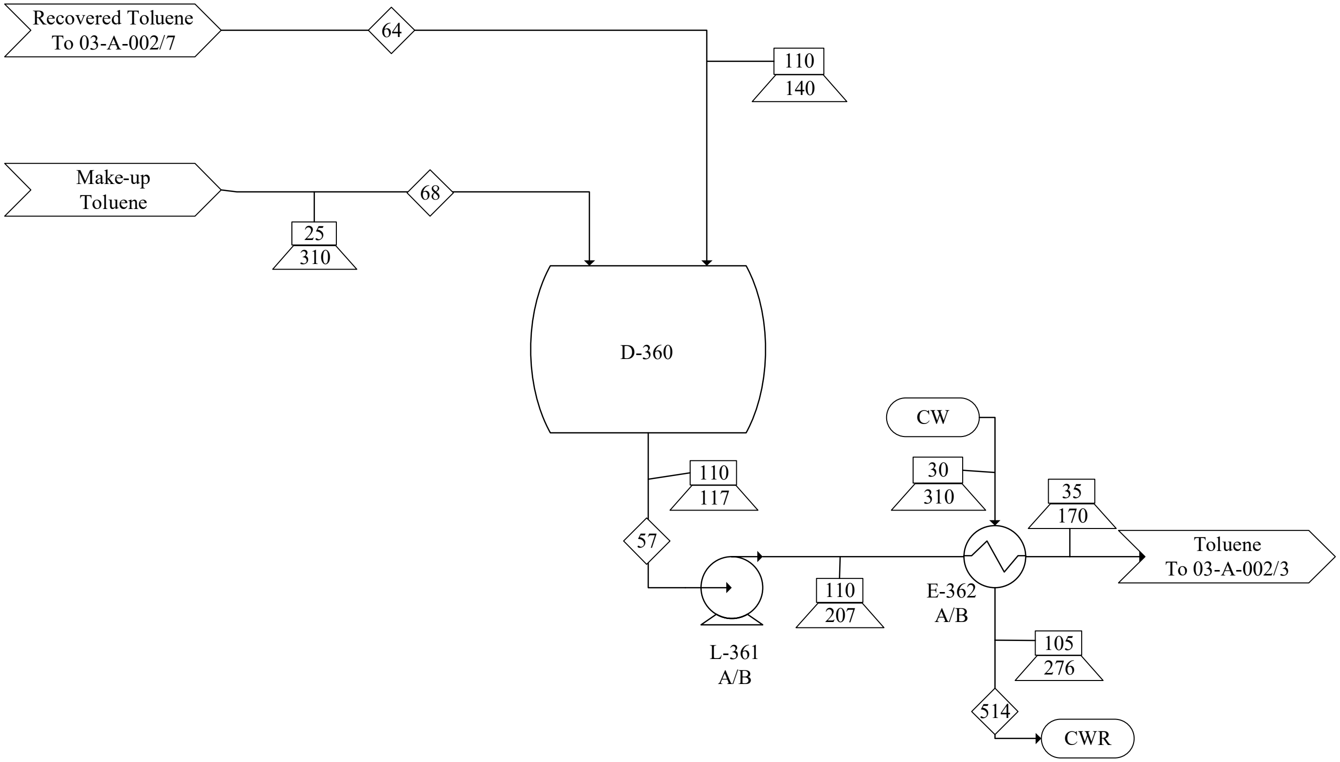
Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/6	0	393

D-360
Toluene Recycle Drum
Height = 0.94 m
Diameter = 0.30 m

L-361 A/B
Toluene Pump
Duty = 0.74 kW

E-362 A/B
Toluene Cooler
Duty = 0.044 MJ/hr
Area = 9.3 m^2



Components/Stream	64	68	57	510
Furfural	0.25	0	0.25	0
Toluene	330	Trace	330	0
Water	0	0	0	140
Total	**330**	**Trace**	**330**	**140**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Make-up Toluene flow is normally zero.

Special Symbols / Nomenclature

◇ # Stream Identification
Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
Pressure (kPa), gas
Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 02-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/7	0	394

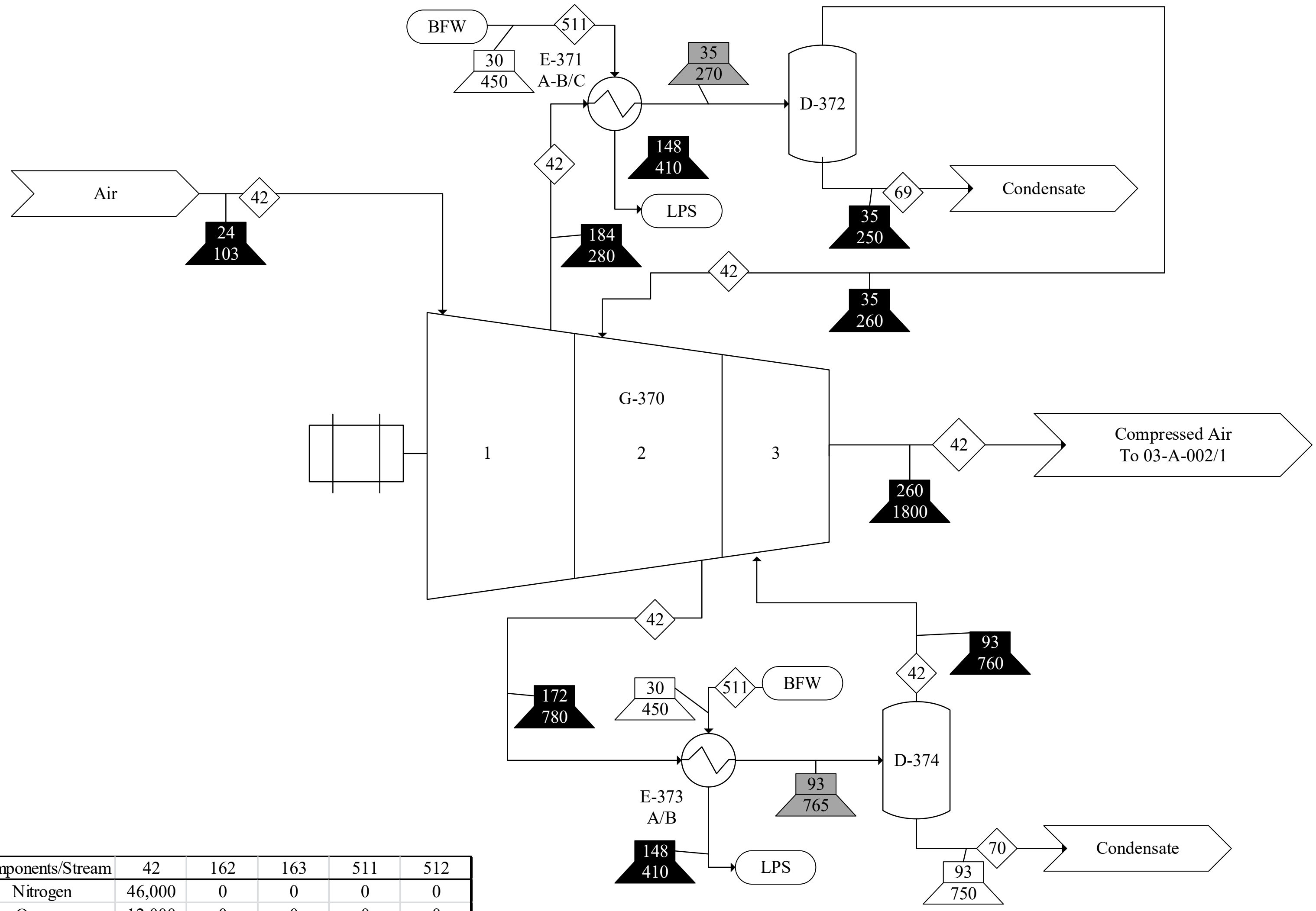
G-370
Three Stage Air Compressor
1st Stage Duty = 2,300 kW
2nd Stage Duty = 2,300 kW
3rd Stage Duty = 2,300 kW

E-371 A-B/C
First Inter-stage cooler
Duty = 8.2 GJ/hr
Area = 470 m^2

D-372
First Inter-stage Knock-out Drum
Diameter = 3.6 m
Height = 7.3 m

E-373 A/B
Second Inter-stage cooler
Duty = 4.9 GJ/hr
Area = 79 ft^2

D-374
Second Inter-stage Knock-out Drum
Diameter = 2.8 m
Height =5.5 m



Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Flows of stream 162 and 163 are normally zero.
5. Air enters at an average humidity of 70%.

Special Symbols / Nomenclature

◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
□ #	Temperature (°C), liquid
▨ #	Temperature (°C), mix phase
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid
▲ #	Pressure (kPa), mix phase

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

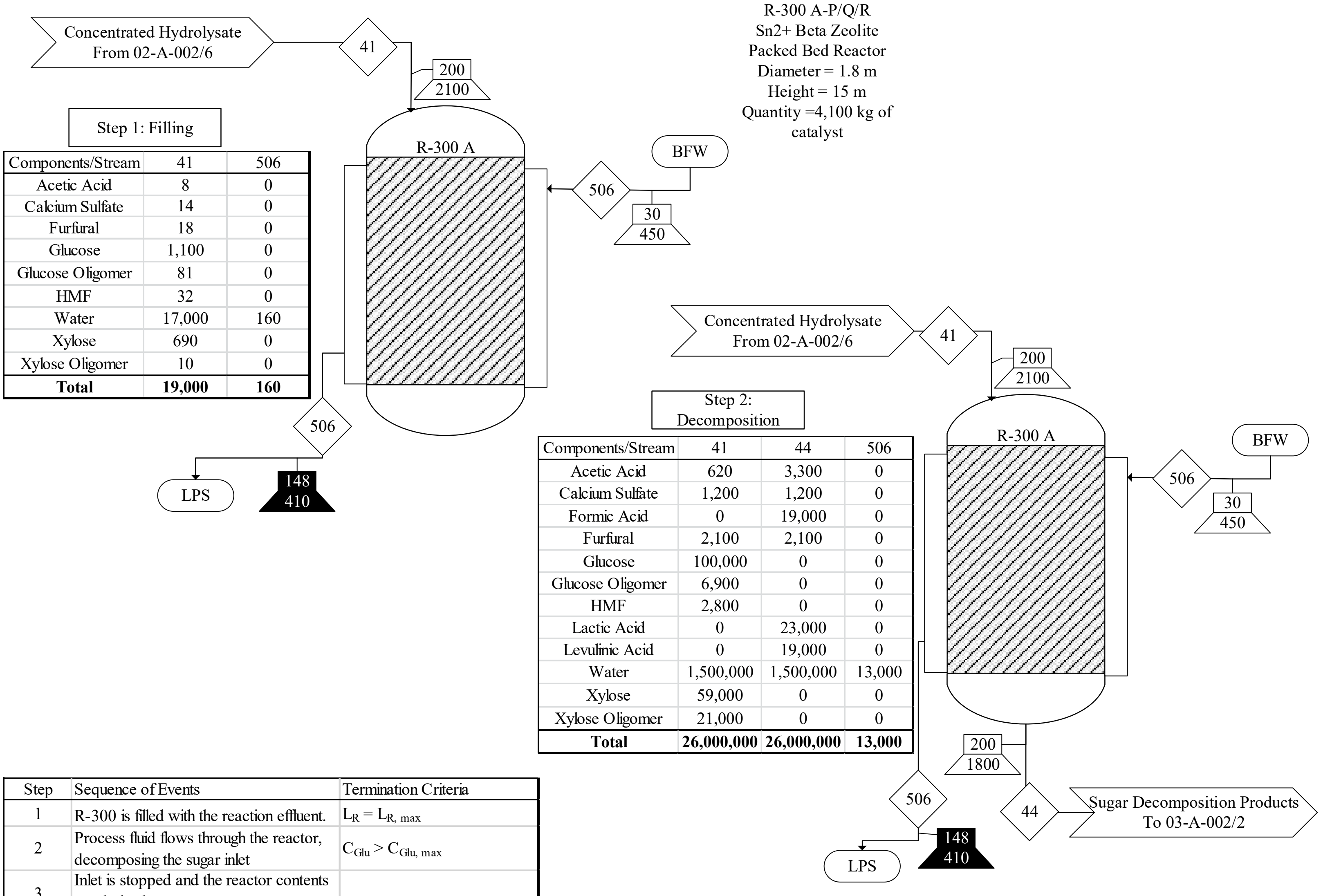
Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/8	0	395

Components/Stream	42	162	163	511	512
Nitrogen	46,000	0	0	0	0
Oxygen	12,000	0	0	0	0
Water	560	Trace	Trace	3,100	1,800
Total	**82,000**	**Trace**	**Trace**	**3,100**	**1,800**



Step 1: Filling

Components/Stream	41	506
Acetic Acid	8	0
Calcium Sulfate	14	0
Furfural	18	0
Glucose	1,100	0
Glucose Oligomer	81	0
HMF	32	0
Water	17,000	160
Xylose	690	0
Xylose Oligomer	10	0
Total	**19,000**	**160**

Step 2: Decomposition

Components/Stream	41	44	506
Acetic Acid	620	3,300	0
Calcium Sulfate	1,200	1,200	0
Formic Acid	0	19,000	0
Furfural	2,100	2,100	0
Glucose	100,000	0	0
Glucose Oligomer	6,900	0	0
HMF	2,800	0	0
Lactic Acid	0	23,000	0
Levulinic Acid	0	19,000	0
Water	1,500,000	1,500,000	13,000
Xylose	59,000	0	0
Xylose Oligomer	21,000	0	0
Total	**26,000,000**	**26,000,000**	**13,000**

Step	Sequence of Events	Termination Criteria
1	R-300 is filled with the reaction effluent.	$L_R = L_{R,\ max}$
2	Process fluid flows through the reactor, decomposing the sugar inlet	$C_{Glu} > C_{Glu,\ max}$
3	Inlet is stopped and the reactor contents are drained.	$L_R < L_{R,\ min}$
4	Compressed air is run through the catalyst, removing the coke (in the form of Carbon Dioxide) produced by decomposition of the sugars.	$C_{CO2} < C_{CO2,\ min}$
5	R-300 is on standby	$C_{Glu} > C_{Glu,\ max}$ of previous reactor

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 2 hours
5. Step 2 is 172 hours
6. Eighteen reactors are used in parallel for continual decomposition.
7. At any time, sixteen reactors are is use and two are being calcined.

Special Symbols / Nomenclature

◇#	Stream Identification
〉	Input/Output
■#	Temperature (°C), gas
☐#	Temperature (°C), liquid
▲#	Pressure (kPa), gas
△#	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

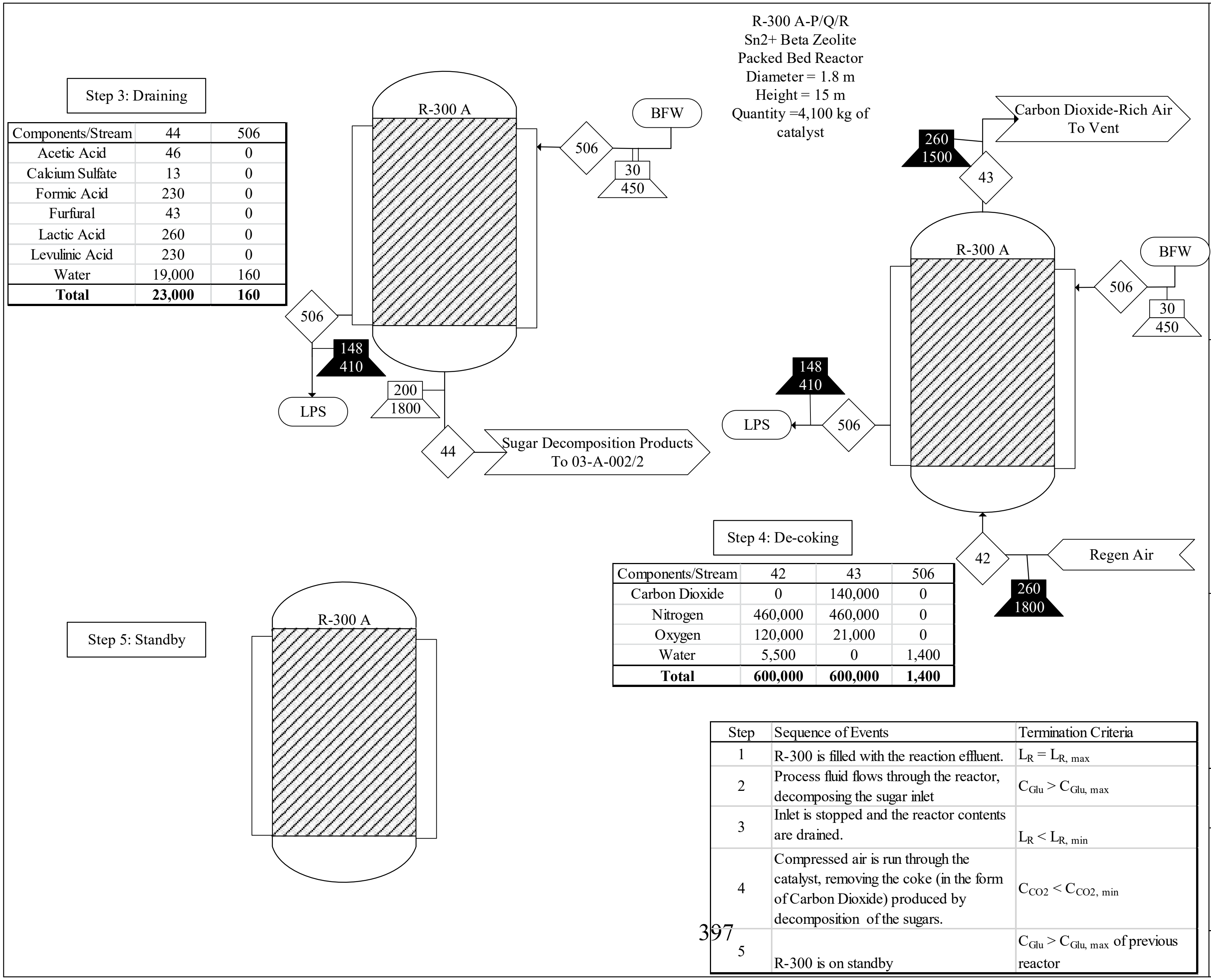
Drawing Type

Supplemental Process Flow Diagram

Title

Catalytic Reactor Semi-Batch Steps

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	03-A-003/1	0	396



Step 3: Draining

Components/Stream	44	506
Acetic Acid	46	0
Calcium Sulfate	13	0
Formic Acid	230	0
Furfural	43	0
Lactic Acid	260	0
Levulinic Acid	230	0
Water	19,000	160
Total	**23,000**	**160**

Step 4: De-coking

Components/Stream	42	43	506
Carbon Dioxide	0	140,000	0
Nitrogen	460,000	460,000	0
Oxygen	120,000	21,000	0
Water	5,500	0	1,400
Total	**600,000**	**600,000**	**1,400**

Step	Sequence of Events	Termination Criteria
1	R-300 is filled with the reaction effluent.	$L_R = L_{R,\ max}$
2	Process fluid flows through the reactor, decomposing the sugar inlet	$C_{Glu} > C_{Glu,\ max}$
3	Inlet is stopped and the reactor contents are drained.	$L_R < L_{R,\ min}$
4	Compressed air is run through the catalyst, removing the coke (in the form of Carbon Dioxide) produced by decomposition of the sugars.	$C_{CO2} < C_{CO2,\ min}$
5	R-300 is on standby	$C_{Glu} > C_{Glu,\ max}$ of previous reactor

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 3 is 2 hours
5. Step 4 is 20 hours
6. Step 5 is 4 hour
7. Eighteen reactors are used in parallel for continual decomposition.
8. At any time, sixteen reactors are is use and two are being calcined.

Special Symbols / Nomenclature

Symbol	Meaning
◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (psia), gas
△ #	Pressure (psia), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

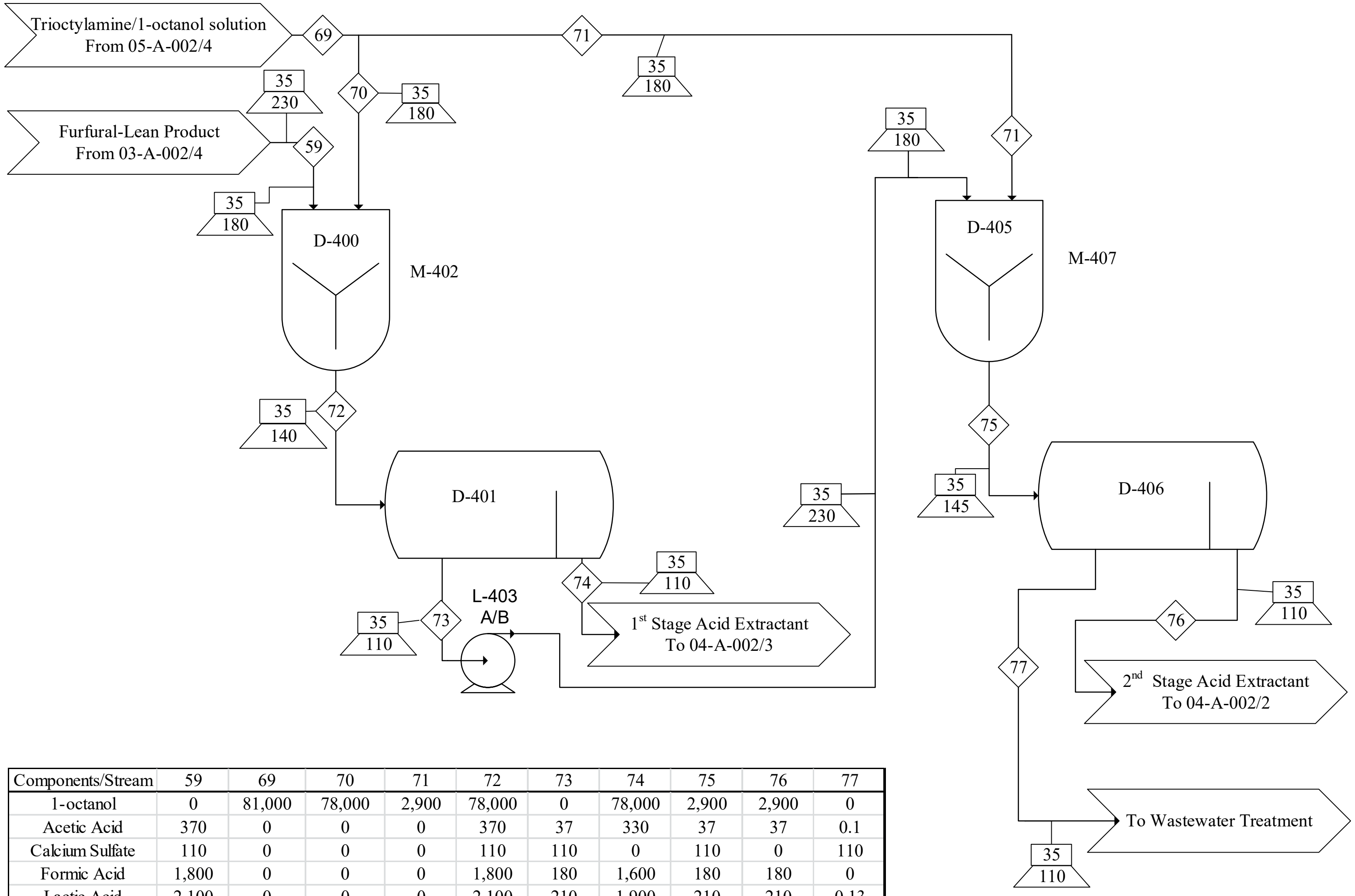
Supplemental Process Flow Diagram

Title

Catalytic Reactor Semi-Batch Steps

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	03-A-003/2	0	397

D-400	D-401	M-402	L-403 A/B	D-405	D-406	M-407
1st Stage TOA Mixing Drum	1st Stage TOA Settler	1st Stage TOA Mixer	1st Stage TOA Pump	2nd Stage TOA Mixing Drum	2nd Stage TOA Settler	2nd Stage TOA Mixer
Height = 8.3 m Diameter = 4.1 m	Height = 4.6 m Diameter = 2.3 m	Power = 54 kW	Duty = 8.3 kW	Height = 8.3 m Diameter = 4.1 m	Height = 4.6 m Diameter = 2.3 m	Power = 8.3 kW



Components/Stream	59	69	70	71	72	73	74	75	76	77
1-octanol	0	81,000	78,000	2,900	78,000	0	78,000	2,900	2,900	0
Acetic Acid	370	0	0	0	370	37	330	37	37	0.1
Calcium Sulfate	110	0	0	0	110	110	0	110	0	110
Formic Acid	1,800	0	0	0	1,800	180	1,600	180	180	0
Lactic Acid	2,100	0	0	0	2,100	210	1,900	210	210	0.13
Levulinic Acid	1,800	0	0	0	1,800	180	1,600	180	180	1.2
Trioctylamine	0	25,000	24,000	860	24,000	0	24,000	860	860	0
Water	150,000	0	0	0	150,000	150,000	0	150,000	0	150,000
Total	**150,000**	**100,000**	**100,000**	**3,800**	**250,000**	**150,000**	**110,000**	**150,000**	**4,400**	**150,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #

〉

■ #

□ #

▲ #

△ #

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

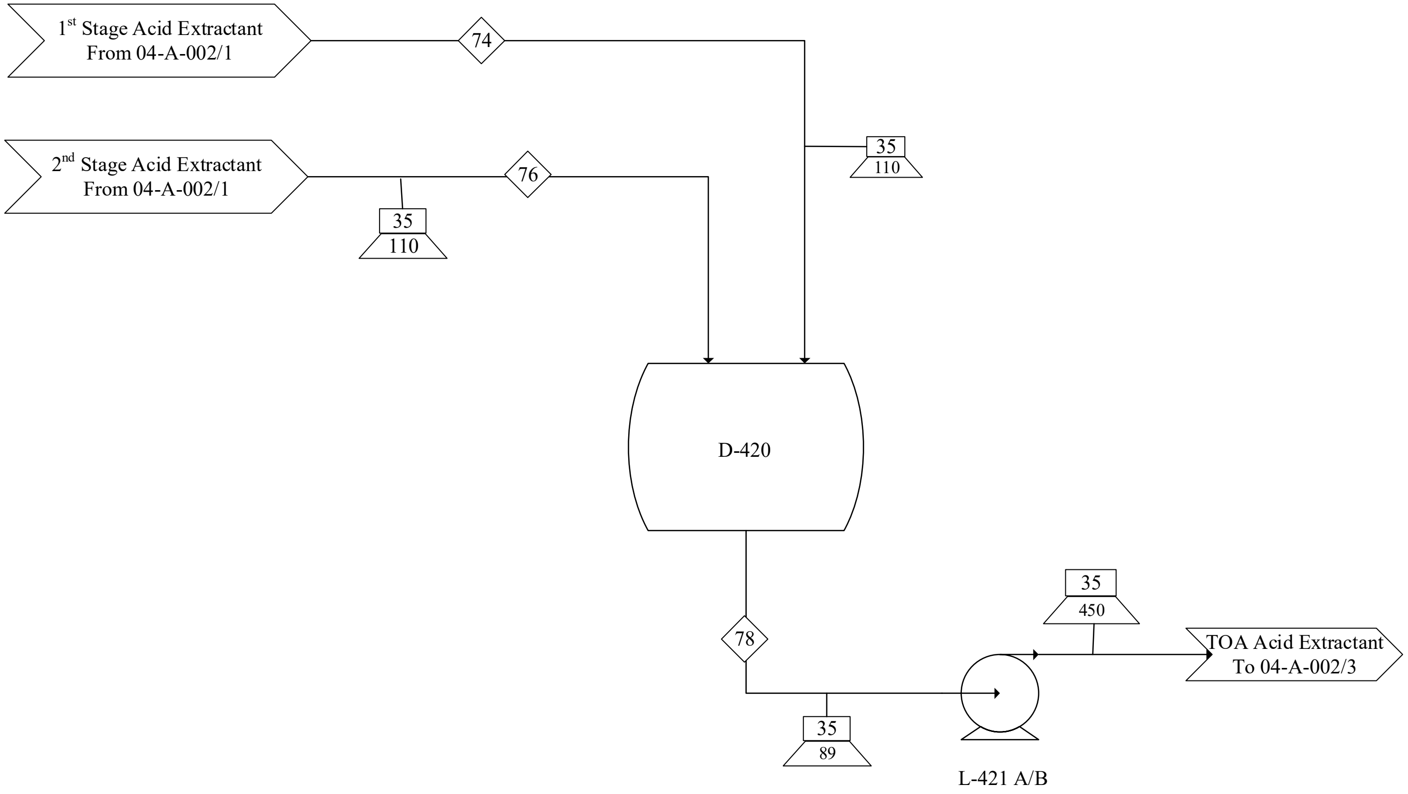
Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/1	0	398

D-420
TOA Acid Extractant
Hold Up Drum
Height = 4.4 m
Diameter = 2.2 m

L-421 A/B
TOA Hold Up Drum
Pump
Duty = 13 kW



Components/Stream	74	76	78
1-octanol	78,000	2,900	81,000
Acetic Acid	330	37	370
Formic Acid	1,600	180	1,800
Lactic Acid	1,900	210	2,100
Levulinic Acid	1,600	180	1,800
Trioctylamine	24,000	860	25,000
Total	**110,000**	**4,400**	**120,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ # Stream Identification
Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
Pressure (kPa), gas
Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001 PFD 03-A-002
PFD 01-A-002 PFD 05-A-002
PFD 02-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/2	0	399

D-430
Volatile Acid Recovery Column
Rectifying Diameter: 0.58 m
Stripping Diameter: 3.9 m
Height: 13 m

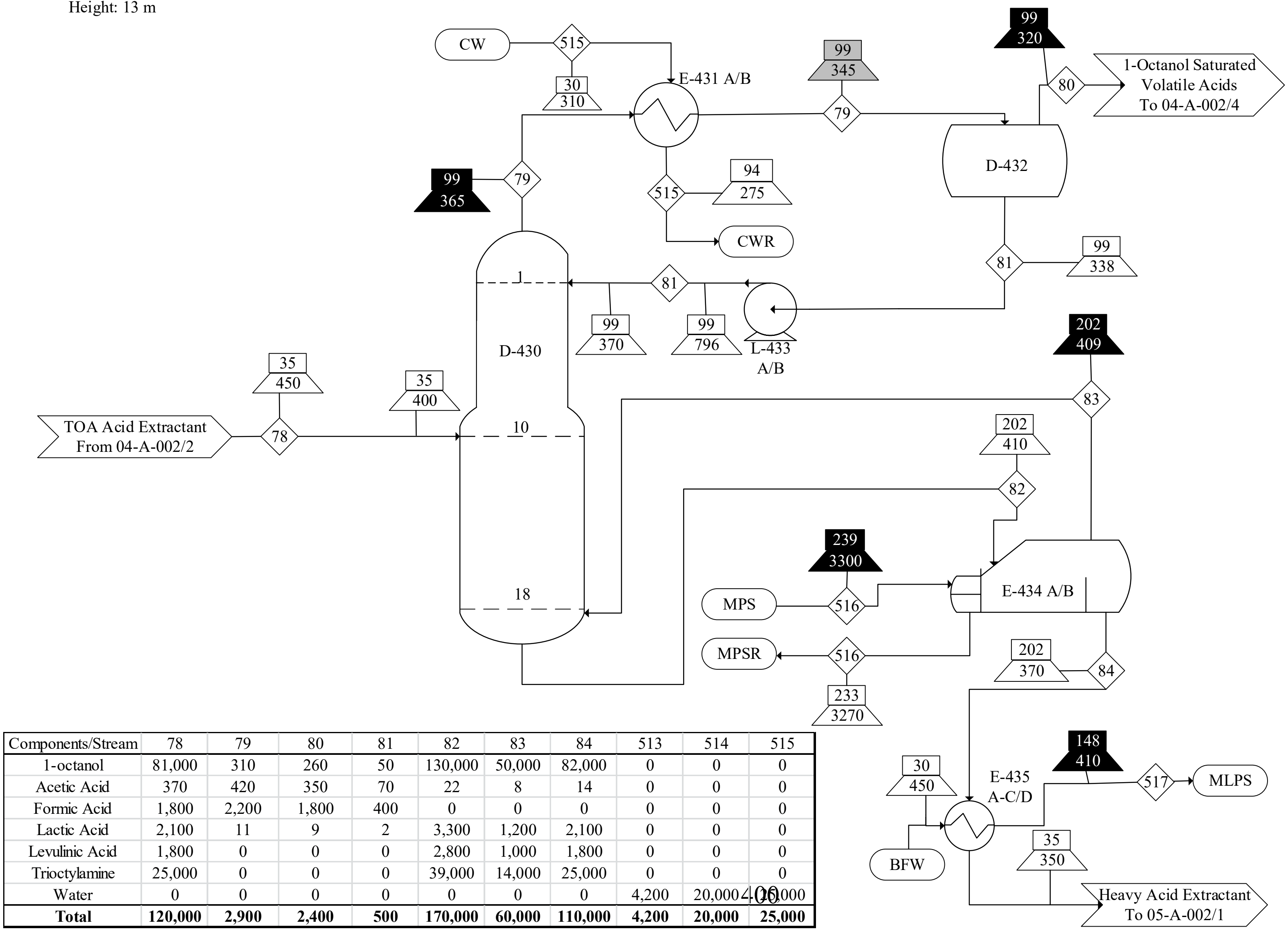
E-431 A/B
Volatile Acid Column Condenser
Duty: 1.5 GJ/hr
Area: 19 m^2

D-432
Volatile Acid Column Reflux Drum
Height: 1.5 m
Length: 0.73 m

L-433 A/B
Volatile Acid Column Reflux Pump
Duty: 0.74 kW

E-434 A/B
Volatile Acid Column Reboiler
Duty: 56 GJ/hr
Area: 590 m^2

E-435 A-C/D
Heavy Acid Extractant Cooler
Duty: 65 GJ/hr
Area: 740 m^2



Components/Stream	78	79	80	81	82	83	84	513	514	515
1-octanol	81,000	310	260	50	130,000	50,000	82,000	0	0	0
Acetic Acid	370	420	350	70	22	8	14	0	0	0
Formic Acid	1,800	2,200	1,800	400	0	0	0	0	0	0
Lactic Acid	2,100	11	9	2	3,300	1,200	2,100	0	0	0
Levulinic Acid	1,800	0	0	0	2,800	1,000	1,800	0	0	0
Trioctylamine	25,000	0	0	0	39,000	14,000	25,000	0	0	0
Water	0	0	0	0	0	0	0	4,200	20,000	25,000
Total	**120,000**	**2,900**	**2,400**	**500**	**170,000**	**60,000**	**110,000**	**4,200**	**20,000**	**25,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Trays have and efficiency of 70%

Special Symbols / Nomenclature

◇# Stream Identification
▷ Input/Output
■# Temperature (°F), gas
☐# Temperature (°F), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG # / Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/3	0	400

D-430
Volatile Acid Recovery Column
Rectifying Diameter: 0.58 m
Stripping Diameter: 3.9 m
Height: 13 m

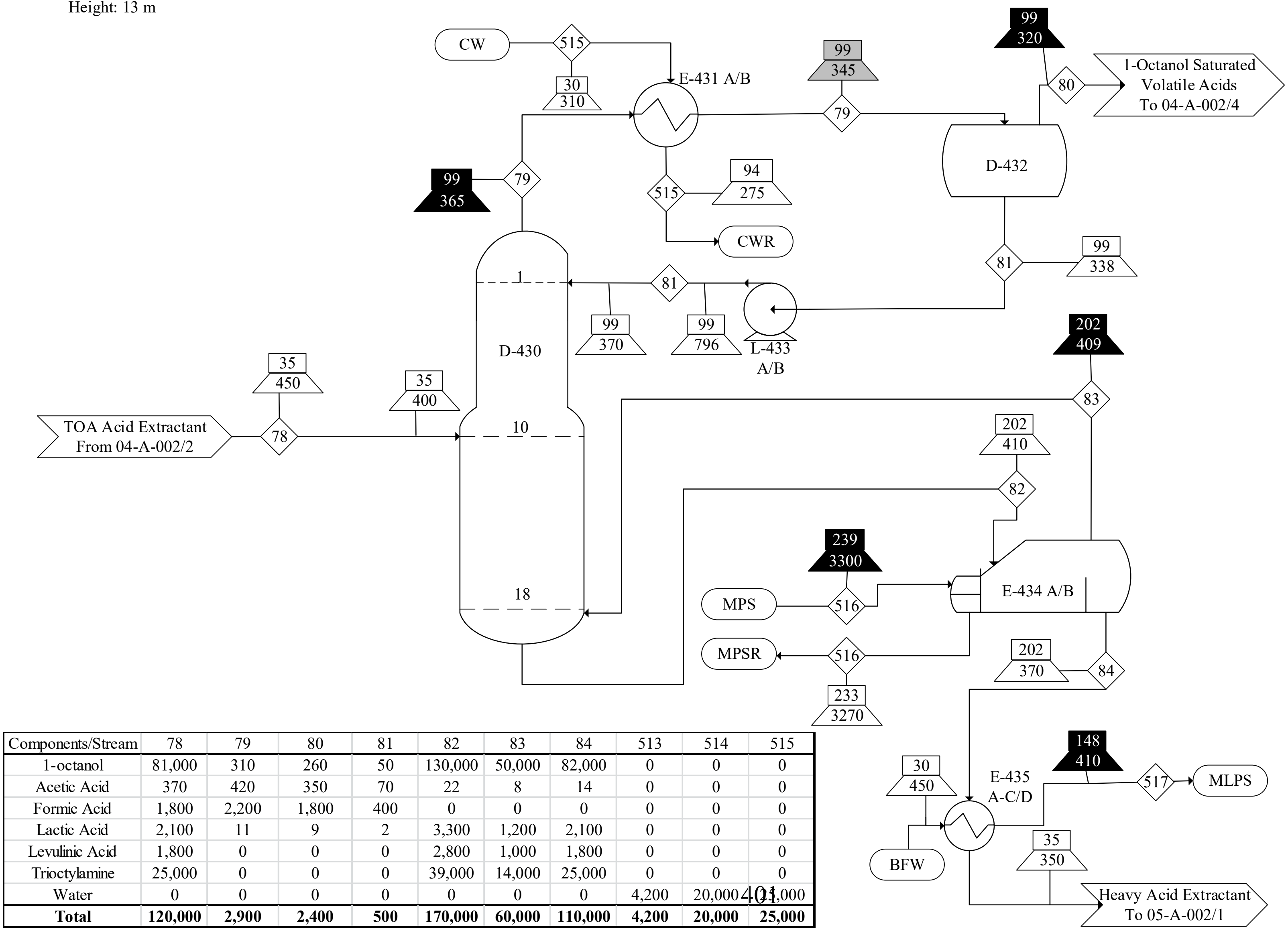
E-431 A/B
Volatile Acid Column Condenser
Duty: 1.5 GJ/hr
Area: 19 m^2

D-432
Volatile Acid Column Reflux Drum
Height: 1.5 m
Length: 0.73 m

L-433 A/B
Volatile Acid Column Reflux Pump
Duty: 0.74 kW

E-434 A/B
Volatile Acid Column Reboiler
Duty: 56 GJ/hr
Area: 590 m^2

E-435 A-C/D
Heavy Acid Extractant Cooler
Duty: 65 GJ/hr
Area: 740 m^2



Components/Stream	78	79	80	81	82	83	84	513	514	515
1-octanol	81,000	310	260	50	130,000	50,000	82,000	0	0	0
Acetic Acid	370	420	350	70	22	8	14	0	0	0
Formic Acid	1,800	2,200	1,800	400	0	0	0	0	0	0
Lactic Acid	2,100	11	9	2	3,300	1,200	2,100	0	0	0
Levulinic Acid	1,800	0	0	0	2,800	1,000	1,800	0	0	0
Trioctylamine	25,000	0	0	0	39,000	14,000	25,000	0	0	0
Water	0	0	0	0	0	0	0	4,200	20,000	25,000
Total	**120,000**	**2,900**	**2,400**	**500**	**170,000**	**60,000**	**110,000**	**4,200**	**20,000**	**25,000**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Pumps have an efficiency of 65%

5. Trays have and efficiency of 70%

Special Symbols / Nomenclature

◇#	Stream Identification
▷	Input/Output
■#	Temperature (°F), gas
□#	Temperature (°F), liquid
▲#	Pressure (kPa), gas
△#	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

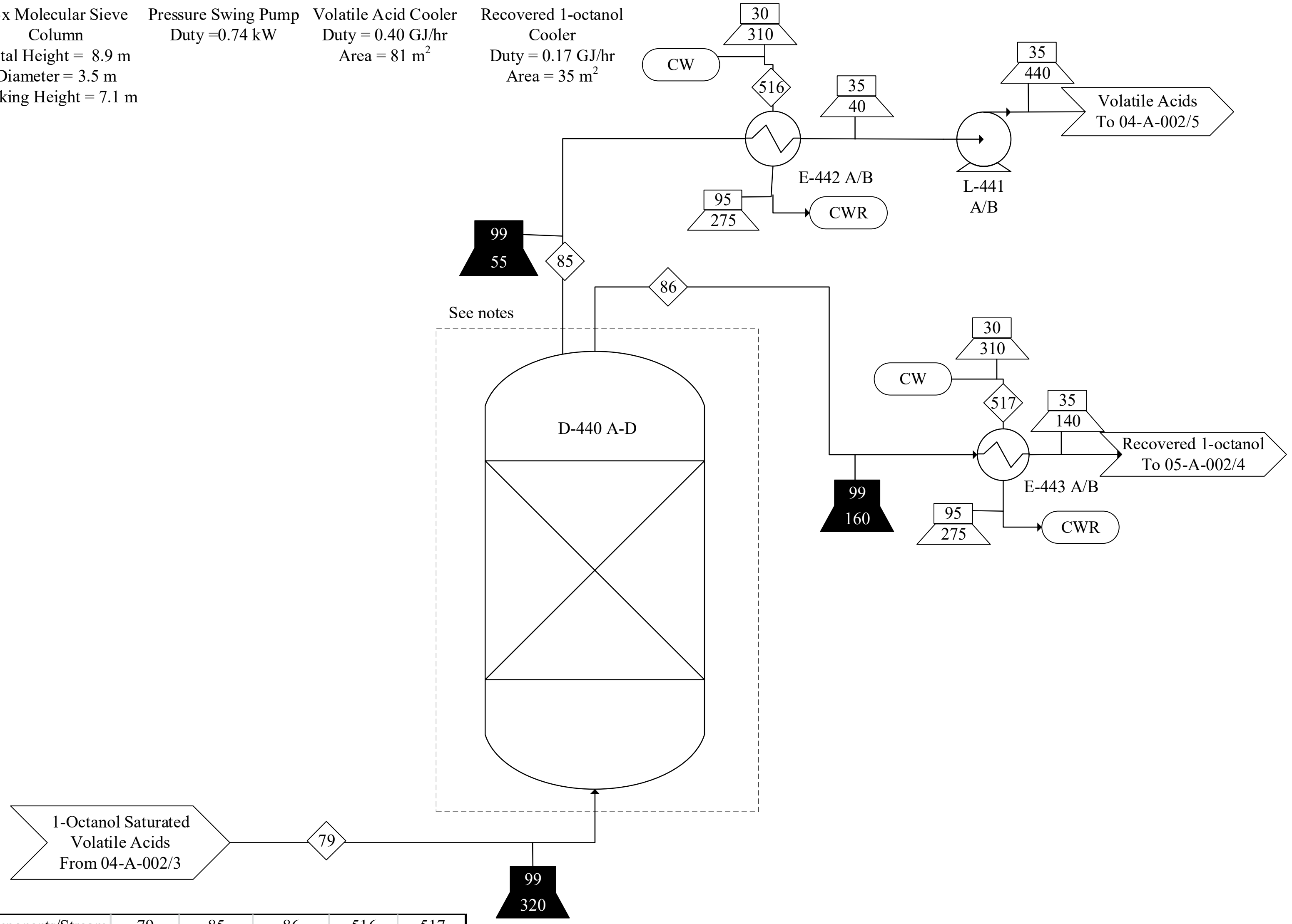
By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/3	0	401

D-440 A-D
13x Molecular Sieve Column
Total Height = 8.9 m
Diameter = 3.5 m
Packing Height = 7.1 m

L-441
Pressure Swing Pump
Duty =0.74 kW

E-442 A/B
Volatile Acid Cooler
Duty = 0.40 GJ/hr
Area = 81 m^2

E-443 A/B
Recovered 1-octanol Cooler
Duty = 0.17 GJ/hr
Area = 35 m^2



Components/Stream	79	85	86	516	517
1-octanol	260	0	260	0	0
Acetic Acid	350	350	0	0	0
Formic Acid	1,800	1,800	0	0	0
Water	0	0	0	1,100	640
Total	**2,400**	**2,100**	**260**	**1,100**	**640**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Compressors have an efficiency of 65%
5. D-440 is a semi-batch process, more information can be seen on sheet 04-A-003

Special Symbols / Nomenclature

◇# Stream Identification
▷ Input/Output
■# Temperature (°F), gas
☐# Temperature (°F), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 03-A-002
BFD 04-A-001	PFD 05-A-002
PFD 01-A-002	Supplemental
PFD 02-A-002	PFD 04-A-003

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/4	0	402

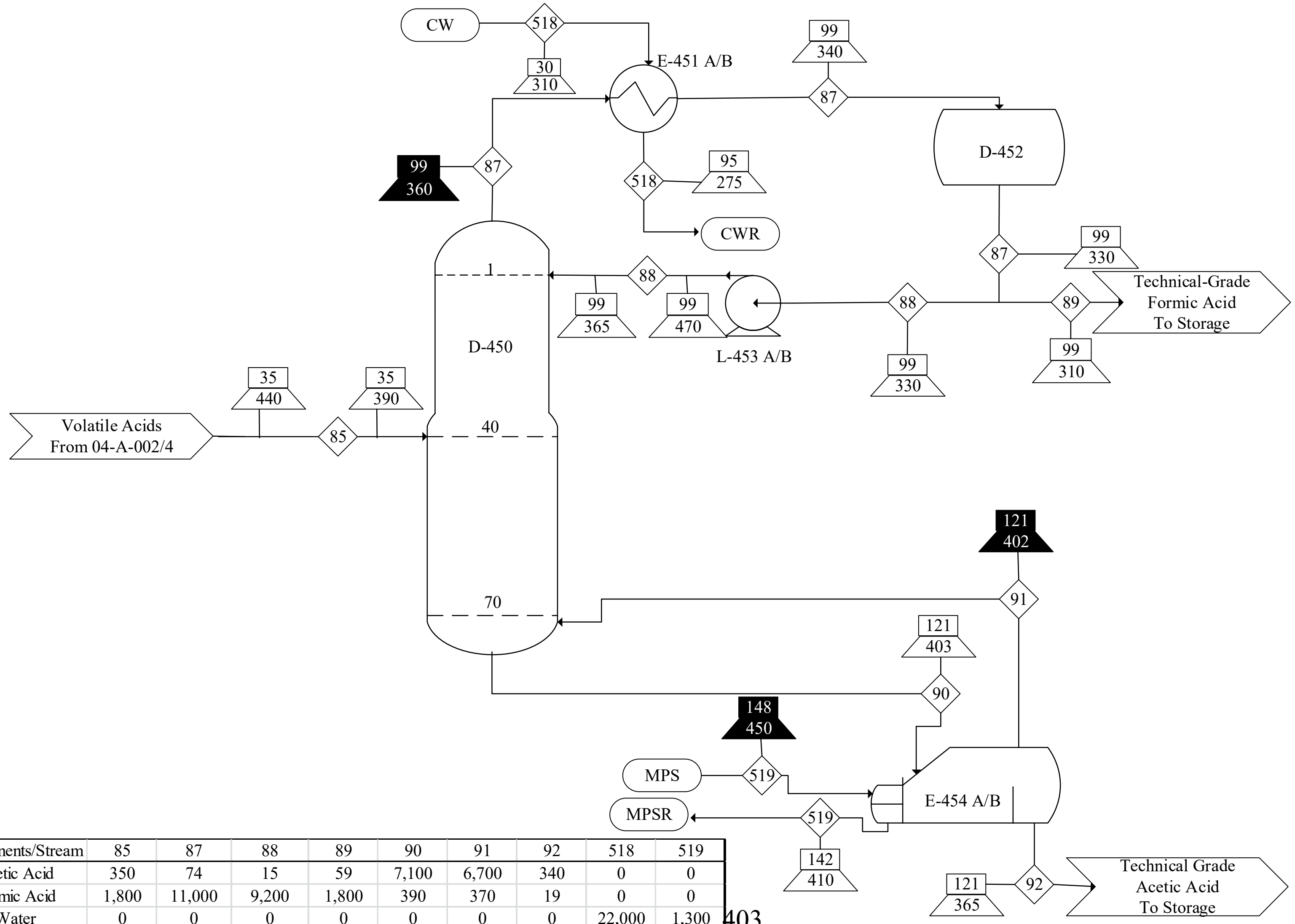
D-450
Formic/Acetic Column
Rectifying Diameter: 1.3 m
Stripping Diameter: 0.85 m
Height: 43 m

E-451
Formic/Acetic Column
Condenser
Duty: 6.0 GJ/hr
Area: 91 m^2

D-452
Formic/Acetic Column
Reflux Drum
Diameter: 1.1 m
Length: 2.2 m

L-453
Formic/Acetic Column
Reflux Pump
Duty: 0.74 kW

E-454
Formic/Acetic Column
Reboiler
Duty: 3.5 GJ/hr
Area: 27 m^2



Components/Stream	85	87	88	89	90	91	92	518	519
Acetic Acid	350	74	15	59	7,100	6,700	340	0	0
Formic Acid	1,800	11,000	9,200	1,800	390	370	19	0	0
Water	0	0	0	0	0	0	0	22,000	1,300
Total	**2,100**	**11,000**	**9,200**	**1,800**	**7,500**	**7,100**	**360**	**22,000**	**1,300**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Trays have and efficiency of 70%

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

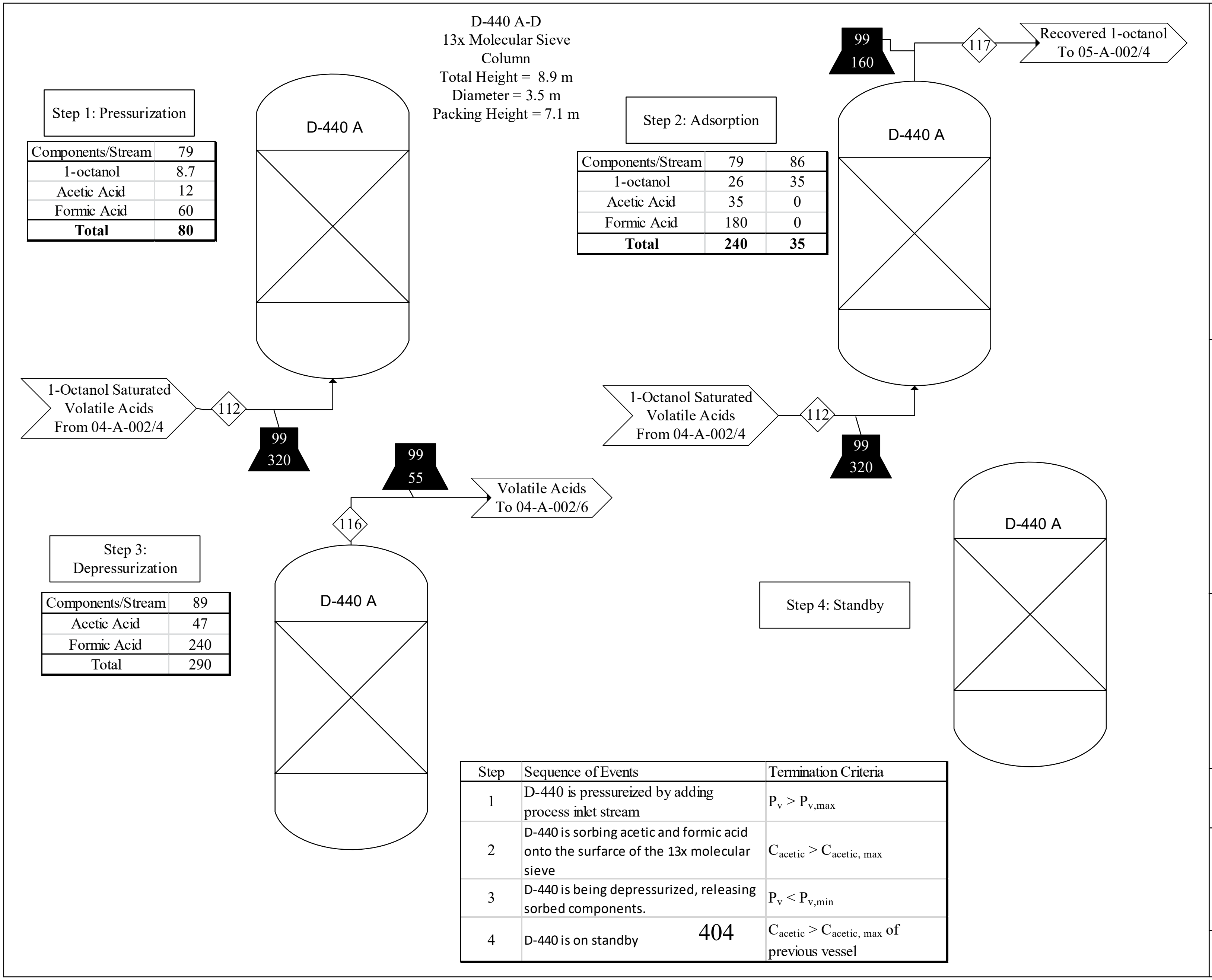
Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/5	0	403



Step 1: Pressurization

Components/Stream	79
1-octanol	8.7
Acetic Acid	12
Formic Acid	60
Total	**80**

Step 2: Adsorption

Components/Stream	79	86
1-octanol	26	35
Acetic Acid	35	0
Formic Acid	180	0
Total	**240**	**35**

Step 3:
Depressurization

Components/Stream	89
Acetic Acid	47
Formic Acid	240
Total	290

Step	Sequence of Events	Termination Criteria
1	D-440 is pressureized by adding process inlet stream	$P_v > P_{v,max}$
2	D-440 is sorbing acetic and formic acid onto the surface of the 13x molecular sieve	$C_{acetic} > C_{acetic,\ max}$
3	D-440 is being depressurized, releasing sorbed components.	$P_v < P_{v,min}$
4	D-440 is on standby	$C_{acetic} > C_{acetic,\ max}$ of previous vessel

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 2 minutes
5. Step 2 is 6 minutes
6. Step 3 is 24 minutes
7. Step 4 is 4 minute
8. Four adsorbers used in parallel for continual adsorption

Special Symbols / Nomenclature

◇#	Stream Identification
〉	Input/Output
■#	Temperature (°C), gas
□#	Temperature (°C), liquid
▲#	Pressure (kPa), gas
△#	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

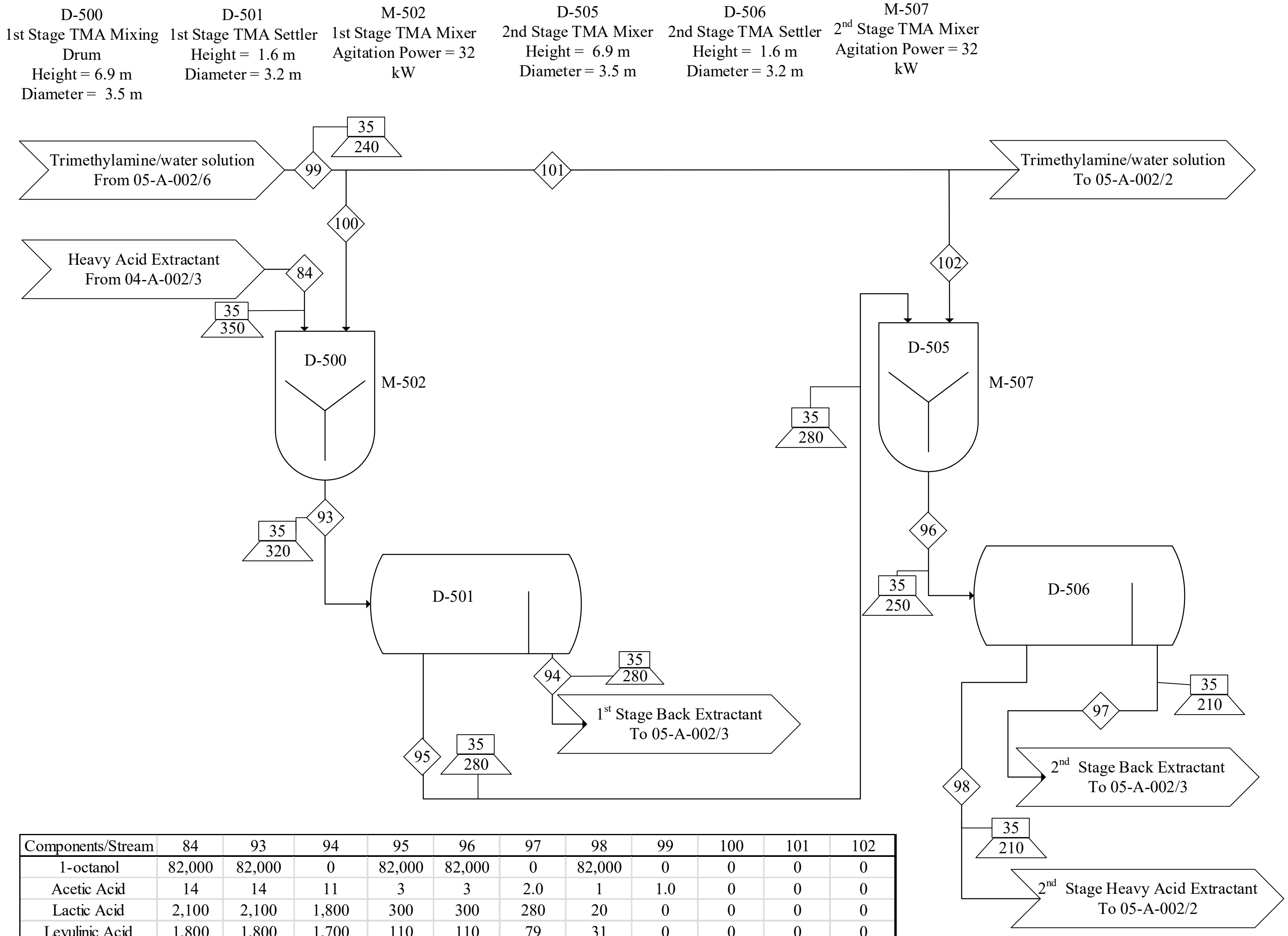
Drawing Type

Supplemental Process Flow Diagram

Title

13x Molecular Sieve Column Steps

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	04-A-003/1	0	404



Components/Stream	84	93	94	95	96	97	98	99	100	101	102
1-octanol	82,000	82,000	0	82,000	82,000	0	82,000	0	0	0	0
Acetic Acid	14	14	11	3	3	2.0	1	1.0	0	0	0
Lactic Acid	2,100	2,100	1,800	300	300	280	20	0	0	0	0
Levulinic Acid	1,800	1,800	1,700	110	110	79	31	0	0	0	0
Trimethylamine	0	4,300	4,300	0	420	420	0	4,800	4,300	500	420
Trioctylamine	25,000	25,000	0	25,000	25,000	0	25,000	0	0	0	0
Water	0	4,200	4,200	0	340	340	0	4,600	4,200	400	340
Total	**110,000**	**120,000**	**13,000**	**100,000**	**100,000**	**1,100**	**100,000**	**9,400**	**8,500**	**820**	**880**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/1	0	405

D-510
3rd Stage TMA Mixing Drum
Height = 6.9 m
Diameter = 3.5 m

D-511
3rd Stage TMA Settler
Height = 1.6 m
Diameter = 3.2 m

M-512
3rd Stage TMA Mixer
Agitation Power = 32 kW

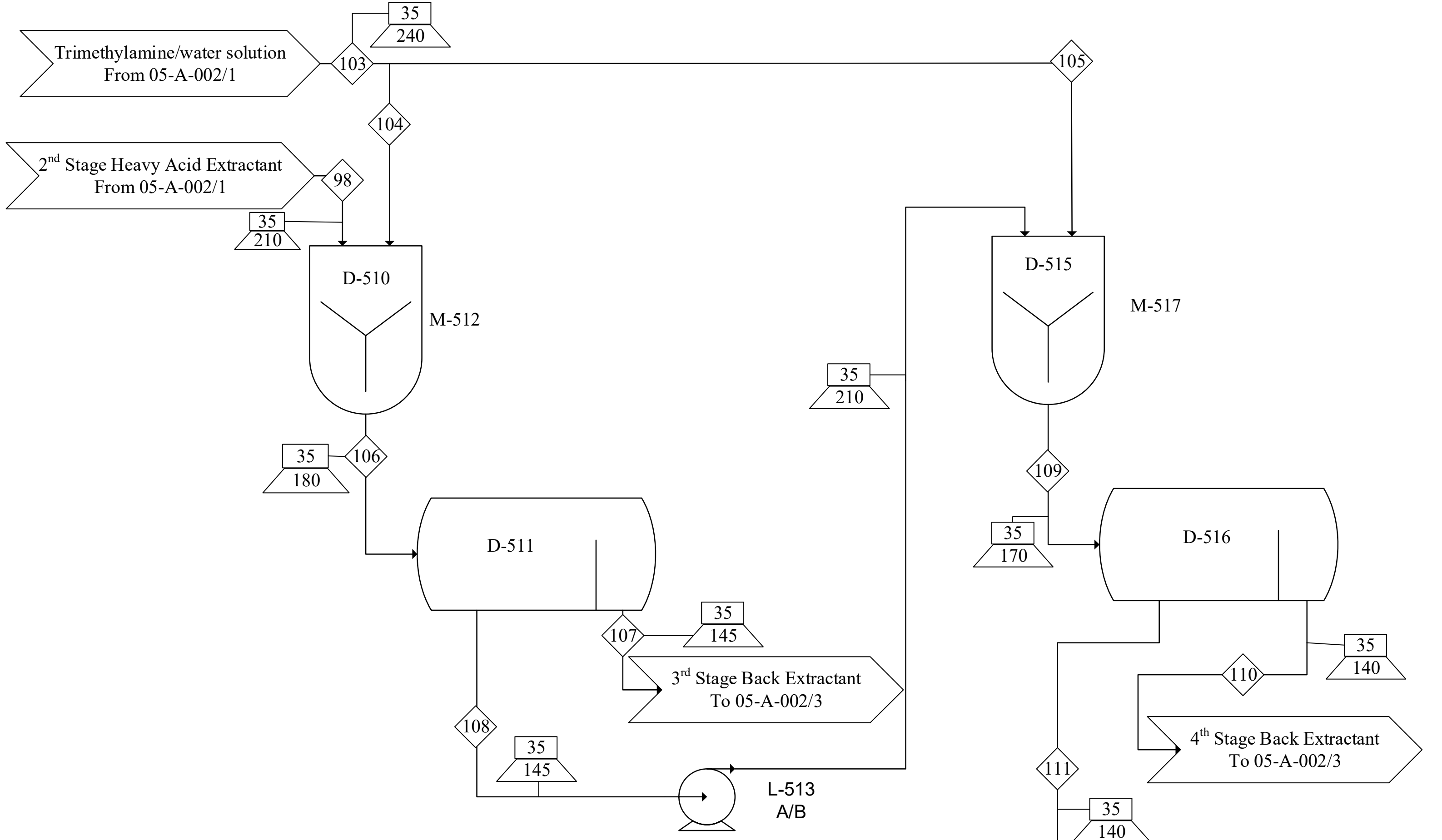
L-513 A/B
3rd Stage TMA Pump
Duty = 4.2 kW

D-515
4th Stage TMA Mixer
Height = 6.9 m
Diameter = 3.5 m

D-516
4th Stage TMA Settler
Height = 1.6 m
Diameter = 3.2 m

M-517
4th Stage TMA Mixer
Agitation Power = 32 kW

L-518 A/B
4th Stage TMA Pump
Duty = 4.2 kW



Components/Stream	98	103	104	105	106	107	108	109	110	111
1-octanol	82,000	0	0	0	82,000	0	82,000	82,000	0	47,000
Acetic Acid	1	0	0	0	1	1	0	0	0	0
Lactic Acid	20	0	0	0	20	17	3	3	3	0.05
Levulinic Acid	31	0	0	0	31	29	2	2	2	0.10
Trimethylamine	0	85	50	35	85	85	0	35	35	0
Trioctylamine	25,000	0	0	0	25,000	0	25,000	25,000	0	25,000
Water	0	65	38	27	65	65	0	27	27	0
Total	**100,000**	**150**	**88**	**62**	**100,000**	**200**	**100,000**	**100,000**	**67**	**100,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

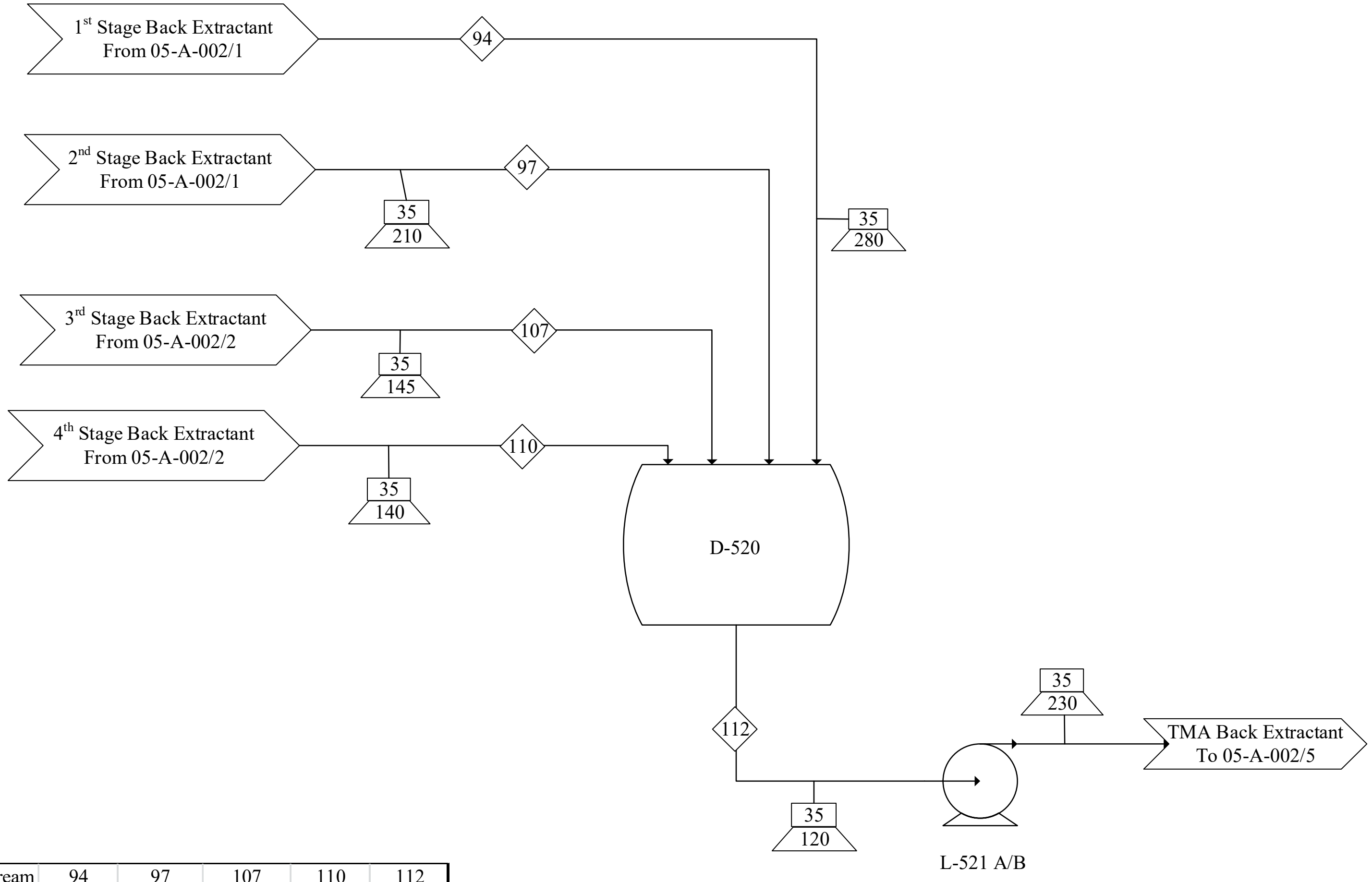
Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/2	0	406

D-520
TMA Acid Extractant
Hold Up Drum
Height = 2.3 m
Diameter = 4.7 m

L-521 A/B
TMA Hold Up Drum
Pump
Duty = 13 kW



Components/Stream	94	97	107	110	112
Acetic Acid	11	2.0	1	0	13
Lactic Acid	1,800	280	17	3	2,100
Levulinic Acid	1,700	79	29	2	1,800
Trimethylamine	4,300	420	85	35	4,800
Water	4,200	340	65	27	4,600
Total	**13,000**	**1,100**	**200**	**67**	**13,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

#	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

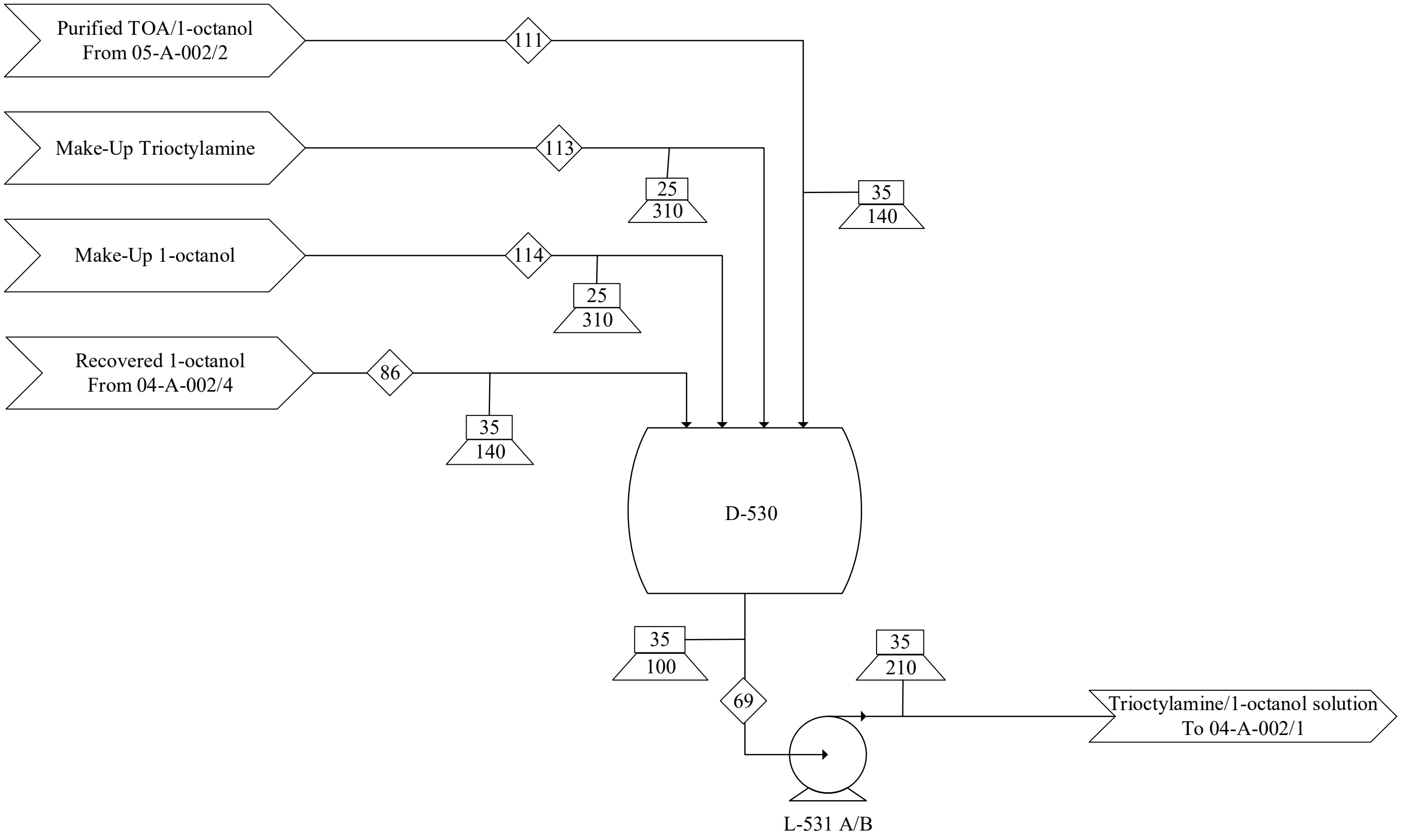
Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/3	0	407

L-531 A/B
TOA Hold Up Drum
Pump
Duty = 3.9 kW

D-530
TOA Acid Extractant
Hold Up Drum
Height = 9.6 m
Diameter = 5.9 m



Components/Stream	69	86	111	113	114
1-octanol	47,000	260	47,000	0	Trace
Acetic Acid	0	0	0	0	0
Lactic Acid	0.05	0	0.05	0	0
Levulinic Acid	0.10	0	0.10	0	0
Trioctylamine	25,000	0	25,000	Trace	0
Total	**100,000**	**260**	**100,000**	**Trace**	**Trace**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Make-up 1-octanol and trioctylamine flows are normally zero.

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
Pressure (kPa), gas
Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001 PFD 03-A-002
PFD 01-A-002 PFD 04-A-002
PFD 02-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/4	0	408

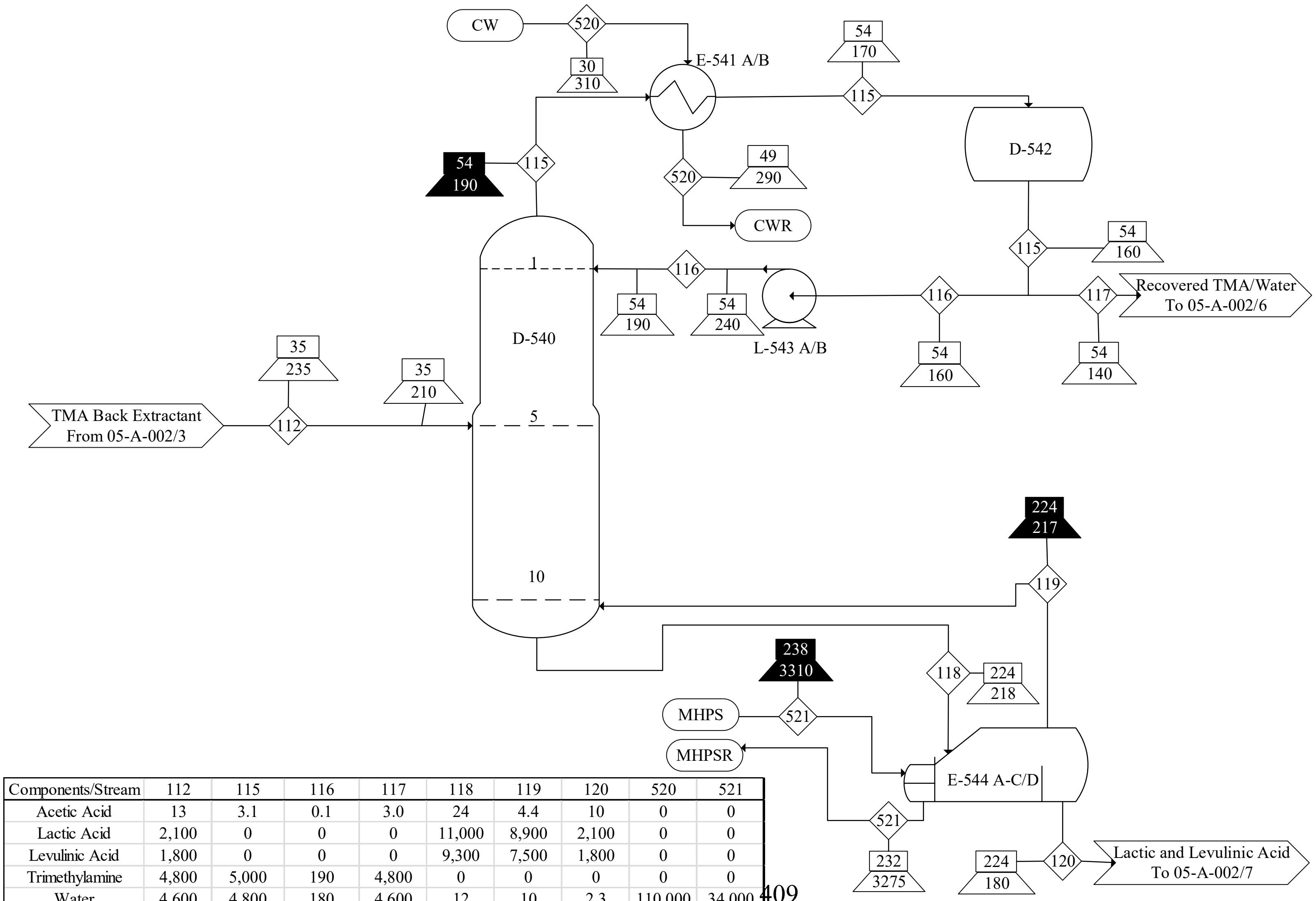
D-540
Back Extractant Solvent
Recovery Column
Rectifying Diameter: 0.91 m
Stripping Diameter: 1.3 m
Height: 7.5 m

E-541 A/B
Back Extractant Recovery
Condenser
Duty: 14 GJ/hr
Area: 270 m^2

D-542
Back Extractant Recovery
Reflux Drum
Height: 2.1 m
Length: 1.0 m

L-543 A/B
Back Extractant Recovery
Reflux Pump
Duty: 1.7 kW

E-544 A-C/D
Back Extractant Recovery
Reboiler
Duty: 17 GJ/hr
Area: 120 m^2



Components/Stream	112	115	116	117	118	119	120	520	521
Acetic Acid	13	3.1	0.1	3.0	24	4.4	10	0	0
Lactic Acid	2,100	0	0	0	11,000	8,900	2,100	0	0
Levulinic Acid	1,800	0	0	0	9,300	7,500	1,800	0	0
Trimethylamine	4,800	5,000	190	4,800	0	0	0	0	0
Water	4,600	4,800	180	4,600	12	10	2.3	110,000	34,000
Total	**13,000**	**9,800**	**370**	**9,800**	**20,000**	**1,600**	**3,900**	**110,000**	**34,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Trays have and efficiency of 70%

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

Title

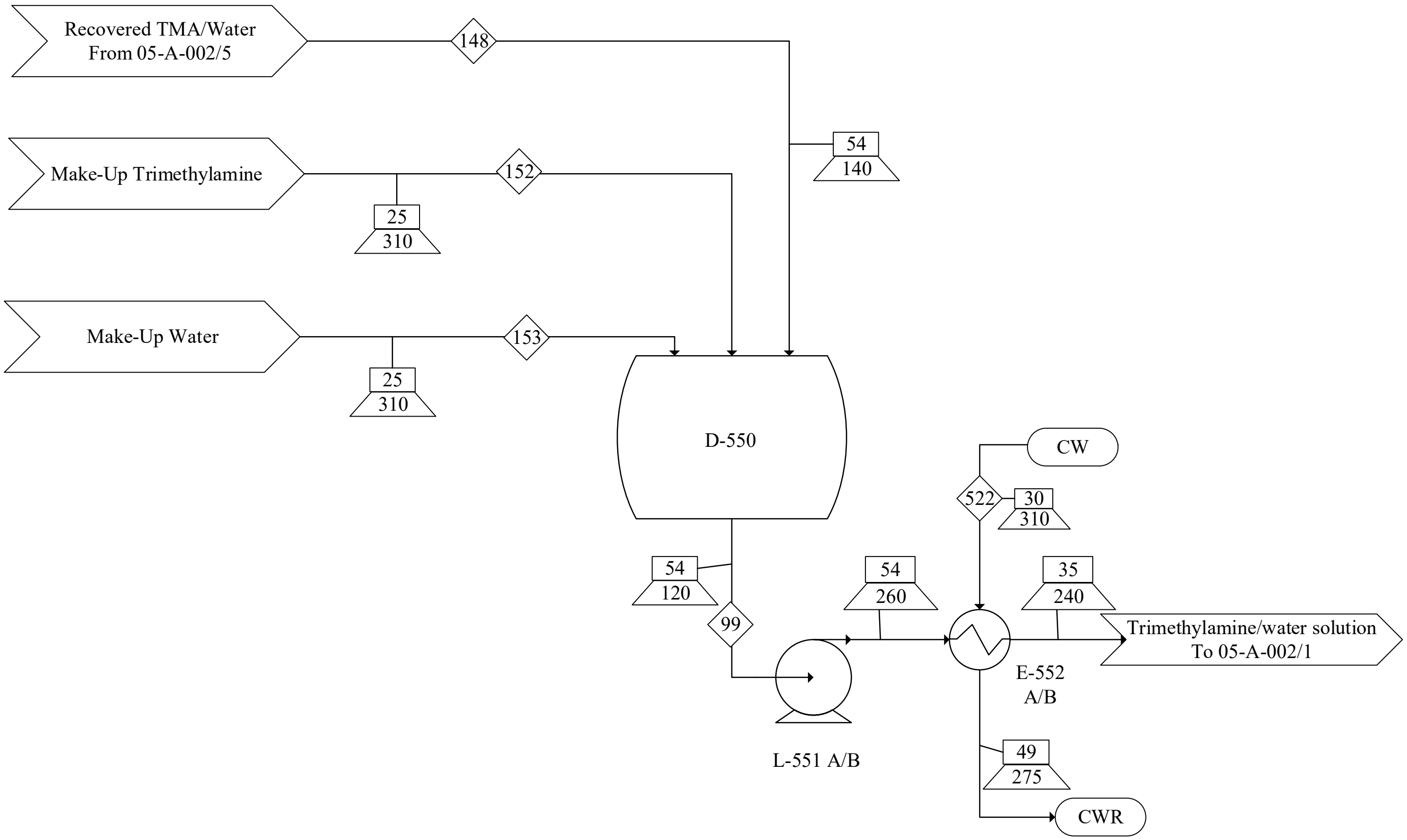
Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/5	0	409

D-550
TMA Acid Extractant
Hold Up Drum
Height = 3.2 m
Diameter = 1.6 m

L-551 A/B
TMA Hold Up Drum
Pump
Duty = 11 kW

E-552 A/B
TMA/Water Cooler
Duty = 0.91 GJ/hr
Area = 160 m^2



Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Make-up trimethylamine and water flows are normally zero.

Special Symbols / Nomenclature

◇ # Stream Identification
〉 Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
▲ # Pressure (kPa), gas
△ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001 PFD 03-A-002
PFD 01-A-002 PFD 04-A-002
PFD 02-A-002

Components/Stream	99	117	121	122	522
Acetic Acid	3.0	3.0	0	0	0
Trimethylamine	4,800	4,800	Trace	0	0
Water	4,600	4,600	0	Trace	400
Total	**9,800**	**9,800**	**Trace**	**Trace**	**400**

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/6	0	410

D-560
Lactic/Levulinic Acid Column
Rectifying Diameter: 0.49 m
Stripping Diameter: 0.83 m
Height: 26 m

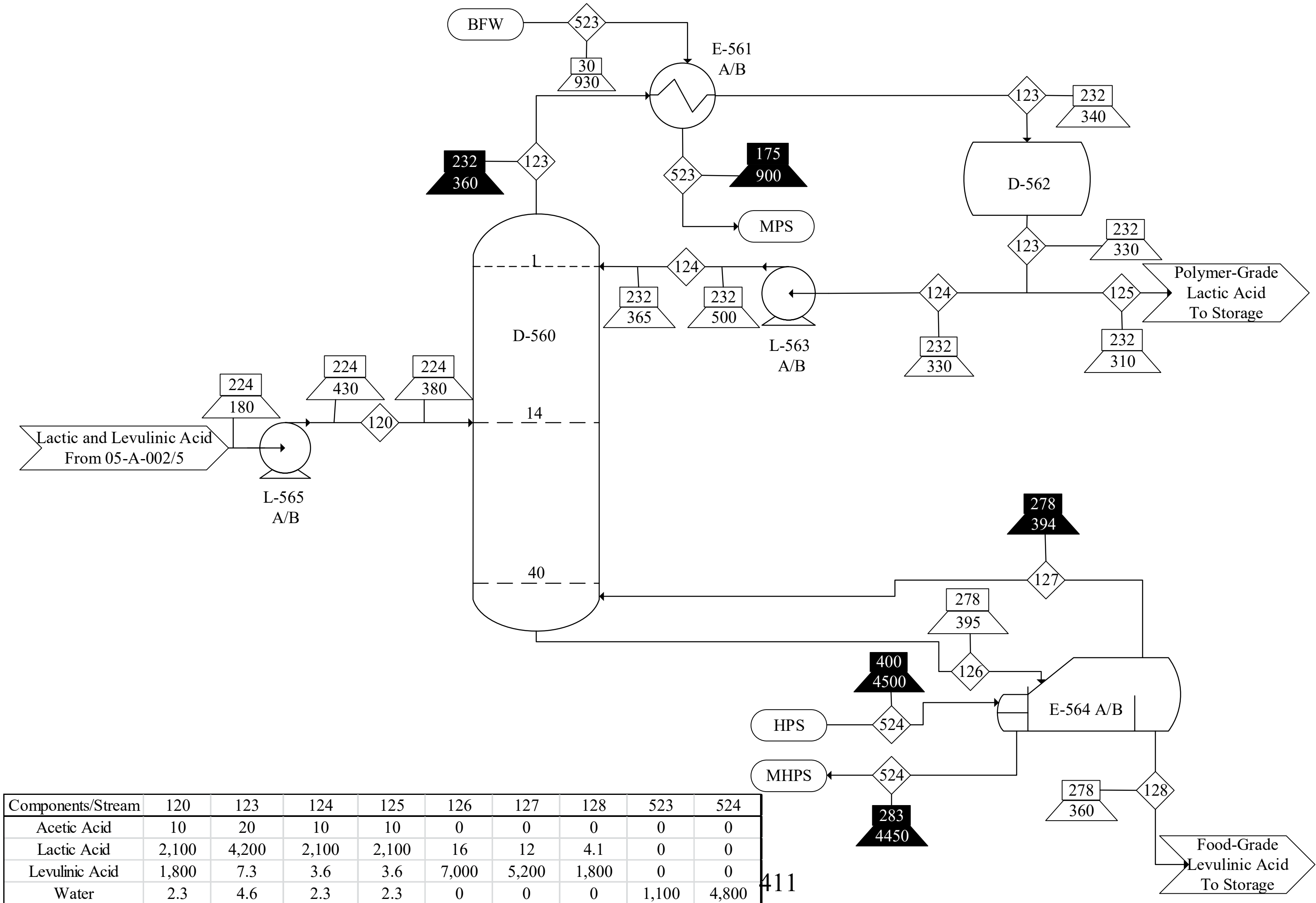
E-561 A/B
Lactic/Levulinic
Condenser
Duty: 3.0 GJ/hr
Area: 110 m^2

D-562
Lactic/Levulinic Reflux
Drum
Height: 0.98 m
Length: 0.49 m

L-563 A/B
Lactic/Levulinic
Reflux Pump
Duty: 0.74 kW

E-564 A/B
Lactic/Levulinic Reboiler
Duty: 3.6 GJ/hr
Area: 16 m^2

L-565 A/B
Lactic/Levulinic
Feed Pump
Duty: 0.74 kW



Components/Stream	120	123	124	125	126	127	128	523	524
Acetic Acid	10	20	10	10	0	0	0	0	0
Lactic Acid	2,100	4,200	2,100	2,100	16	12	4.1	0	0
Levulinic Acid	1,800	7.3	3.6	3.6	7,000	5,200	1,800	0	0
Water	2.3	4.6	2.3	2.3	0	0	0	1,100	4,800
Total	**3,900**	**4,200**	**2,100**	**2,100**	**7,000**	**5,200**	**1,800**	**1,100**	**4,800**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ # Stream Identification
〉 Input/Output
■ # Temperature (°F), gas
☐ # Temperature (°F), liquid
▲ # Pressure (kPa), gas
△ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001 PFD 03-A-002
PFD 01-A-002 PFD 04-A-002
PFD 02-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/7	0	411

Figure 45. Process Drawings for the Sn-Beta + CaSO4 Alternative

D-100
Sulfuric Acid Mixing Vessel
Height = 2.0 m
Diameter = 1.0 m

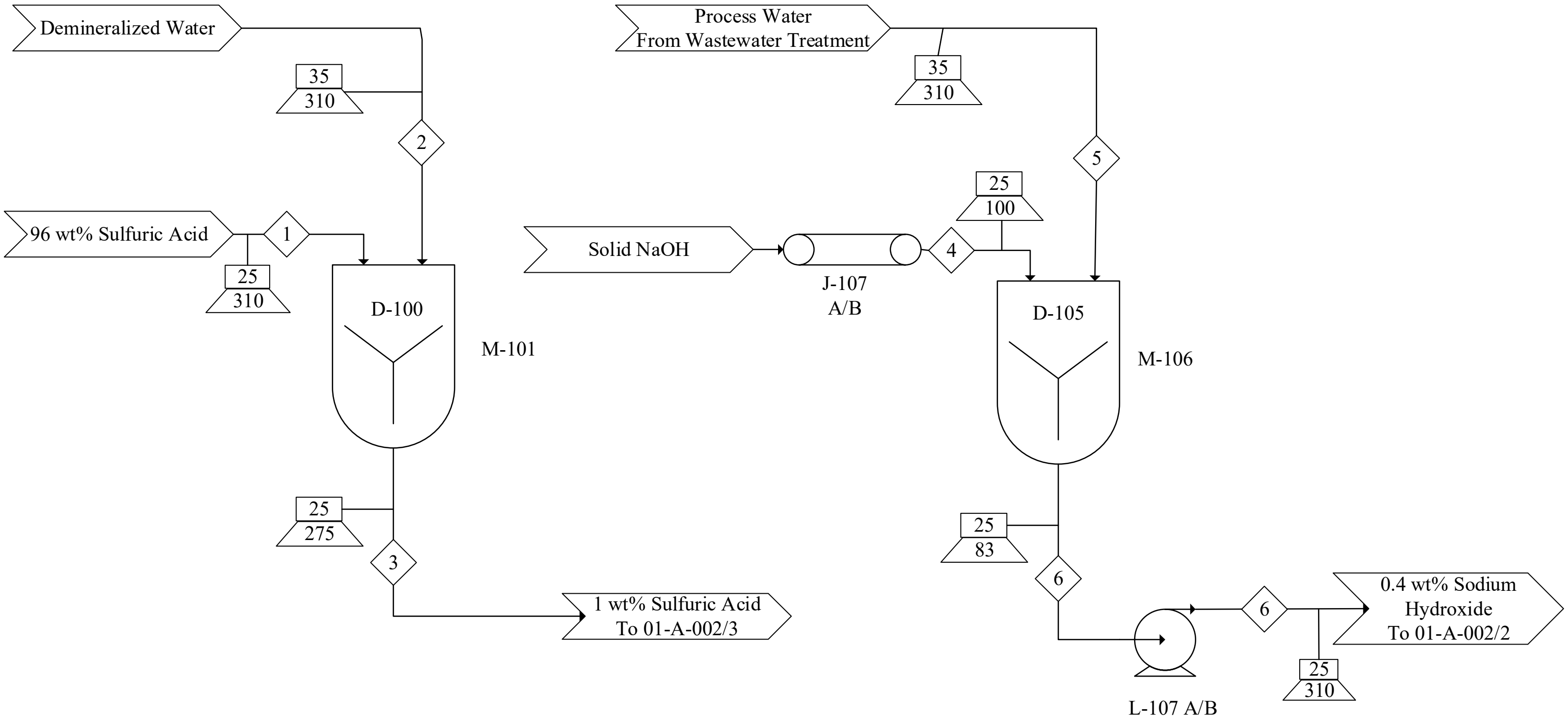
M-101
Sulfuric Acid Mixer
Agitator Power = 2.1 kW

D-105
NaOH Mixing Vessel
Height = 2.1 m
Diameter = 4.2 m

M-106
NaOH Mixer
Agitator Power = 13 kW

L-107 A/B
NaOH Feed Pump
Duty = 12 kW

J-108 A/B
NaOH Feed Belt Conveyor
Length = 150 m
Width = 0.61 m
Duty = 1.0 hp



Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type
Process Flow Diagram

Title
Process Area 01: Corn Stover Preteatment

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	01-A-002/1	0	413

Components/Stream	1	2	3	4	5	6
Sodium Hydroxide	0	0	0	450	0	450
Sulfuric Acid	180	0	180	0	0	0
Water	7.7	18,000	18,000	0	110,000	110,000
Total	**190**	**18,000**	**18,000**	**450**	**110,000**	**110,000**

D-110 A-C/D
Deacetylation Vessel
Height = 8.5 m
Diameter = 4.3 m

J-111 A-C/D
Deacetylation Screw
Conveyor
Duty = 3.7 kW
Length = 15 m
Width = 0.45 m

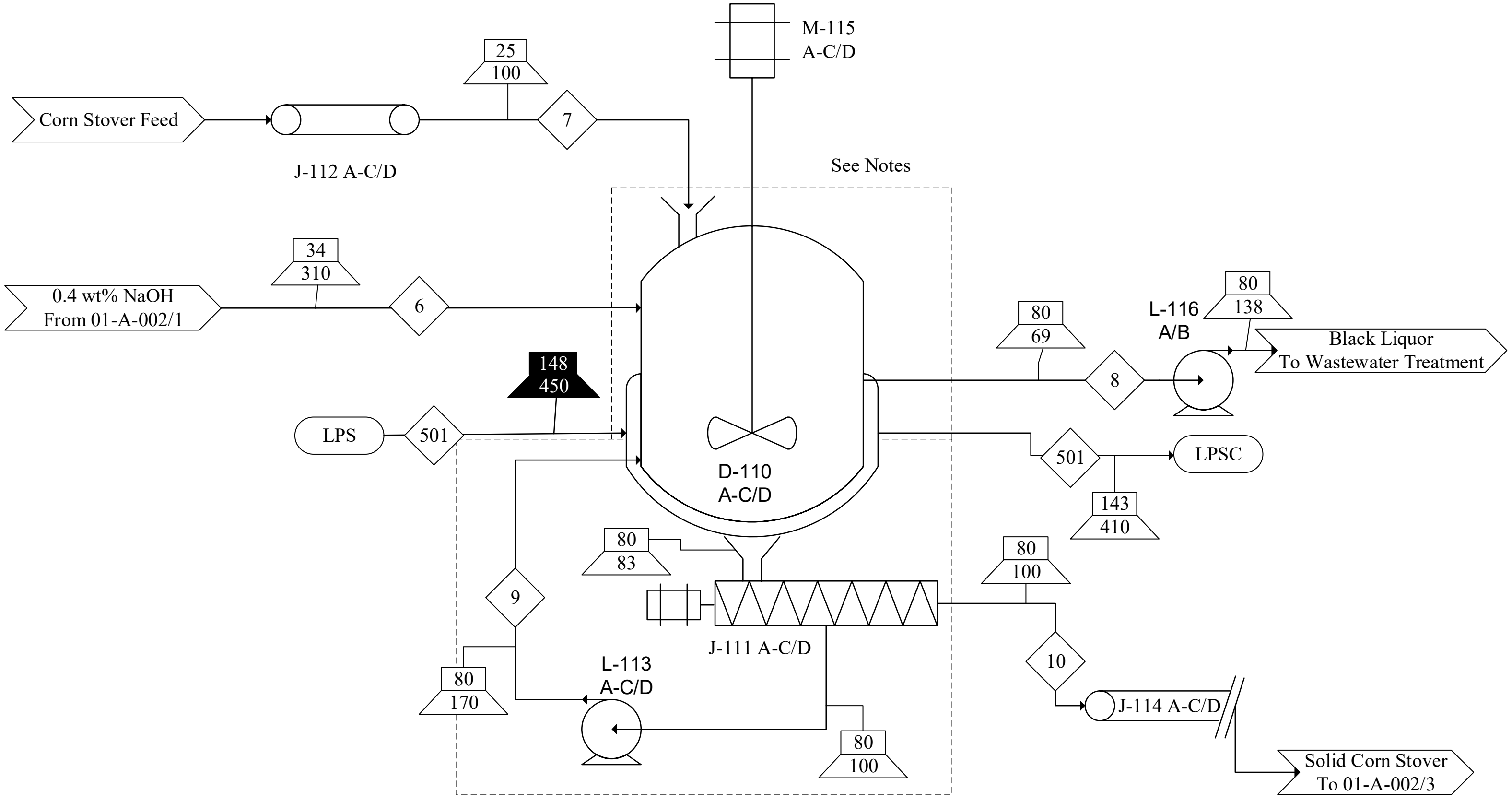
J-112 A-C/D
Pre-Deacetylation Corn
Stover Belt Conveyor
Duty = 2.2 kW
Length = 1.5 m
Width = 0.45 m

L-113 A-C/D
Pre-Deacetylation Corn
Stover Pump
Duty = 0.74 kW

J-114 A-C/D
Post-Deacetylation
Corn Stover Belt
Conveyor
Duty = 2.2 kW
Length = 1.5 m
Width = 2.0 0.45 m

M-115 A-C/D
Deacetylation Mixer
Agitation Power= 96 kW

L-116 A/B
Black Liquor Pump
Duty = 3.7 kW



Components/Stream	6	7	8	9	10	501
Acetate	0	0	430	0	0	0
Ash	0	0	1,000	0	0	0
Corn Stover	0	27,000	0	0	21,000	0
Extractables	0	0	4,000	0	0	0
Lignin	0	0	860	0	0	0
Sodium Hydroxide	450	0	450	0	0	0
Water	110,000	0	110,000	4,200	0	10,000
Xylan	0	0	100	0	0	0
Total	**110,000**	**27,000**	**120,000**	**4,200**	**21,000**	**10,000**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. D-100 is a semi-batch process, see supplemental PFD 01-A-003 for more information.

5. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 04-A-002
BFD 01-A-001	PFD 05-A-002
PFD 02-A-002	Supplemental
PFD 03-A-002	PFD 01-A-003

Drawing Type
Process Flow Diagram

Title

Process Area 01: Corn Stover Preteatment

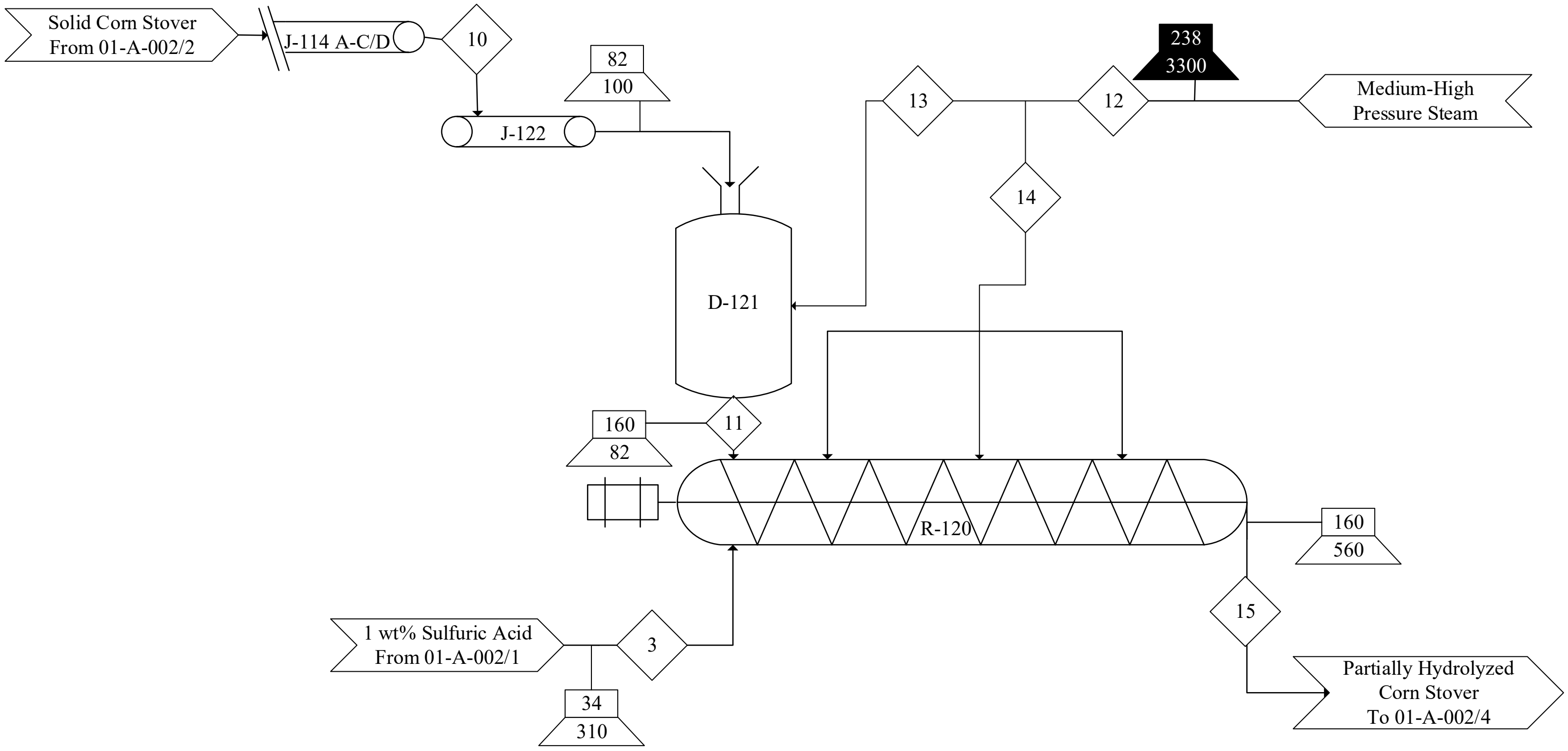
By/Date	DWG # / Sheet #	Rev	Page
AJK 7/05/20	01-A-002/2	0	414

D-121
Acid Presteamer
Height = 5.8 m
Diameter = 2.9 m

R-120
Acid Hydrolysis
Reactor
Length = 10 m
Diameter = 1.0 m
Power = 29 kW

J-122
Pre-Hydrolysis
Corn Stover Belt
Conveyor
Power = 1.5 kW
Length = 15 m
Width = 6.1 m

J-114 A-C/D
Post-Deacetylation
Corn Stover Belt
Conveyor
Duty = 2.2 kW
Length = 15 m
Width = 6.1 m



Components/Stream	3	10	11	12	13	14	15
Acetic Acid	0	0	0	0	0	0	59
Corn Stover	0	21,000	21,000	0	0	0	14,000
Furfural	0	0	0	0	0	0	190
Glucose	0	0	0	0	0	0	1,100
Glucose Oligomer	0	0	0	0	0	0	280
HMF	0	0	0	0	0	0	260
Sulfuric Acid	180	0	0	0	0	0	180
Water	18,000	0	1,000	40,000	1,000	39,000	59,000
Xylose	0	0	0	0	0	0	5,500
Xylose Oligomer	0	0	0	0	0	0	82
Total	**18,000**	**214,000**	**22,000**	**40,000**	**1,000**	**39,000**	**82,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. High pressure steam is directly injected into the reactor to keep it isothermal.
5. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type
Process Flow Diagram

Title

Process Area 01: Corn Stover Pretreatment

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	01-A-002/3	0	415

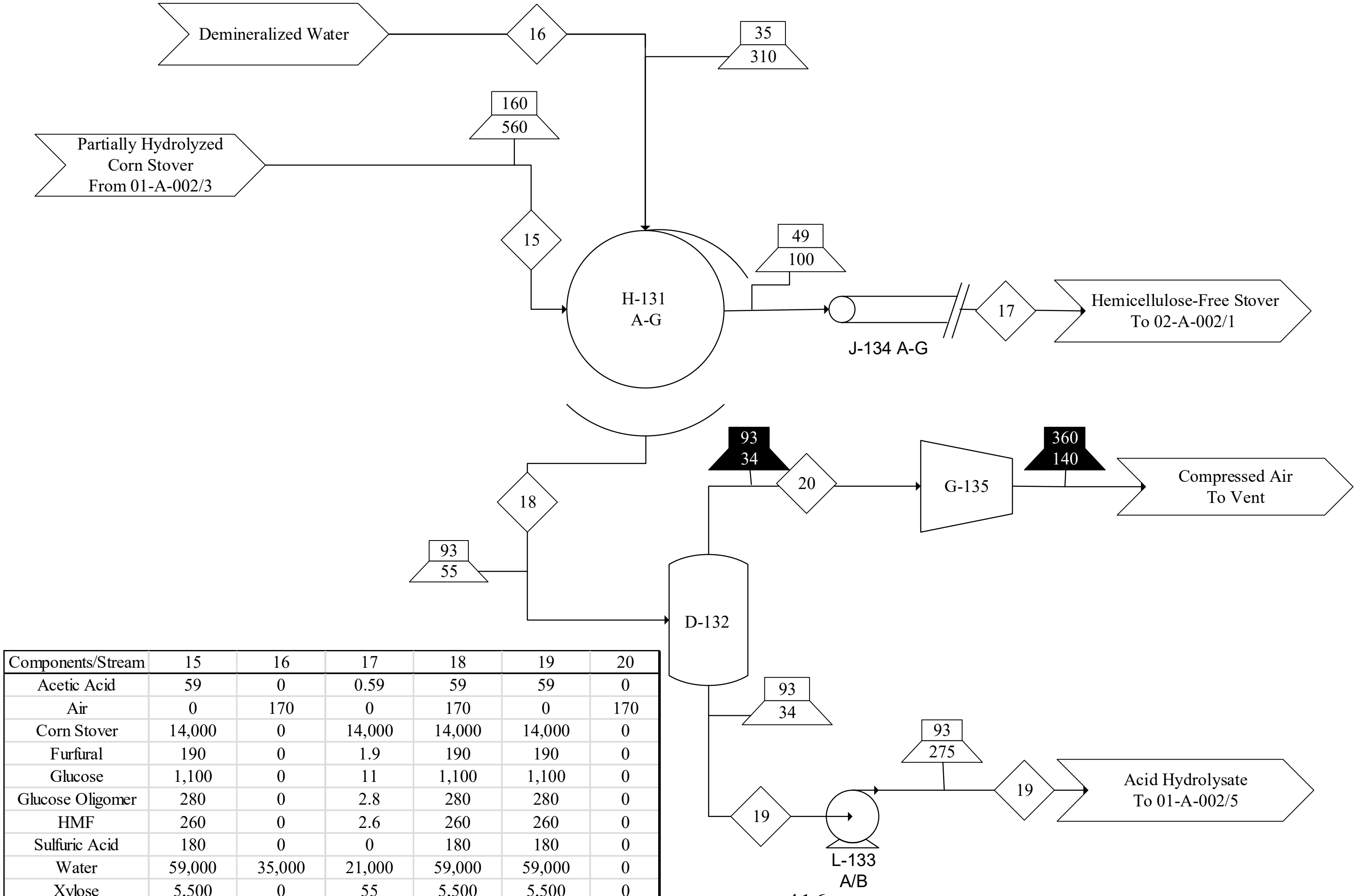
H-131, A-G
Acid Rotary Vacuum Filter
Duty = 1.5 kW
Volume = 15 m^3

D-132
Acid Filtration Knockout Drum
Diameter = 4.0 m
Height =7.9 m

L-133 A/B
Acid Filtration Pump
Duty = 19 kW

J-134 A-G
Acid Filtration Corn Stover Belt Conveyor
Power = 1.5 kW
Length = 150 m
Width =0.61 m

G-135
Air Compressor
Duty = 13 kW



Components/Stream	15	16	17	18	19	20
Acetic Acid	59	0	0.59	59	59	0
Air	0	170	0	170	0	170
Corn Stover	14,000	0	14,000	14,000	14,000	0
Furfural	190	0	1.9	190	190	0
Glucose	1,100	0	11	1,100	1,100	0
Glucose Oligomer	280	0	2.8	280	280	0
HMF	260	0	2.6	260	260	0
Sulfuric Acid	180	0	0	180	180	0
Water	59,000	35,000	21,000	59,000	59,000	0
Xylose	5,500	0	55	5,500	5,500	0
Xylose Oligomer	82	0	82	82	82	0
Total	**82,000**	**35,000**	**35,000**	**82,000**	**82,000**	**170**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. H-121 is run at vacuum conditions to remove liquids from the solids.
5. Process water washes the stover to ensure high product recovery.
6. Pumps have an efficiency of 65%
7. Compressors have an efficiency of 65%.

Special Symbols / Nomenclature

◇# Stream Identification
〉 Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001 PFD 04-A-002
PFD 02-A-002 PFD 05-A-002
PFD 03-A-002

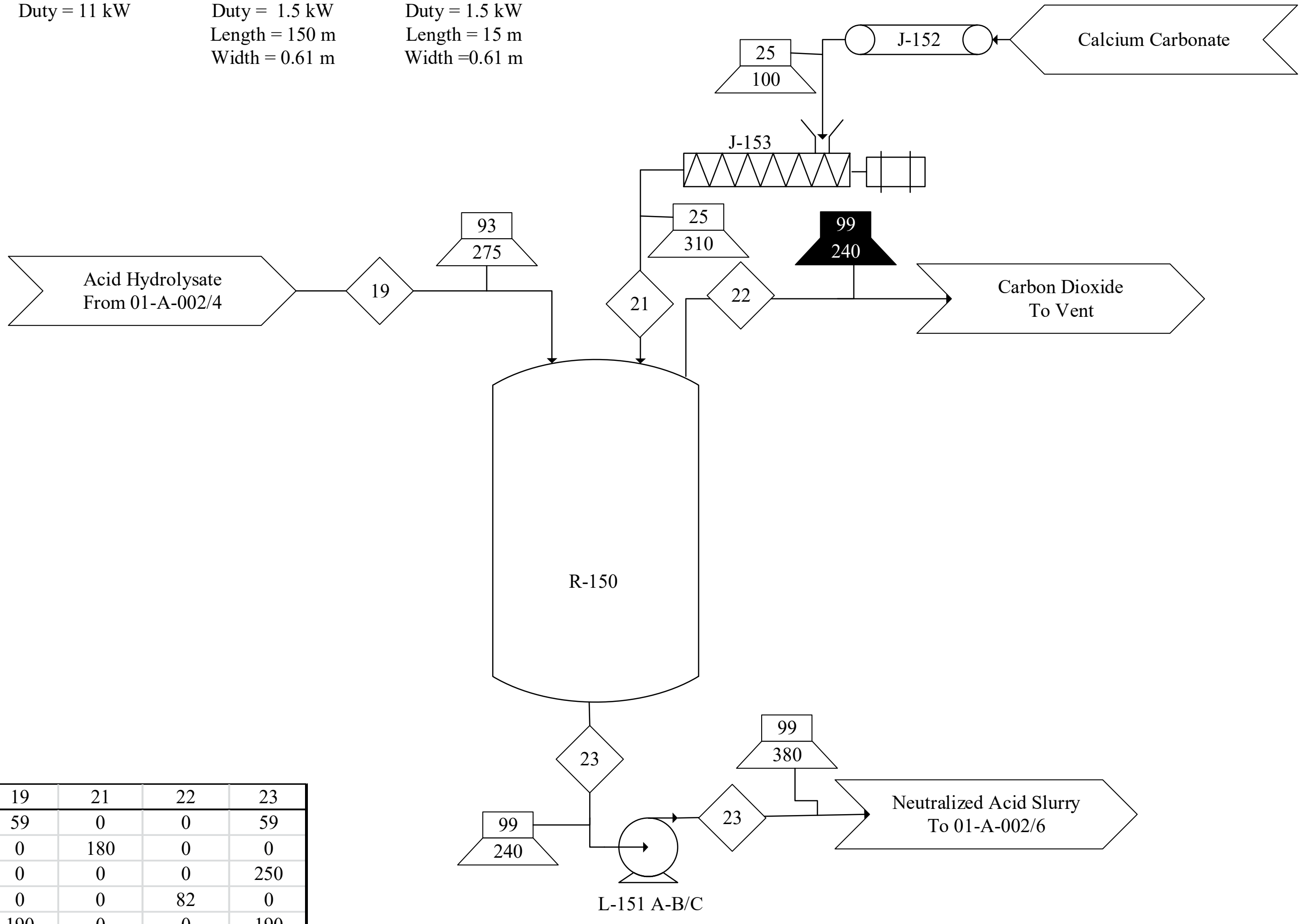
Drawing Type
Process Flow Diagram

Title

Process Area 01: Corn Stover Pretreatment

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	01-A-002/4	0	416

R-150	L-151 A-B/C	J-152	J-153
Neutralization Reactor	Acid Phase Separation Slurry Pump	Calcium Carbonate Feed Conveyor	Calcium Carbonate Screw Conveyor
Diameter =5.8 m	Duty = 11 kW	Duty = 1.5 kW	Duty = 1.5 kW
Height = 11 m		Length = 150 m	Length = 15 m
		Width = 0.61 m	Width =0.61 m



Components/Stream	19	21	22	23
Acetic Acid	59	0	0	59
Calcium Carbonate	0	180	0	0
Calcium Sulfate	0	0	0	250
Carbon Dioxide	0	0	82	0
Furfural	190	0	0	190
Glucose	1,100	0	0	1,100
Glucose Oligomer	280	0	0	280
HMF	260	0	0	260
Sulfuric Acid	180	0	0	0
Water	59,000	0	0	59,000
Xylose	5,500	0	0	5,500
Xylose Oligomer	82	0	0	82
Total	**82,000**	**180**	**42**	**82,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇#	Stream Identification
⟩	Input/Output
■#	Temperature (°C), gas
□#	Temperature (°C), liquid
▲#	Pressure (kPa), gas
△#	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001 PFD 04-A-002
PFD 02-A-002 PFD 05-A-002
PFD 03-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 01: Corn Stover Pretreatment

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	01-A-002/5	0	417

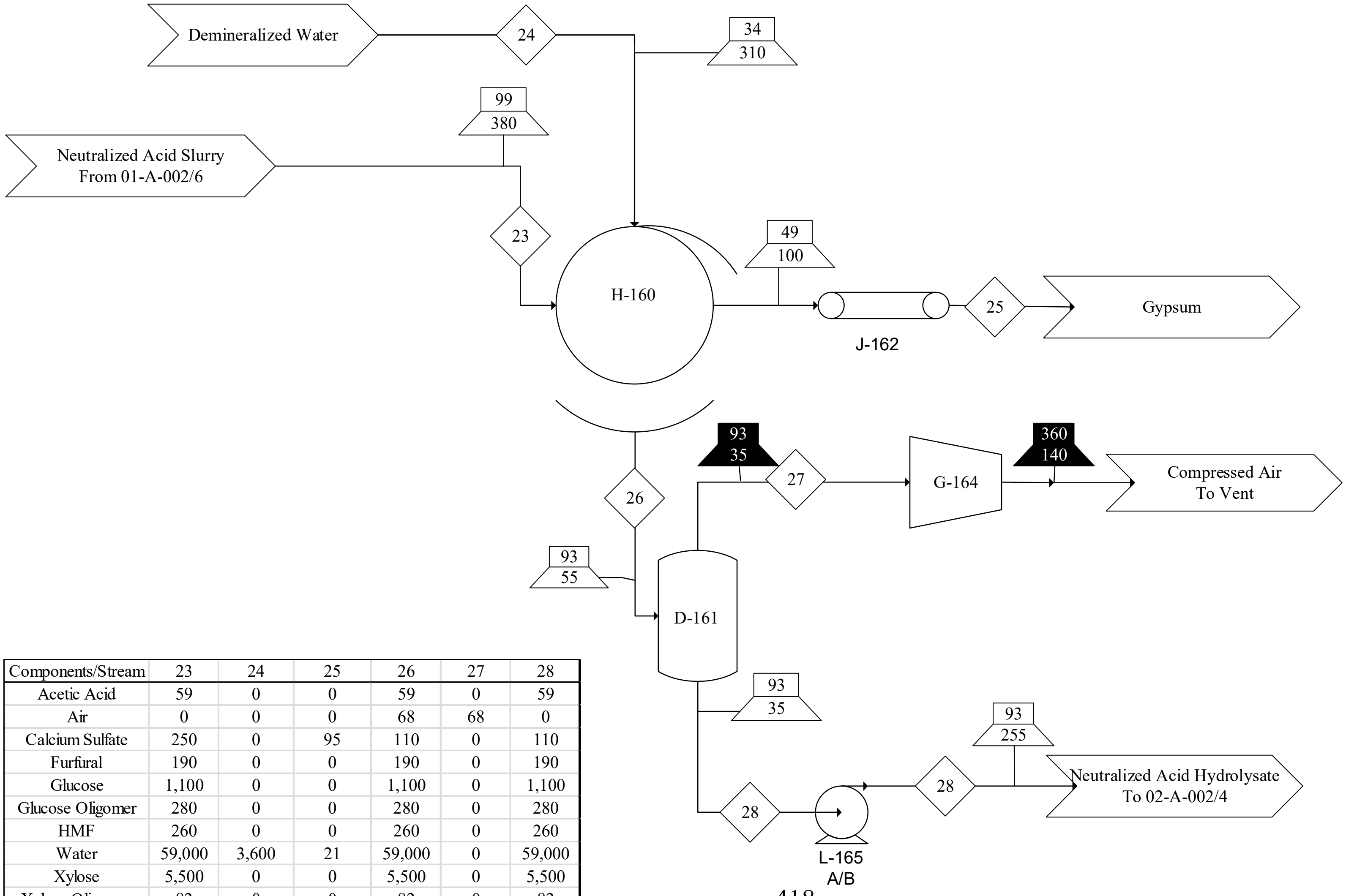
H-160
Gypsum Rotary
Vacuum Filtration
Duty = 0.74 kW
Volume = 5.1 m^3

D-161
Gypsum Filtration
Knockout Drum
Diameter = 3.1 m
Height = 6.1 m

J-162
Gypsum Filtration
Corn Stover Belt
Conveyor
Power = 1.5 kW
Length = 150 m
Width = 0.61 m

L-165 A/B
Gypsum Filtration
Pump
Duty = 46 kW

G-164
Gypsum Filtration
Air Compressor
Duty = 4.4 kW



Components/Stream	23	24	25	26	27	28
Acetic Acid	59	0	0	59	0	59
Air	0	0	0	68	68	0
Calcium Sulfate	250	0	95	110	0	110
Furfural	190	0	0	190	0	190
Glucose	1,100	0	0	1,100	0	1,100
Glucose Oligomer	280	0	0	280	0	280
HMF	260	0	0	260	0	260
Water	59,000	3,600	21	59,000	0	59,000
Xylose	5,500	0	0	5,500	0	5,500
Xylose Oligomer	82	0	0	82	0	82
Total	**82,000**	**3,600**	**120**	**82,000**	**68**	**82,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 01-A-001
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

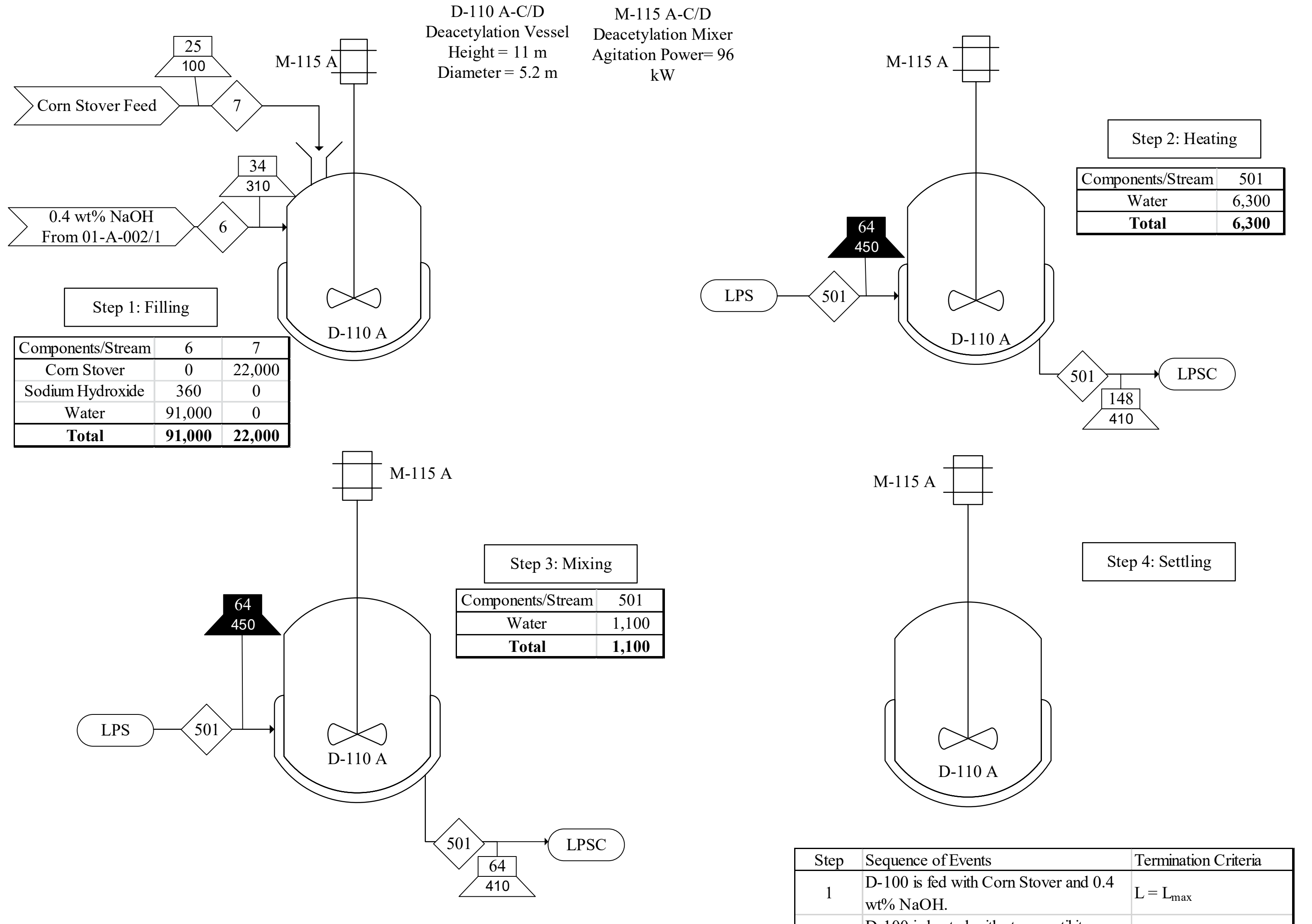
Drawing Type

Process Flow Diagram

Title

Process Area 01: Corn Stover Pretreatment

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	01-A-002/7	0	418



Step 1: Filling

Components/Stream	6	7
Corn Stover	0	22,000
Sodium Hydroxide	360	0
Water	91,000	0
Total	**91,000**	**22,000**

Step 2: Heating

Components/Stream	501
Water	6,300
Total	**6,300**

Step 3: Mixing

Components/Stream	501
Water	1,100
Total	**1,100**

Step 4: Settling

Step	Sequence of Events	Termination Criteria
1	D-100 is fed with Corn Stover and 0.4 wt% NaOH.	$L = L_{max}$
2	D-100 is heated with steam until it reaches 180° F	$T = T_{operating}$
3	D-100 is held at the operating temperature and is continually stired	$C_{Ash\ (aq)} > C_{Ash\ (aq),\ max}$
4	Stirring is stopped and the solids are allowed to settle to the bottom	$\rho_{(aq)} < \rho_{(aq),\ min}$

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 10 minutes
5. Step 2 is 10 minutes
6. Step 3 is 60 minutes
7. Step 4 is 5 minute
8. Three vessels are used in parallel for continual deacetylation.
9. Each vessel cycle are staggered by 0.8 hours.

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

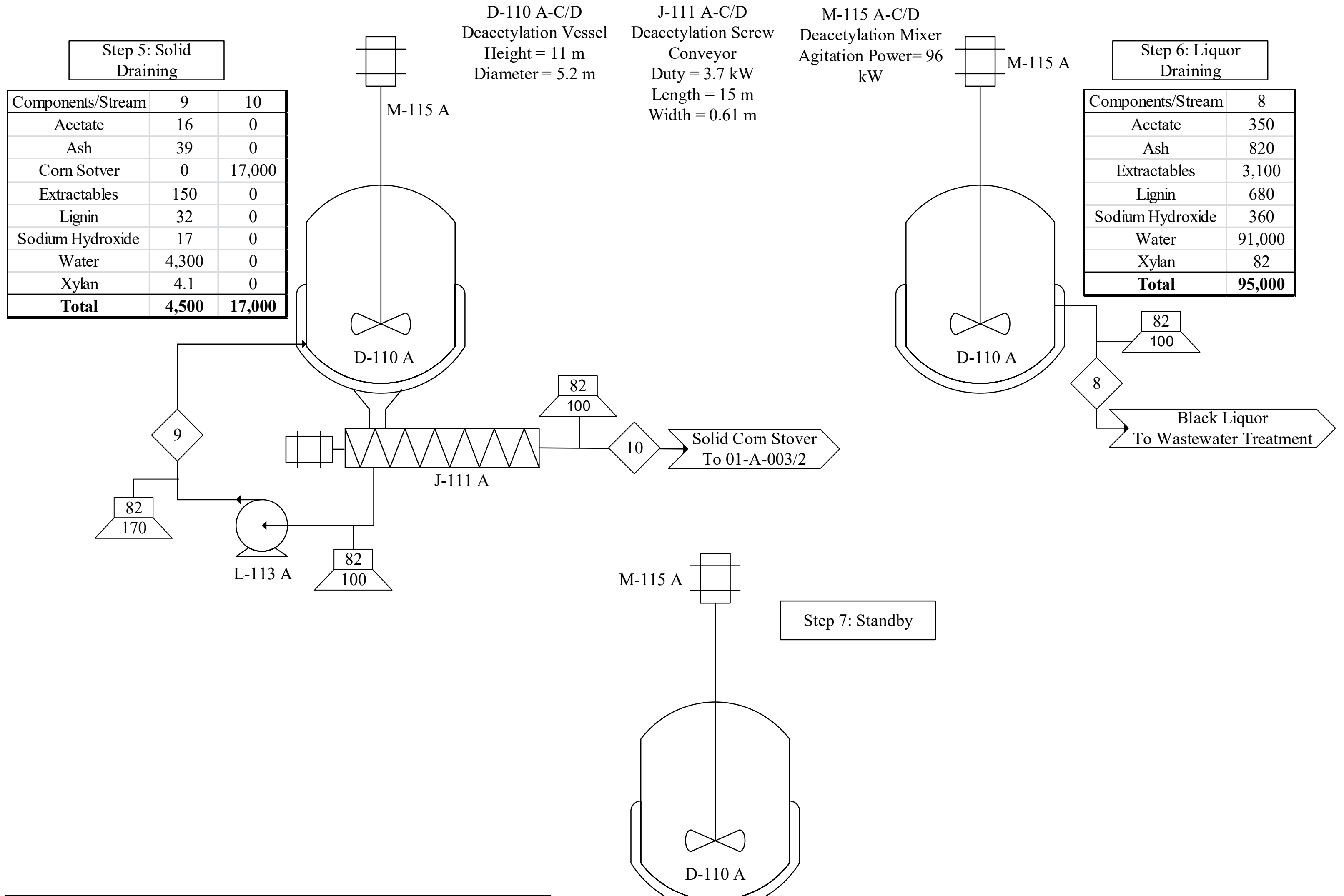
Drawing Type

Supplemental Process Flow Diagram

Title

Deacetylation Semi-Batch Steps

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	01-A-003/1	0	419



Components/Stream	9	10
Acetate	16	0
Ash	39	0
Corn Sotver	0	17,000
Extractables	150	0
Lignin	32	0
Sodium Hydroxide	17	0
Water	4,300	0
Xylan	4.1	0
Total	**4,500**	**17,000**

Components/Stream	8
Acetate	350
Ash	820
Extractables	3,100
Lignin	680
Sodium Hydroxide	360
Water	91,000
Xylan	82
Total	**95,000**

Step	Sequence of Events	Termination Criteria
5	Solids are screw conveyed from the bottom of the tank. Recovered liquids are recycled back to the vessel.	$L_{solids} < L_{solids,\ min}$
6	The remaining liquids within the vessel are drained.	$L_{liquids} < L_{liquids,\ min}$
7	D-100 is on standby.	$C_{Ash\ (aq)} > 1/2\ C_{Ash\ (aq),\ max}$ of vessel in Step 3

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 5 is 48 minutes
5. Step 6 is 10 minutes
6. Step 7 is 1 minute
7. Three vessels are used in parallel for continual deacetylation.

Special Symbols / Nomenclature

◇ #	Stream Identification
⌦	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

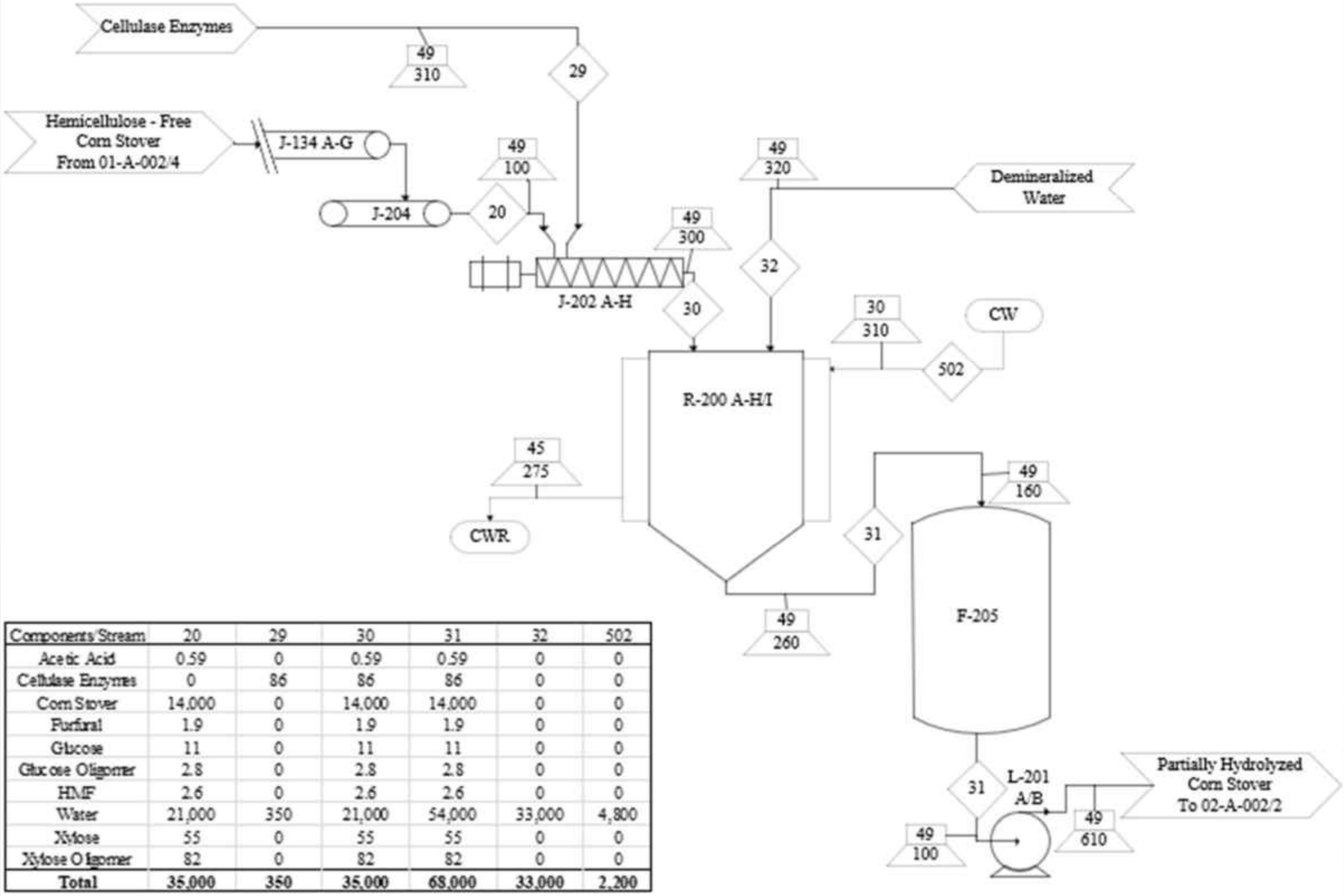
Supplemental Process Flow Diagram

Title

Deacetylation Semi-Batch Steps

By/Date	DWG # / Sheet #	Rev	Page
AJK 7/05/20	01-A-003/2	0	420

R-200 A-H/I	L-201 A/B	J-202 A-H	J-204	F-205	J-134 A-G
Continuous, High-Solids Enzymatic Reactor	Enzymatic Slurry Pump	Pre-Enzymatic Screw Conveyor	Corn Stover Belt Conveyor	Pre-Batch Hold Up	Acid Filtration Corn Stover Belt Conveyor
Height = 17 m	Duty = 33 kW	Duty = 2.2 kW	Duty = 1.5 kW	Height: 18 m	Power = 1.5 kW
Diameter= 4.3 m		Length = 15 m	Length = 15 m	Diameter: 9.1 m	Length = 150 m
		Width = 0.45 m	Width = 0.61 m		Width = 0.61 m



Components/Stream	20	29	30	31	32	502
Acetic Acid	0.59	0	0.59	0.59	0	0
Cellulase Enzymes	0	86	86	86	0	0
Corn Stover	14,000	0	14,000	14,000	0	0
Furfural	1.9	0	1.9	1.9	0	0
Glucose	11	0	11	11	0	0
Glucose Oligomer	2.8	0	2.8	2.8	0	0
HMF	2.6	0	2.6	2.6	0	0
Water	21,000	350	21,000	54,000	33,000	4,800
Xylose	55	0	55	55	0	0
Xylose Oligomer	82	0	82	82	0	0
Total	**35,000**	**350**	**35,000**	**68,000**	**33,000**	**2,200**

Notes:

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Due to high solids concentration enzymatic reaction starts in a flow reactor. When the stream reaches the bottom of the reactor (24 hrs) it is a pump-able concertation.
6. Conveyors have an efficiency of 88%.

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 02-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 03-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 02: Glucose Recovery

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	02-A-002/1	0	421

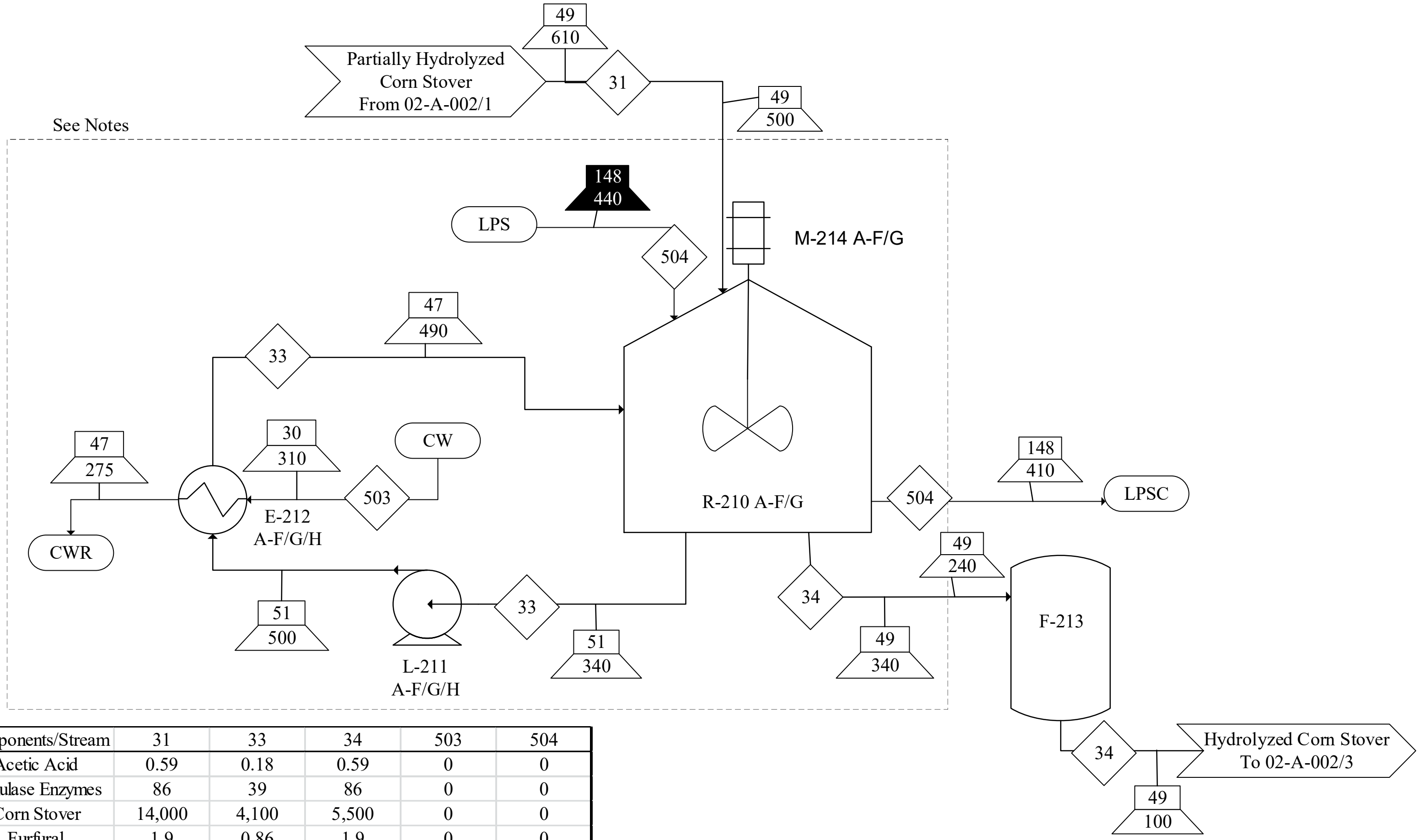
R-210 A-F/G
Batch Enzymatic Reactor
Height = 16 m
Diameter= 8.2 m

L-211 A-F/G/H
Batch Hydrolysis Recycle Pump
Duty = 1.5 kW

E-212 A-F/G/H
Batch Hydrolysis Recycle Heat Exchanger
Duty = 0.057 GJ/hr
Area = 1.7 m^2

F-213
Post-Batch Hold Up
Height: 18 m
Diameter: 9.1 m

M-214 A-F/G
Batch Hydrolysis Mixer
Agitation Power= 430 kW



Components/Stream	31	33	34	503	504
Acetic Acid	0.59	0.18	0.59	0	0
Cellulase Enzymes	86	39	86	0	0
Corn Stover	14,000	4,100	5,500	0	0
Furfural	1.9	0.86	1.9	0	0
Glucose	11	820	8,100	0	0
Glucose Oligomer	2.8	33	350	0	0
HMF	2.6	1	2.6	0	0
Water	54,000	18,000	54,000	9,500	250
Xylose	55	18	55	0	0
Xylose Oligomer	82	0	82	0	0
Total	**68,000**	**23,000**	**68,000**	**9,500**	**250**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. R-210 is a semi-batch process, more information can be seen on Supplemental PFD 02-A-003.
6. Pressure drops in stream 38 due to the 54 foot height increase to get to the top of the reactor.

Special Symbols / Nomenclature

#	Stream Identification
#	Input/Output
#	Temperature (°F), gas
#	Temperature (°F), liquid
#	Pressure (kPa), gas
#	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 04-A-002
BFD 02-A-001	PFD 05-A-002
PFD 01-A-002	Supplemental
PFD 03-A-002	PFD 02-A-003

Drawing Type

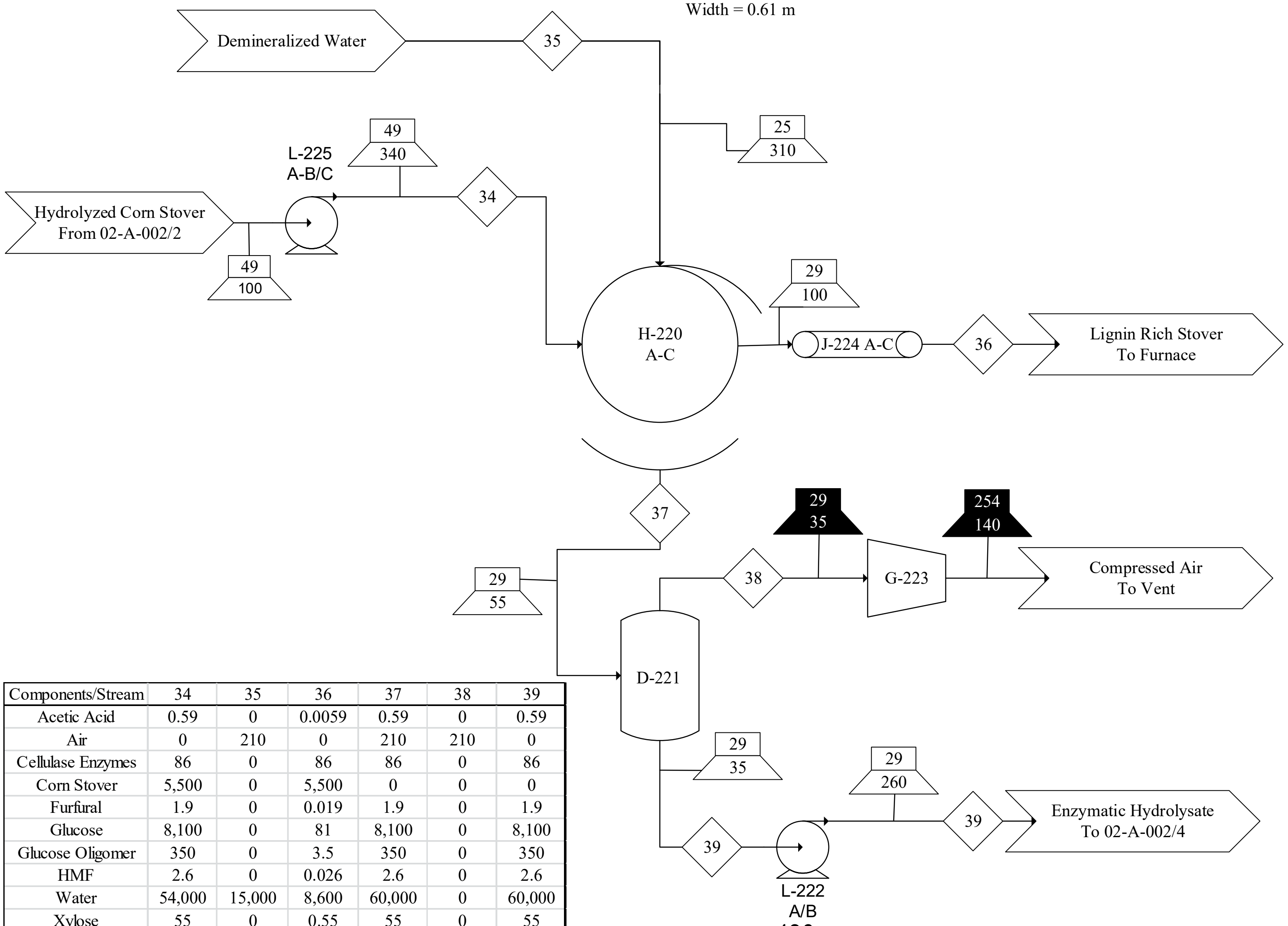
Process Flow Diagram

Title

Process Area 02: Glucose Recovery

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	02-A-002/2	0	422

H-220, A-C	D-221	L-222 A/B	G-223	J-224 A-C	L-225 A-B/C
Enzymatic Rotary Vacuum Filtration	Enzymatic Filtration Knockout Drum	Enzymatic Filtration Pump	Enzymatic Filtration Air Compressor	Lignin Rich Belt Conveyor	Enzymatic Filtration Feed Slurry Pump
Duty = 1.5 kW Volume = 540 ft^3	Diameter = 3.1 m Height = 6.7 m	Duty = 20 kW	Duty 13 kW	Power = 1.5 kW Length = 150 m Width = 0.61 m	Duty = 4.5 kW



Components/Stream	34	35	36	37	38	39
Acetic Acid	0.59	0	0.0059	0.59	0	0.59
Air	0	210	0	210	210	0
Cellulase Enzymes	86	0	86	86	0	86
Corn Stover	5,500	0	5,500	0	0	0
Furfural	1.9	0	0.019	1.9	0	1.9
Glucose	8,100	0	81	8,100	0	8,100
Glucose Oligomer	350	0	3.5	350	0	350
HMF	2.6	0	0.026	2.6	0	2.6
Water	54,000	15,000	8,600	60,000	0	60,000
Xylose	55	0	0.55	55	0	55
Xylose Oligomer	82	0	0.82	82	0	82
Total	**68,000**	**15,000**	**14,000**	**68,000**	**210**	**68,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Compressors have an efficiency of 65%.

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 02-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 03-A-002

Drawing Type

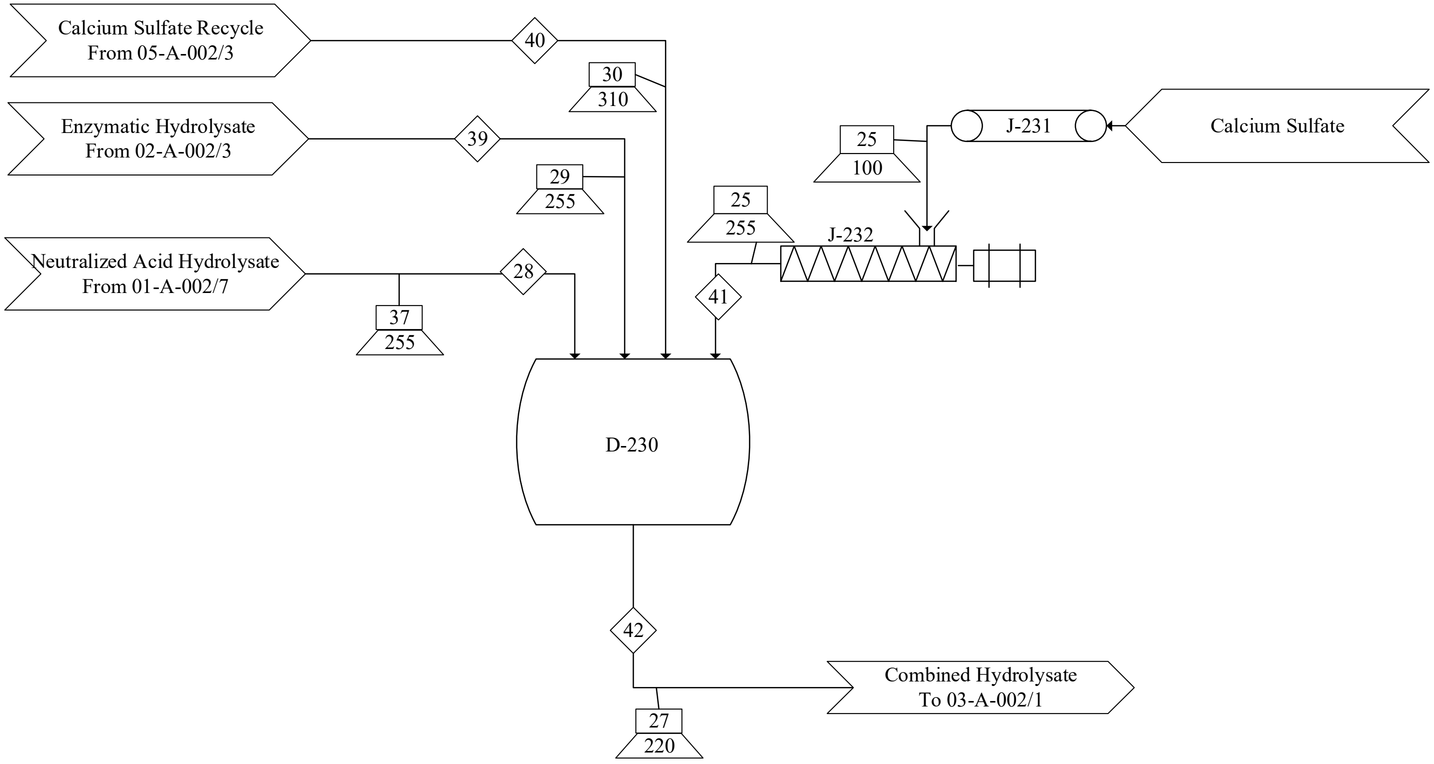
Process Flow Diagram

Title

Process Area 02: Glucose Recovery

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	02-A-002/3	0	423

D-230	J-231	J-232
Hydrolysate Mixing Drum	Calcium Sulfate Feed Conveyor	Calcium Sulfate Screw Conveyor
Diameter = 2.5 m	Duty = 1.5 kW	Duty = 1.5 kW
Height = 4.9 m	Length = 150 m	Length = 15 m
	Width = 0.61 m	Width =0.61 m



Components/Stream	28	39	40	41	42
Acetic Acid	59	0.59	0	0	59
Calcium Sulfate	110	0	31	130	270
Furfural	190	1.9	0	0	190
Glucose	1,100	8,100	0	0	9,200
Glucose Oligomer	280	350	0	0	630
HMF	260	2.6	0	0	260
Water	59,000	60,000	23,000	0	140,000
Xylose	5,500	55	0	0	5,500
Xylose Oligomer	82	82	0	0	180
Total	**82,000**	**68,000**	**23,000**	**130**	**150,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇#

⮞

■#

☐#

▲#

△#

Reference Drawings:

I/O 00-A-001
BFD 02-A-001
PFD 01-A-002
PFD 03-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 02: Glucose Recovery

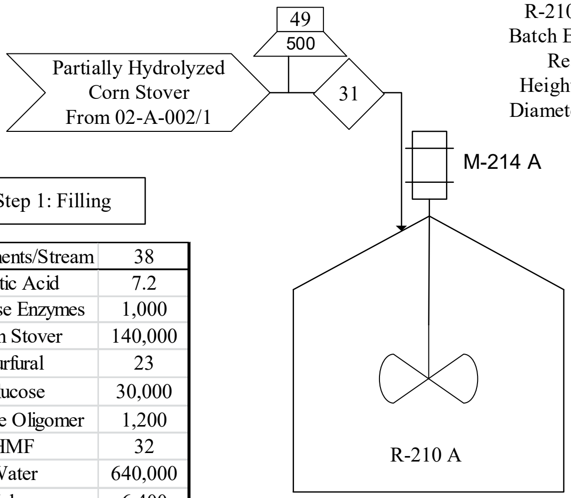
By/Date	DWG #/ Sheet #	Rev	
AJK 7/05/20	02-A-002/4	0	424

R-210 A-F/G
Batch Enzymatic Reactor
Height = 16 m
Diameter= 8.2 m

L-211 A-F/G/H
Batch Hydrolysis Recycle Pump
Duty = 1.5 kW

E-212 A-F/G/H
Batch Hydrolysis Recycle Heat Exchanger
Duty = 0.056 GJ/hr
Area = 1.7 m^2

M-214 A-F/G
Batch Hydrolysis Mixer
Agitation Power= 430 kW

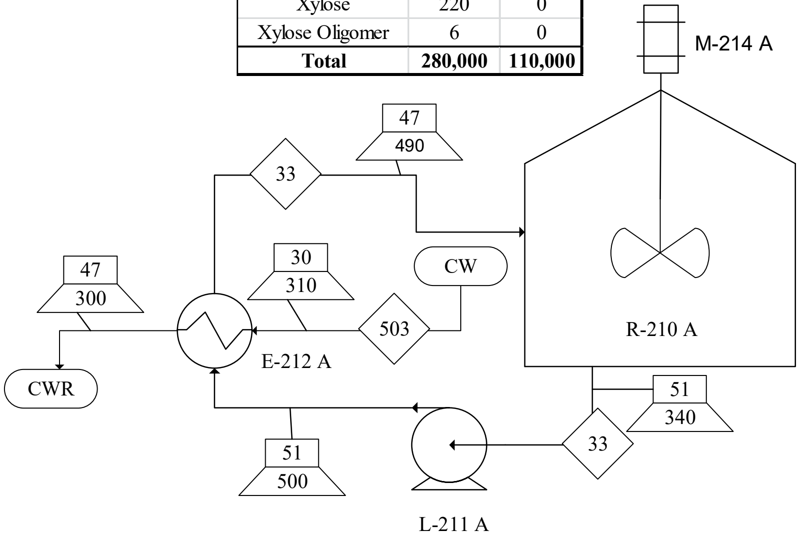


Step 1: Filling

Components/Stream	38
Acetic Acid	7.2
Cellulase Enzymes	1,000
Corn Stover	140,000
Furfural	23
Glucose	30,000
Glucose Oligomer	1,200
HMF	32
Water	640,000
Xylose	6,400
Xylose Oligomer	9.5
Total	**820,000**

Step 2: Hydrolysis

Components/Stream	33	503
Acetic Acid	5.5	0
Cellulase Enzymes	350	0
Corn Stover	50,000	0
Furfural	7.7	0
Glucose	10,000	0
Glucose Oligomer	400	0
HMF	11	0
Water	220,000	110,000
Xylose	220	0
Xylose Oligomer	6	0
Total	**280,000**	**110,000**



Step	Sequence of Events	Termination Criteria
1	R-210 is filled with partially hydrolyzed corn stover from the continuous reactor (R-200).	$L = L_{max}$
2	R-210 is stirred and its contents recycled to maintain vessel temperature.	$C_{\text{Glucose (aq)}} > C_{\text{Gkucose (aq), min}}$
3	R-210 is drained of its contents.	$L_{\text{liquids}} < L_{\text{liquids, min}}$
4	R-210 is filled with low pressure steam to steralize the reactor.	$P_{\text{vessel}} = P_{\text{vessel, max}}$
5	R-210 is on standby.	Vessel ahead is 11 hours into reaction

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 1 hour
5. Step 2 is 60 hours
6. Step 3 is 1 hour
7. Step 4 is 1 hour
8. Step 5 is 3 hours
9. Six reactors are used in parallel for continual hydrolysis.

Special Symbols / Nomenclature

◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

Supplemental Process Flow Diagram

Title

Enzymatic Hydrolysis Semi-Batch Steps

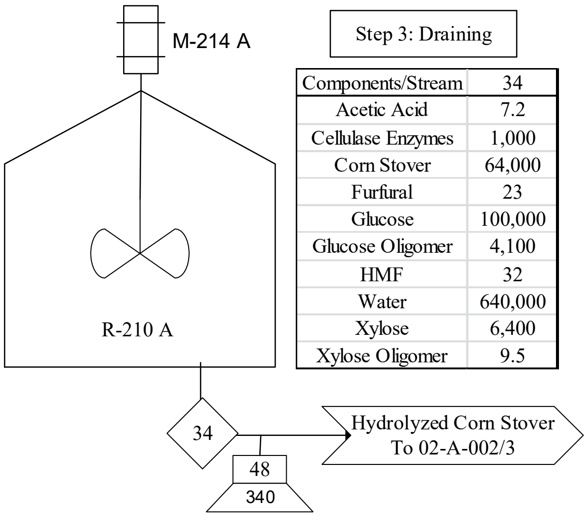
By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	02-A-003/1	0	425

R-210 A-F/G
Batch Enzymatic Reactor
Height = 16 m
Diameter= 8.2 m

E-212 A-F/G/H
Batch Hydrolysis Recycle Heat Exchanger
Duty = 0.057 GJ/hr
Area = 1.7 m^2

M-214 A-F/G
Batch Hydrolysis Mixer
Agitation Power= 430 kW

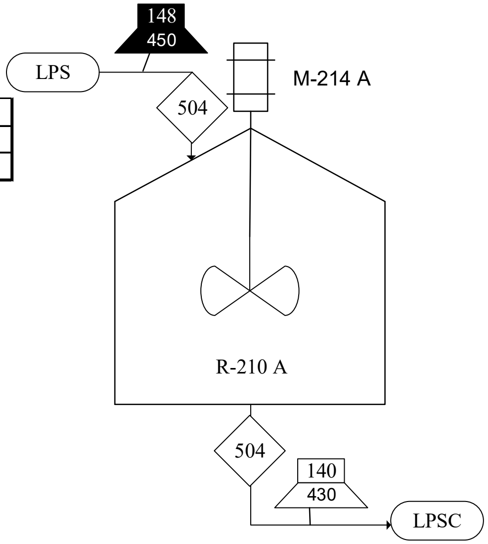
Step 3: Draining



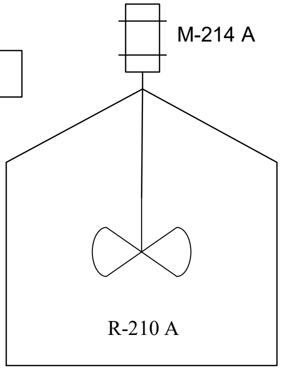
Components/Stream	34
Acetic Acid	7.2
Cellulase Enzymes	1,000
Corn Stover	64,000
Furfural	23
Glucose	100,000
Glucose Oligomer	4,100
HMF	32
Water	640,000
Xylose	6,400
Xylose Oligomer	9.5

Step 4: Sterilization

Components/Stream	504
Water	2,700
Total	**2,700**



Step 5: Standby



Step	Sequence of Events	Termination Criteria
1	R-210 is filled with partially hydrolyzed corn stover from the continuous reactor (R-200).	$L = L_{max}$
2	R-210 is stirred and its contents recycled to maintain vessel temperature.	$C_{Glucose\ (aq)} > C_{Gkucose\ (aq),\ min}$
3	R-210 is drained of its contents.	$L_{liquids} < L_{liquids,\ min}$
4	R-210 is filled with low pressure steam to steralize the reactor.	$P_{vessel} = P_{vessel,\ max}$
5	R-210 is on standby.	Vessel ahead is 11 hours into reaction

426

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 1 hour
5. Step 2 is 60 hours
6. Step 3 is 1 hour
7. Step 4 is 1 hour
8. Step 5 is 3 hours
9. Six reactors are used in parallel for continual hydrolysis.

Special Symbols / Nomenclature

◇ #	Stream Identification
⇨	Input/Output
■ #	Temperature (°C), gas
□ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

Supplemental Process Flow Diagram

Title

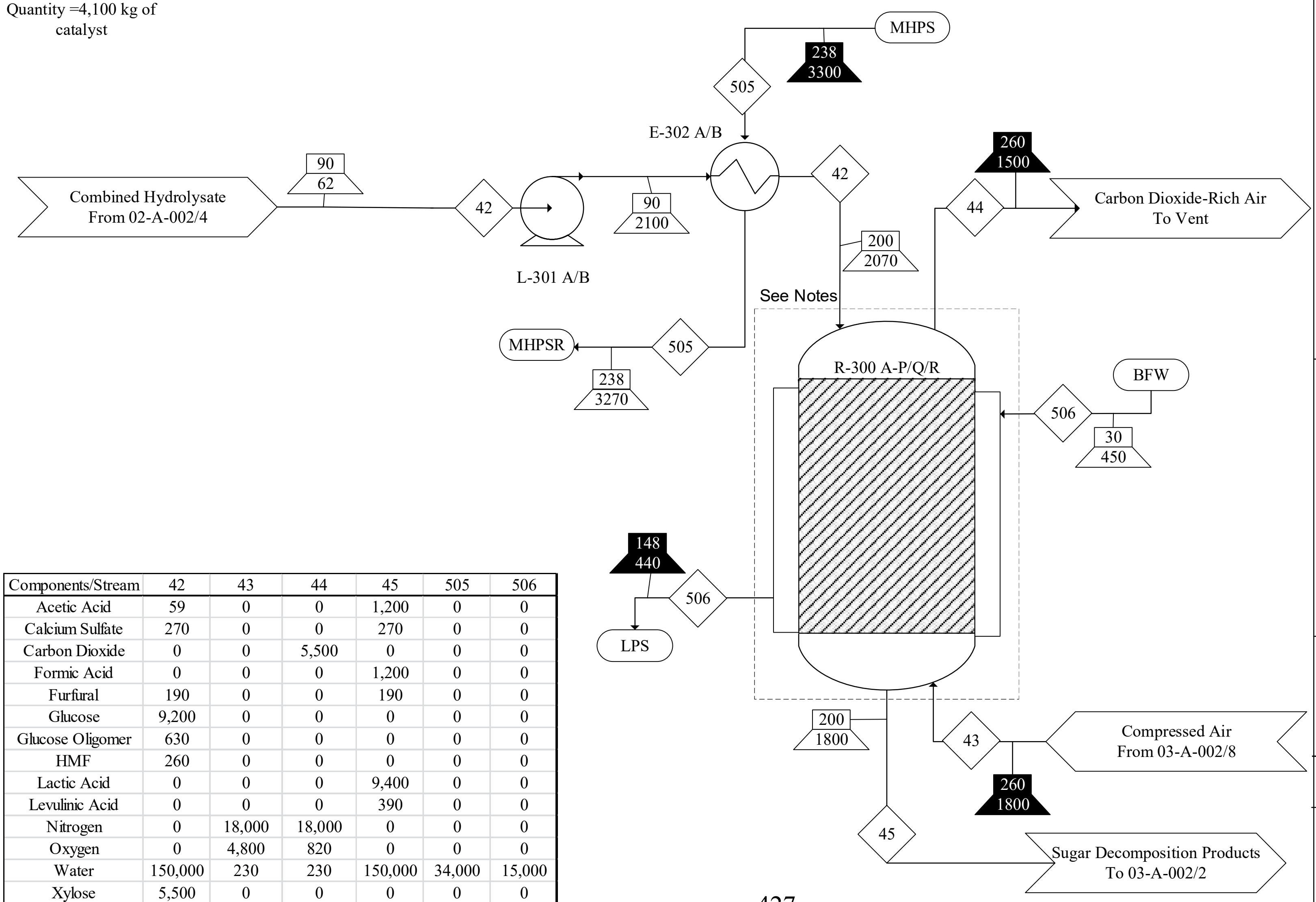
Enzymatic Hydrolysis Semi-Batch Steps

By/Date	DWG # / Sheet #	Rev	Page
AJK 7/05/20	02-A-003/2	0	426

R-300 A-P/Q/R
Sn2+ Beta Zeolite
Packed Bed Reactor
Diameter = 1.8 m
Height = 15 m
Quantity =4,100 kg of catalyst

L-301 A/B
Pre-Catalytic
Decomposition Pump
Duty = 140 kW

E-302 A/B
Catalytic Decomposition
Preheater
Duty = 94 GJ/hr
Area = 710 m^2



Components/Stream	42	43	44	45	505	506
Acetic Acid	59	0	0	1,200	0	0
Calcium Sulfate	270	0	0	270	0	0
Carbon Dioxide	0	0	5,500	0	0	0
Formic Acid	0	0	0	1,200	0	0
Furfural	190	0	0	190	0	0
Glucose	9,200	0	0	0	0	0
Glucose Oligomer	630	0	0	0	0	0
HMF	260	0	0	0	0	0
Lactic Acid	0	0	0	9,400	0	0
Levulinic Acid	0	0	0	390	0	0
Nitrogen	0	18,000	18,000	0	0	0
Oxygen	0	4,800	820	0	0	0
Water	150,000	230	230	150,000	34,000	15,000
Xylose	5,500	0	0	0	0	0
Xylose Oligomer	82	0	0	0	0	0
Total	**160,000**	**23,000**	**25,000**	**160,000**	**34,000**	**15,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Due to coke formation, two extra reactors are present and the catalyst will need to be calcined every two weeks.
6. Coke formation makes the reactor a semi-batch process, see drawing 03-A-003 for more details.

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
□# Temperature (°C), liquid
Pressure (kPa), gas
Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 04-A-002
BFD 03-A-001	PFD 05-A-002
PFD 01-A-002	Supplemental
PFD 02-A-002	PFD 03-A-003

Drawing Type
Process Flow Diagram

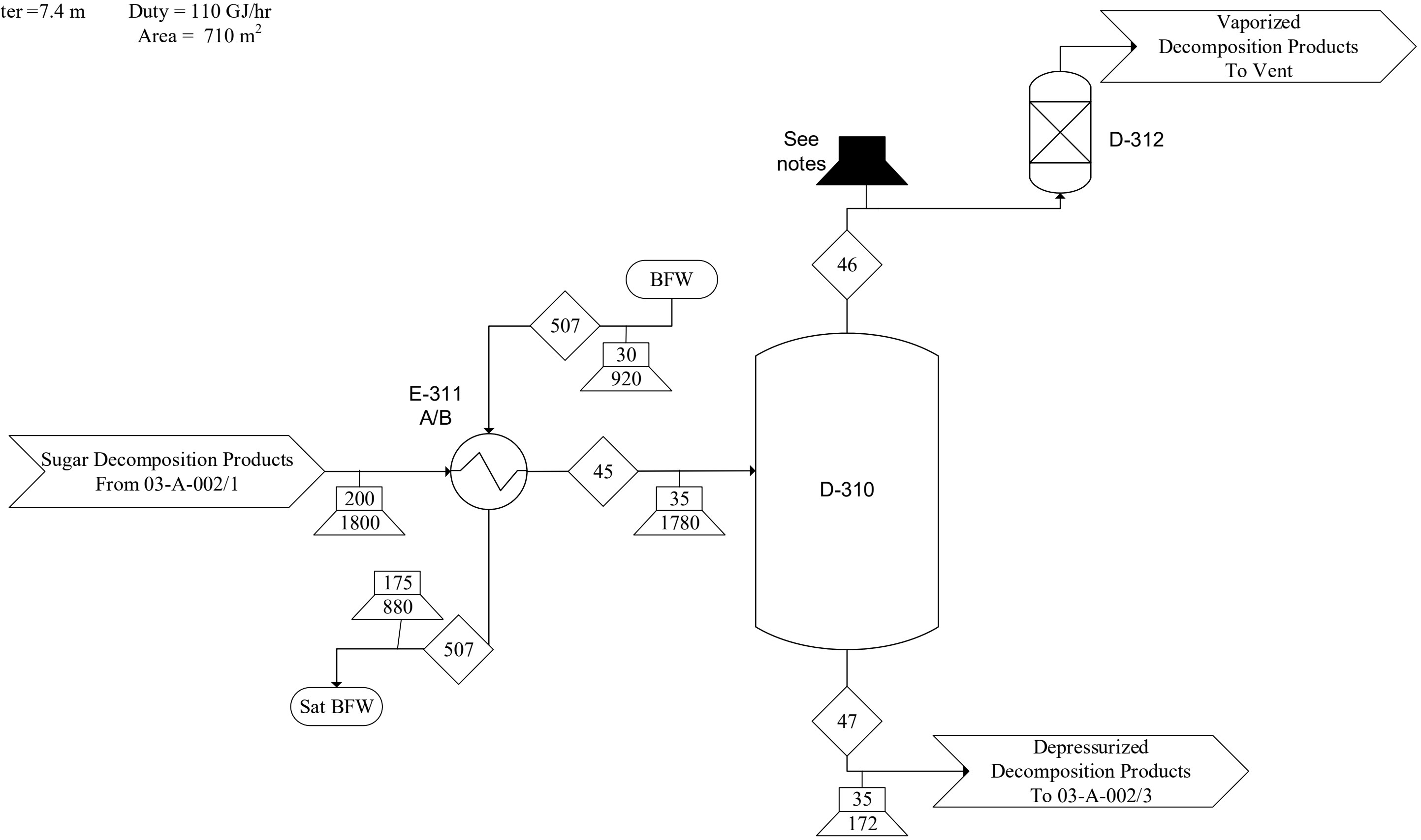
Title
Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG # / Sheet #	Rev	Page
AJK 7/05/20	03-A-002/1	0	427

D-310
Furfural Flash Drum
Height = 3.7 m
Diameter =7.4 m

E-311 A-D/E/F
Pre-Furfural
Extraction Cooler
Duty = 110 GJ/hr
Area = 710 m^2

D-312
Chemical Filter
See Notes



Components/Stream	45	46	47	507
Acetic Acid	1,200	0	1,200	0
Calcium Sulfate	270	0	270	0
Formic Acid	1,200	Trace	1,200	0
Furfural	190	0	190	0
Lactic Acid	9,400	0	9,400	0
Levulinic Acid	390	0	390	0
Water	150,000	Trace	150,000	190,000
Total	**160,000**	**Trace**	**160,000**	**190,000**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Stream 73 normally has no flow. Since flow is dependent on process upsets, temperature and pressure are not specified.

5. A chemical filter is added on stream 73 to remove formic acid, allowing water vapor to vent to the atmosphere. Since the flow of stream 73 is particularily small, D-312 will also be small. Design of D-312 will be specified in the future, more detailed designs.

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
Pressure (kPa), gas
Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001 PFD 04-A-002
PFD 01-A-002 PFD 05-A-002
PFD 02-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	03-A-002/2	0	428

D-320
1st Stage Furfural Extraction Mixing Drum
Height = 7.4 m
Diameter = 3.7 m

D-321
1st Stage Furfural Extraction Settler
Height = 1.6 m
Diameter = 3.2 m

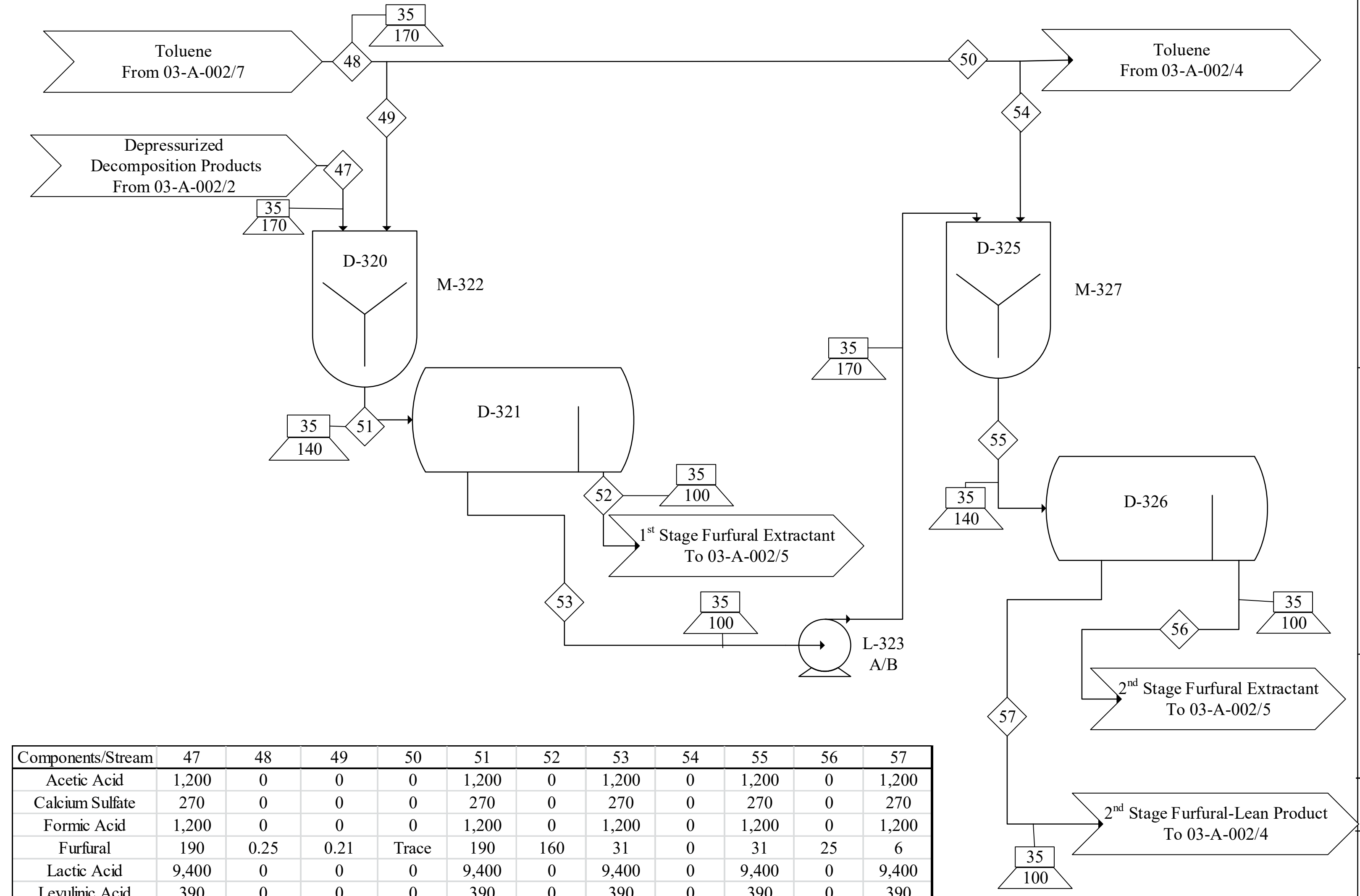
M-322
1st Stage Furfural Mixer
Duty =39 kW

L-323 A/B
1st Stage Furfural Pump
Duty = 5.0 kW

D-325
2nd Stage Furfural Extraction Mixing Drum
Height = 7.4 m
Diameter = 3.7 m

D-326
2nd Stage Furfural Extraction Settler
Height = 1.6 m
Diameter = 3.2 m

M-327
2nd Stage Furfural Mixer
Duty =39 kW



Components/Stream	47	48	49	50	51	52	53	54	55	56	57
Acetic Acid	1,200	0	0	0	1,200	0	1,200	0	1,200	0	1,200
Calcium Sulfate	270	0	0	0	270	0	270	0	270	0	270
Formic Acid	1,200	0	0	0	1,200	0	1,200	0	1,200	0	1,200
Furfural	190	0.25	0.21	Trace	190	160	31	0	31	25	6
Lactic Acid	9,400	0	0	0	9,400	0	9,400	0	9,400	0	9,400
Levulinic Acid	390	0	0	0	390	0	390	0	390	0	390
Toluene	0	330	270	55	270	270	0	43	43	43	0
Water	150,000	0	0	0	150,000	Trace	150,000	0	150,000	Trace	150,000
Total	**150,000**	**330**	**270**	**120**	**150,000**	**430**	**150,000**	**43**	**150,000**	**68**	**150,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇#	Stream Identification
〉	Input/Output
■#	Temperature (°C), gas
☐#	Temperature (°C), liquid
▲#	Pressure (kPa), gas
△#	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	03-A-002/3	0	429

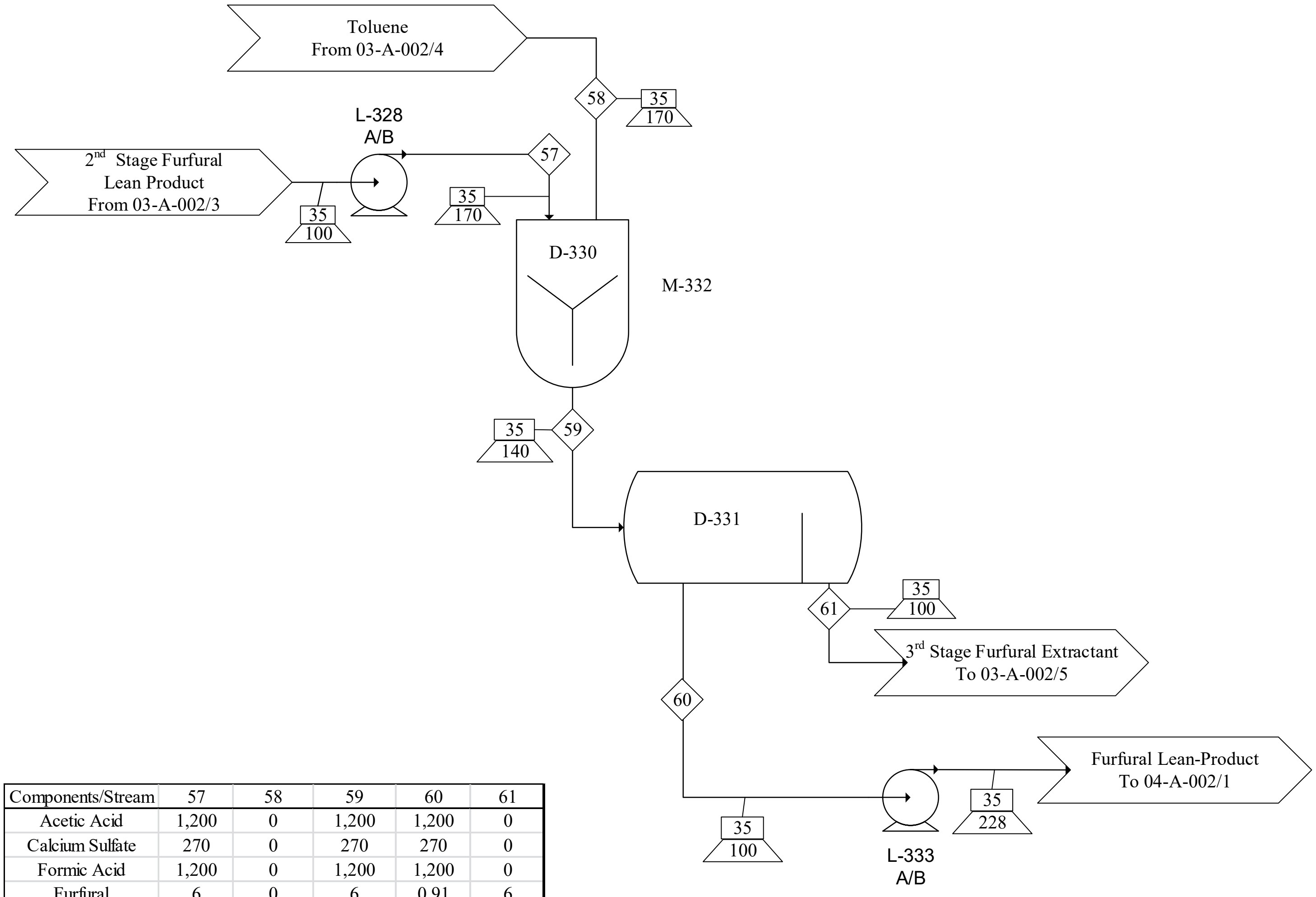
D-330
3rd Stage Furfural
Extraction Mixing Drum
Height = 7.4 m
Diameter = 3.7 m

D-331
3rd Stage Furfural
Extraction Settler
Height = 1.6 m
Diameter = 3.2 m

M-332
3rd Stage Furfural Mixer
Duty = 39 kW

L-328 A/B
2nd Stage Furfural Pump
Duty = 5.0 kW

L-333 A/B
3rd Stage Furfural Pump
Duty = 9.0 kW



Components/Stream	57	58	59	60	61
Acetic Acid	1,200	0	1,200	1,200	0
Calcium Sulfate	270	0	270	270	0
Formic Acid	1,200	0	1,200	1,200	0
Furfural	6	0	6	0.91	6
Lactic Acid	9,400	0	9,400	9,400	0
Levulinic Acid	390	0	390	390	0
Toluene	0	12	12	0	12
Water	150,000	0	150,000	150,000	0
Total	**150,000**	**12**	**150,000**	**150,000**	**18**

430

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ # Stream Identification
Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
▲ # Pressure (kPa), gas
△ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

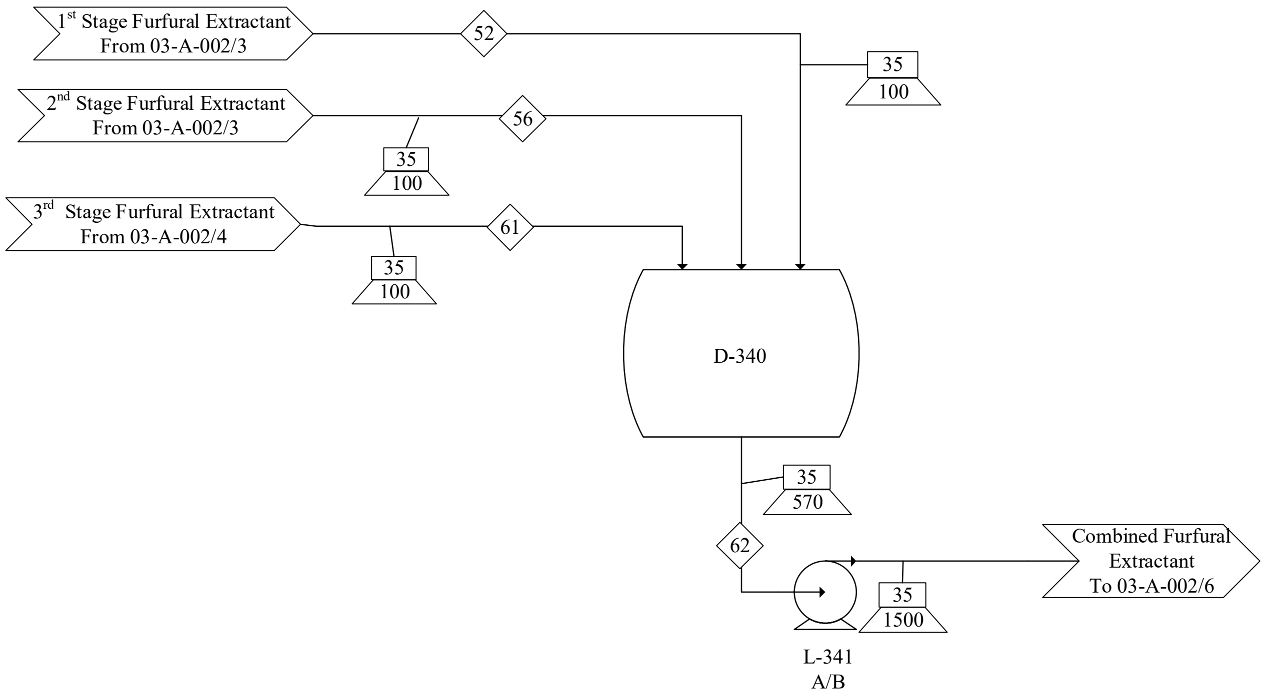
Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG # Sheet #	Rev	Page
AJK 7/05/20	03-A-002/4	0	430

D-340
Furfural Extractant
Holdup Drum
Height = 0.79 m
Diameter = 0.39 m

L-341 A/B
Furfural Extractant
Pump
Duty = 0.74 kW



Components/Stream	52	56	61	62
Furfural	160	25	6	190
Toluene	270	43	18	330
Total	**430**	**68**	**18**	**510**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #

▷

■ #

□ #

▲ #

△ #

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	
AJK 7/05/20	03-A-002/5	0	431

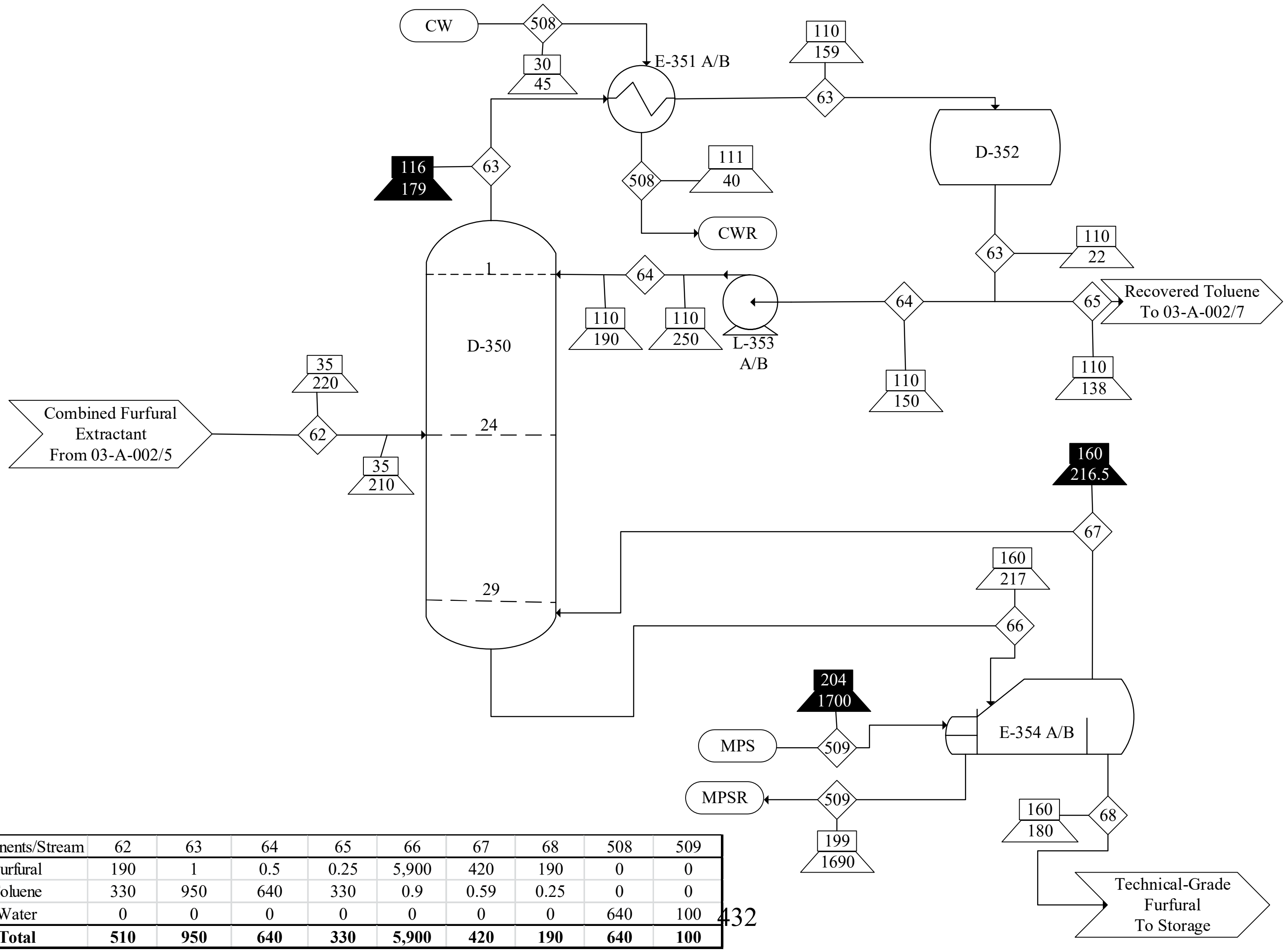
D-350
Toluene Recovery Column
Diameter: 0.23 m
Height: 12 m

E-351 A/B
Toluene Recovery Condenser
Duty: 0.20 GJ/hr
Area: 3.9 m^2

D-352
Toluene Recovery Reflux Drum
Height: 0.28 m
Length: 0.85 m

L-353 A/B
Toluene Recovery Reflux Pump
Duty: 0.74 kW

E-354 A/B
Toluene Recovery Reboiler
Duty: 0.30 GJ/hr
Area: 2.4 m^2



Components/Stream	62	63	64	65	66	67	68	508	509
Furfural	190	1	0.5	0.25	5,900	420	190	0	0
Toluene	330	950	640	330	0.9	0.59	0.25	0	0
Water	0	0	0	0	0	0	0	640	100
Total	**510**	**950**	**640**	**330**	**5,900**	**420**	**190**	**640**	**100**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Trays have and efficiency of 70%
6. Pressures are different in streams 91 and 92 due to control valve on stream 91 being after the pump.

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
□# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

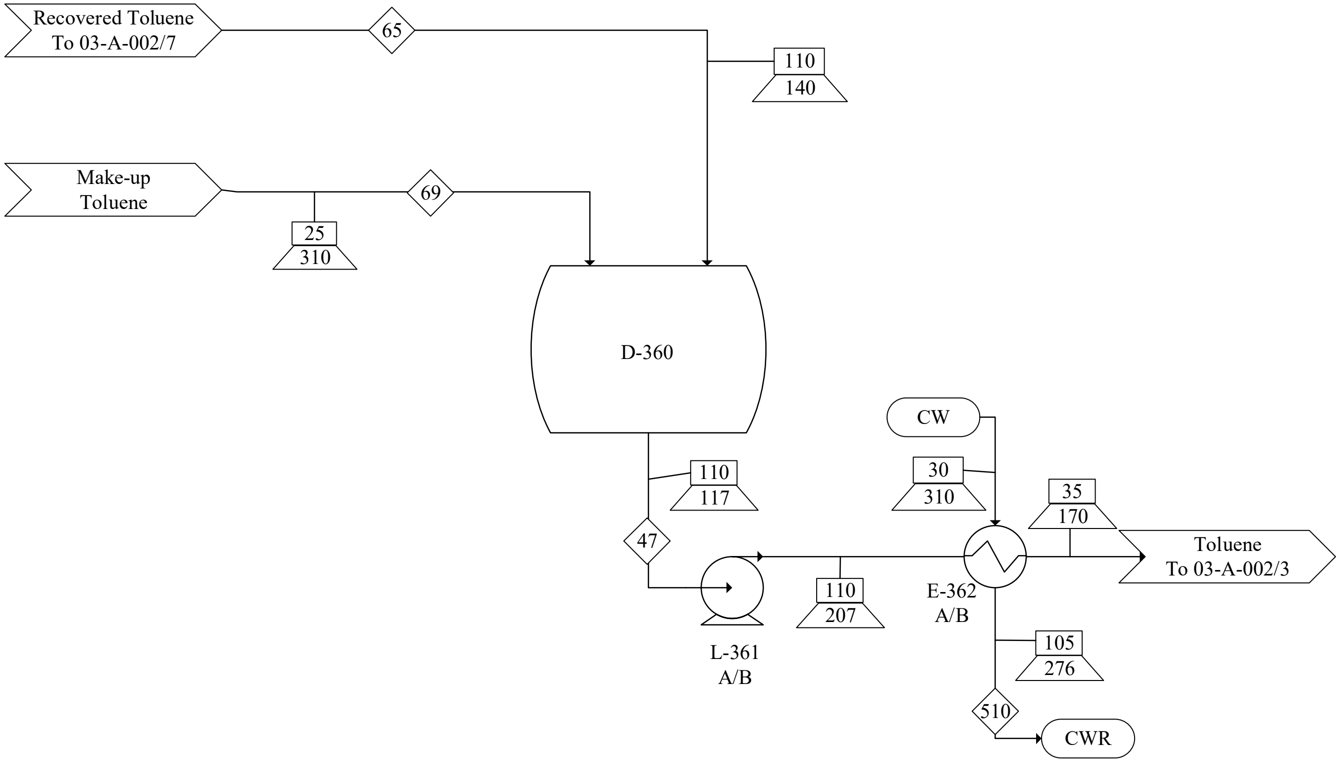
Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/6	0	432

D-360
Toluene Recycle Drum
Height = 0.94 m
Diameter = 0.30 m

L-361 A/B
Toluene Pump
Duty = 0.74 kW

E-362 A/B
Toluene Cooler
Duty = 0.044 MJ/hr
Area = 9.3 m^2



Components/Stream	47	65	69	510
Furfural	0.25	0.25	0	0
Toluene	330	330	Trace	0
Water	0	0	0	140
Total	**330**	**330**	**Trace**	**140**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Make-up Toluene flow is normally zero.

Special Symbols / Nomenclature

◇ # Stream Identification
▷ Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
▲ # Pressure (kPa), gas
△ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG # / Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/7	0	433

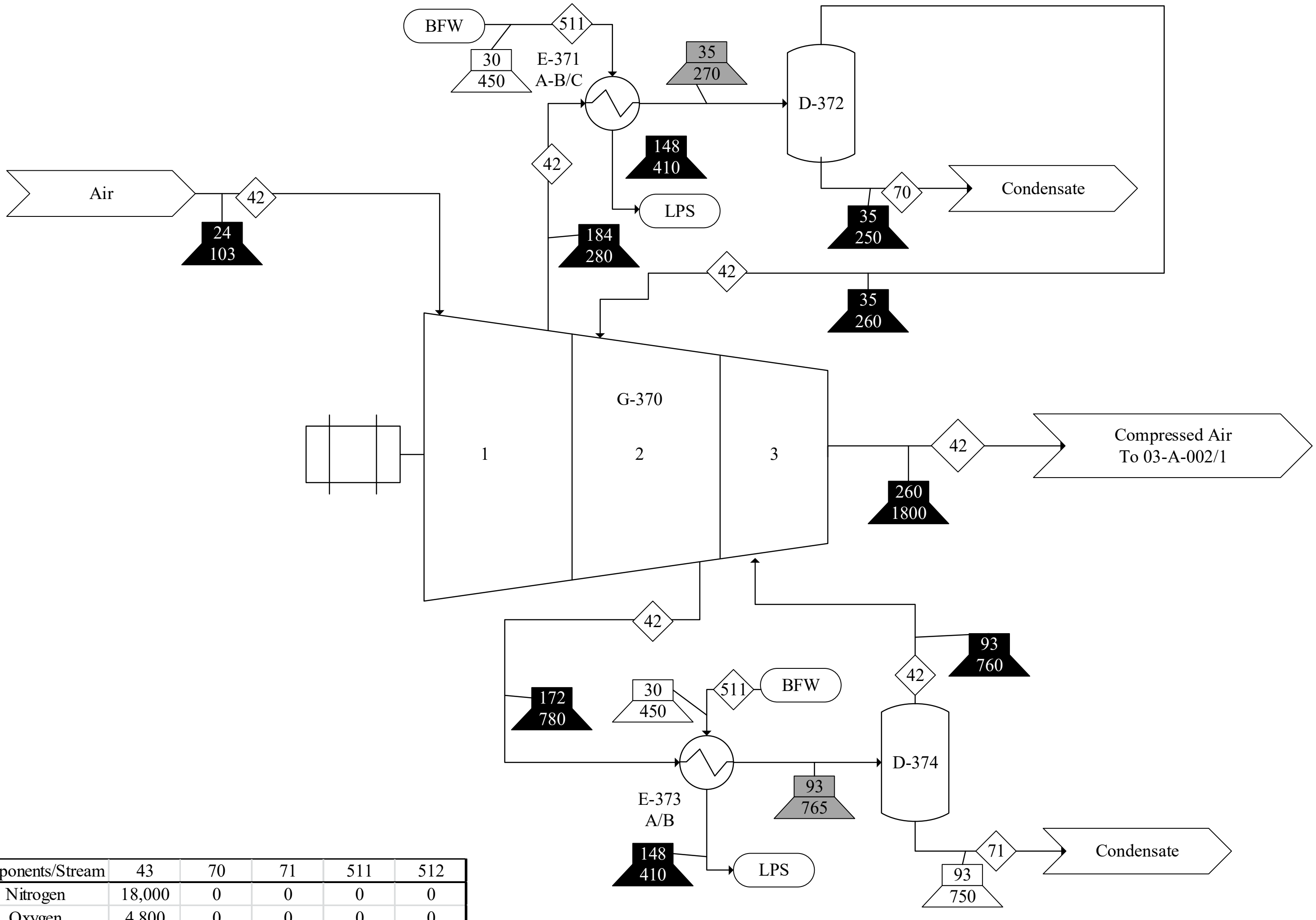
G-370
Three Stage Air Compressor
1st Stage Duty = 880 kW
2nd Stage Duty = 880 kW
3rd Stage Duty = 880 kW

E-371 A-B/C
First Inter-stage cooler
Duty = 3.3 GJ/hr
Area = 370 m^2

D-372
First Inter-stage Knock-out Drum
Diameter = 2.7 m
Height = 5.3 m

E-373 A/B
Second Inter-stage cooler
Duty = 1.9 GJ/hr
Area = 31 m^2

D-374
Second Inter-stage Knock-out Drum
Diameter = 2.0 m
Height =4.0 m



Components/Stream	43	70	71	511	512
Nitrogen	18,000	0	0	0	0
Oxygen	4,800	0	0	0	0
Water	230	Trace	Trace	1,200	720
Total	**23,000**	**Trace**	**Trace**	**1,200**	**720**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Flows of stream 162 and 163 are normally zero.
5. Air enters at an average humidity of 70%.

Special Symbols / Nomenclature

◇# Stream Identification
▷ Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
▨# Temperature (°C), mix phase
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid
▲# Pressure (kPa), mix phase

Reference Drawings:

I/O 00-A-001
BFD 03-A-001
PFD 01-A-002
PFD 02-A-002
PFD 04-A-002
PFD 05-A-002

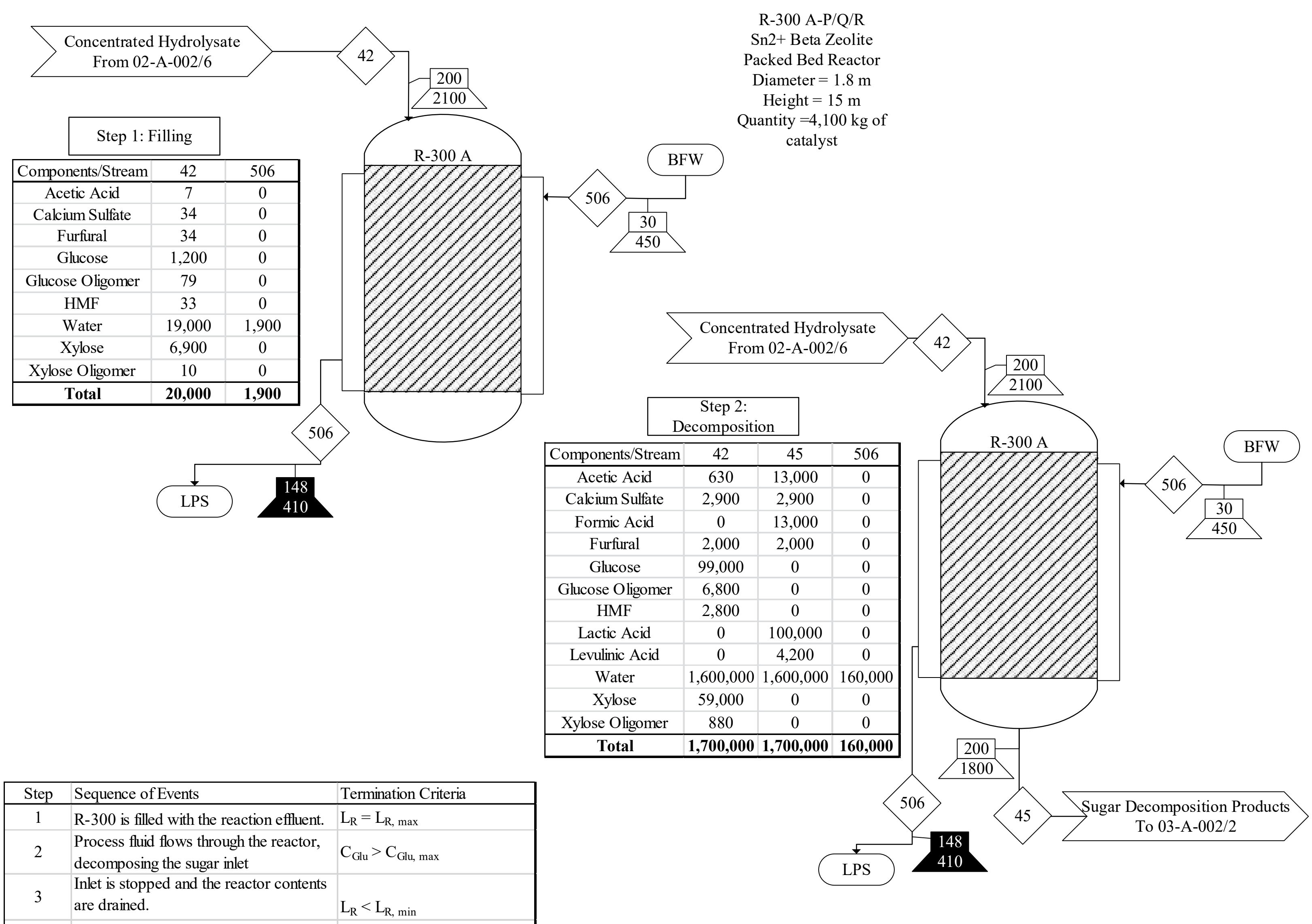
Drawing Type

Process Flow Diagram

Title

Process Area 03: Catalytic Decomposition and Furfural Extraction

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	03-A-002/8	0	434



R-300 A-P/Q/R
Sn2+ Beta Zeolite
Packed Bed Reactor
Diameter = 1.8 m
Height = 15 m
Quantity =4,100 kg of
catalyst

Step 1: Filling

Components/Stream	42	506
Acetic Acid	7	0
Calcium Sulfate	34	0
Furfural	34	0
Glucose	1,200	0
Glucose Oligomer	79	0
HMF	33	0
Water	19,000	1,900
Xylose	6,900	0
Xylose Oligomer	10	0
Total	**20,000**	**1,900**

Step 2:
Decomposition

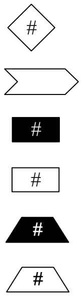
Components/Stream	42	45	506
Acetic Acid	630	13,000	0
Calcium Sulfate	2,900	2,900	0
Formic Acid	0	13,000	0
Furfural	2,000	2,000	0
Glucose	99,000	0	0
Glucose Oligomer	6,800	0	0
HMF	2,800	0	0
Lactic Acid	0	100,000	0
Levulinic Acid	0	4,200	0
Water	1,600,000	1,600,000	160,000
Xylose	59,000	0	0
Xylose Oligomer	880	0	0
Total	**1,700,000**	**1,700,000**	**160,000**

Step	Sequence of Events	Termination Criteria
1	R-300 is filled with the reaction effluent.	$L_R = L_{R,\ max}$
2	Process fluid flows through the reactor, decomposing the sugar inlet	$C_{Glu} > C_{Glu,\ max}$
3	Inlet is stopped and the reactor contents are drained.	$L_R < L_{R,\ min}$
4	Compressed air is run through the catalyst, removing the coke (in the form of Carbon Dioxide) produced by decomposition of the sugars.	$C_{CO2} < C_{CO2,\ min}$
5	R-300 is on standby	$C_{Glu} > C_{Glu,\ max}$ of previous reactor

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 2 hours
5. Step 2 is 172 hours
6. Eighteen reactors are used in parallel for continual decomposition.
7. At any time, sixteen reactors are is use and two are being calcined.

Special Symbols / Nomenclature



Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

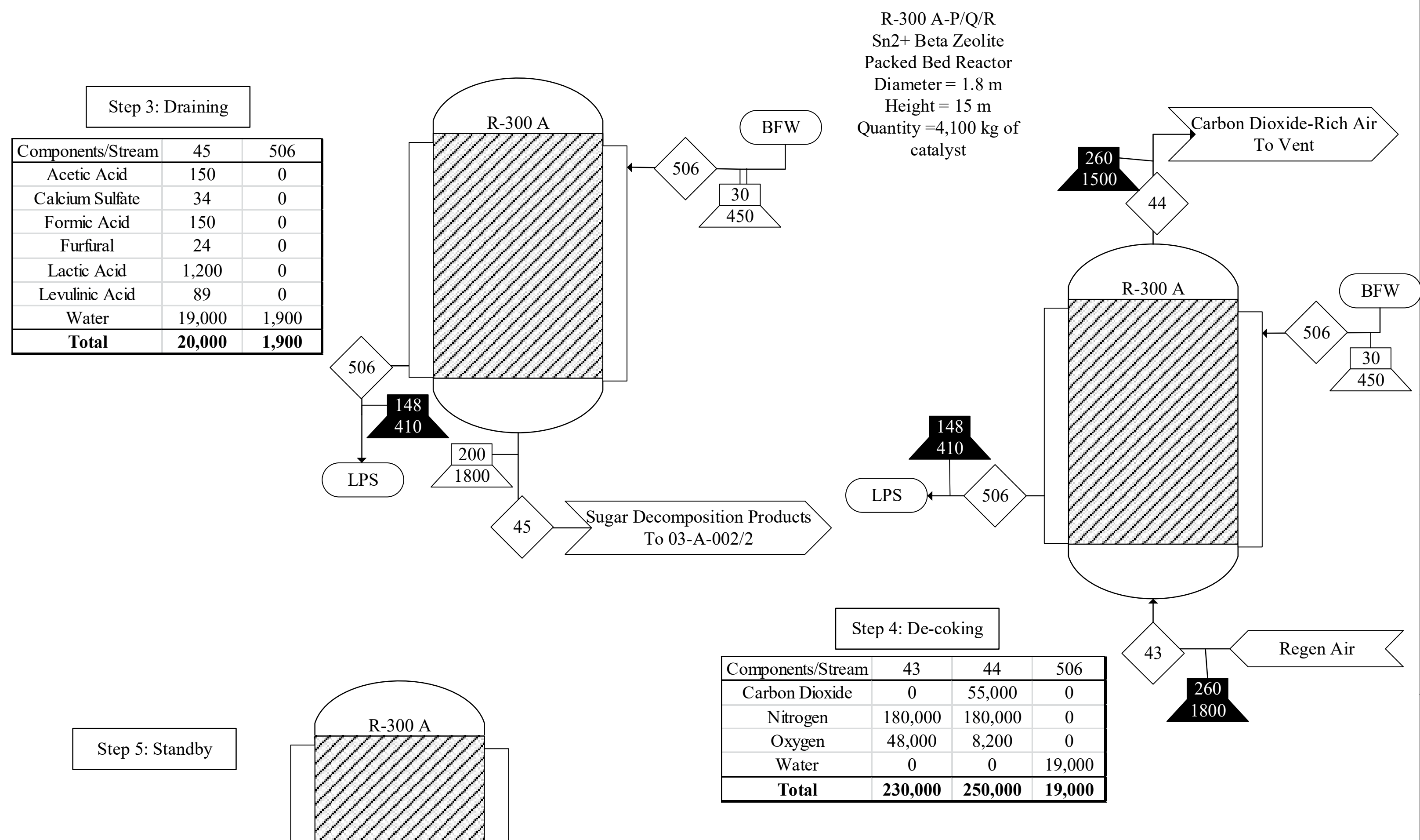
Supplemental Process Flow Diagram

Title

Catalytic Reactor Semi-Batch Steps

By/Date	DWG #/ Sheet #	Rev	
AJK 7/05/20	03-A-003/1	0	435

R-300 A-P/Q/R
Sn2+ Beta Zeolite
Packed Bed Reactor
Diameter = 1.8 m
Height = 15 m
Quantity =4,100 kg of catalyst



Step 3: Draining

Components/Stream	45	506
Acetic Acid	150	0
Calcium Sulfate	34	0
Formic Acid	150	0
Furfural	24	0
Lactic Acid	1,200	0
Levulinic Acid	89	0
Water	19,000	1,900
Total	**20,000**	**1,900**

Step 4: De-coking

Components/Stream	43	44	506
Carbon Dioxide	0	55,000	0
Nitrogen	180,000	180,000	0
Oxygen	48,000	8,200	0
Water	0	0	19,000
Total	**230,000**	**250,000**	**19,000**

Step 5: Standby

Step	Sequence of Events	Termination Criteria
1	R-300 is filled with the reaction effluent.	$L_R = L_{R,\ max}$
2	Process fluid flows through the reactor, decomposing the sugar inlet	$C_{Glu} > C_{Glu,\ max}$
3	Inlet is stopped and the reactor contents are drained.	$L_R < L_{R,\ min}$
4	Compressed air is run through the catalyst, removing the coke (in the form of Carbon Dioxide) produced by decomposition of the sugars.	$C_{CO2} < C_{CO2,\ min}$
5	R-300 is on standby	$C_{Glu} > C_{Glu,\ max}$ of previous reactor

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 3 is 2 hours
5. Step 4 is 20 hours
6. Step 5 is 4 hour
7. Eighteen reactors are used in parallel for continual decomposition.
8. At any time, sixteen reactors are is use and two are being calcined.

Special Symbols / Nomenclature

◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°C), gas
□ #	Temperature (°C), liquid
▲ #	Pressure (psia), gas
△ #	Pressure (psia), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

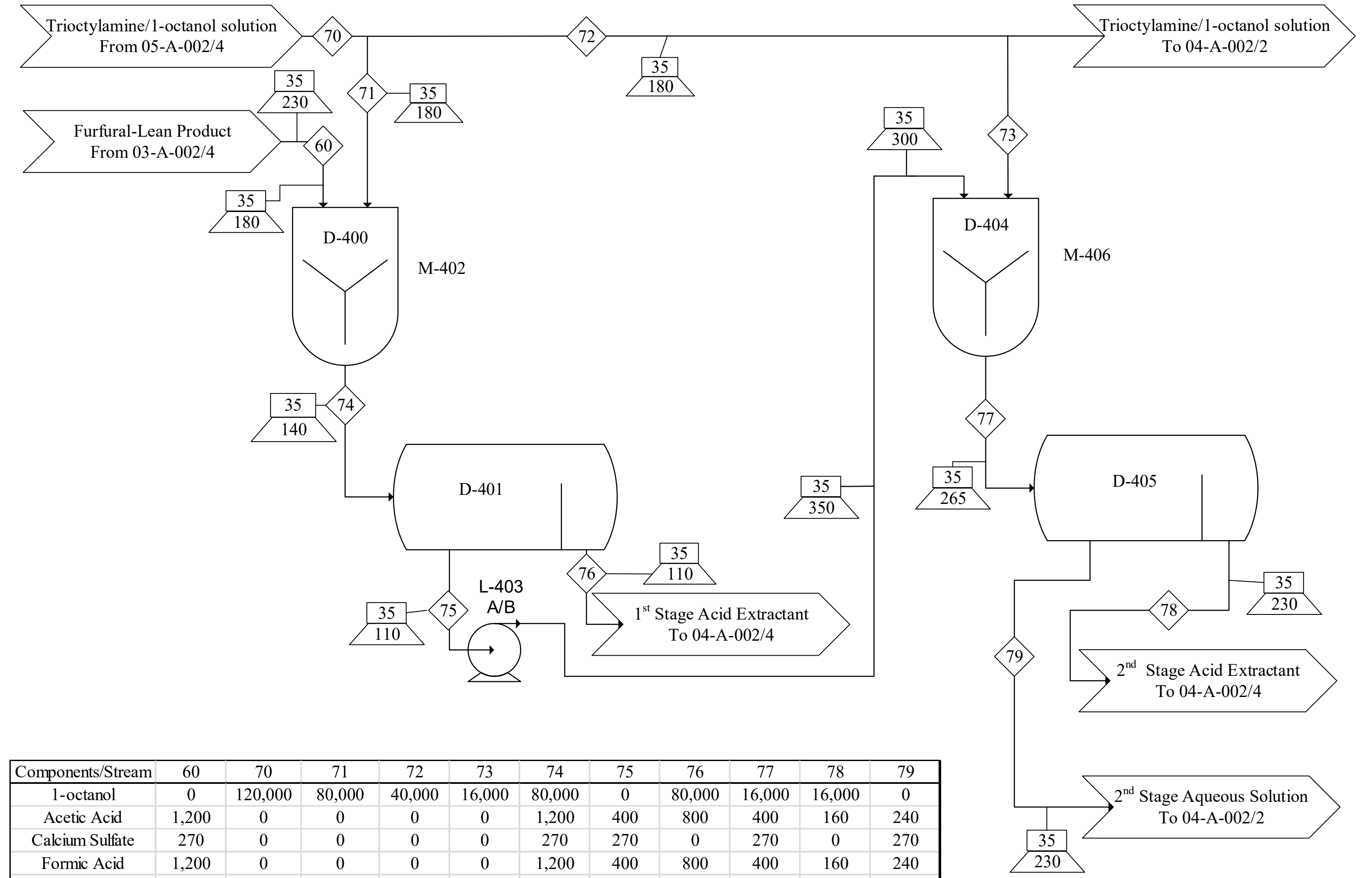
Supplemental Process Flow Diagram

Title

Catalytic Reactor Semi-Batch Steps

By/Date	DWG #/ Sheet #	Rev	Page
AJK 7/05/20	03-A-003/2	0	436

D-400	D-401	M-402	L-403 A/B	D-404	D-405	M-406
1st Stage TOA Mixing Drum	1st Stage TOA Settler	1st Stage TOA Mixer	1st Stage TOA Pump	2nd Stage TOA Mixing Drum	2nd Stage TOA Settler	2nd Stage TOA Mixer
Height = 8.0 m Diameter = 3.9 m	Height = 2.4 m Diameter = 4.8 m	Power = 54 kW	Duty = 8.6 kW	Height = 8.0 m Diameter = 3.9 m	Height = 2.4 m Diameter = 4.8 m	Power = 54 kW



Components/Stream	60	70	71	72	73	74	75	76	77	78	79
1-octanol	0	120,000	80,000	40,000	16,000	80,000	0	80,000	16,000	16,000	0
Acetic Acid	1,200	0	0	0	0	1,200	400	800	400	160	240
Calcium Sulfate	270	0	0	0	0	270	270	0	270	0	270
Formic Acid	1,200	0	0	0	0	1,200	400	800	400	160	240
Lactic Acid	9,400	0	0	0	0	9,400	3,100	6,300	3,100	1,200	1,900
Levulinic Acid	390	0	0	0	0	390	130	260	130	52	78
Trioctylamine	0	37,000	25,000	12,000	4,800	37,000	0	37,000	4,800	4,800	0
Water	150,000	0	0	0	0	150,000	150,000	0	150,000	0	150,000
Total	**160,000**	**160,000**	**110,000**	**52,000**	**21,000**	**320,000**	**150,000**	**120,000**	**250,000**	**22,000**	**150,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #	Stream Identification
⇨	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type

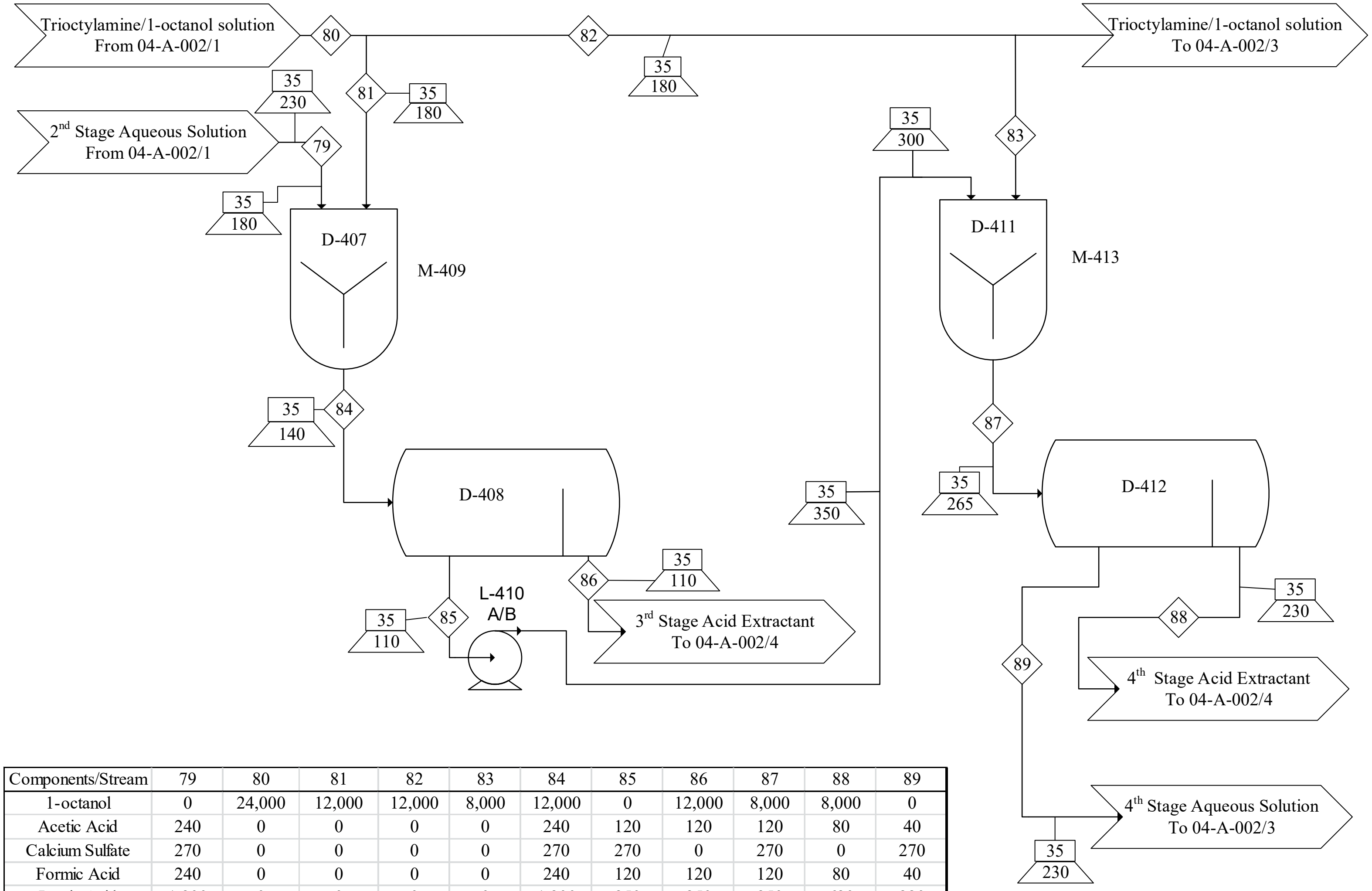
Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/1	0	437

D-407	D-408	M-409	L-410 A/B	D-411	D-412	M-413
3rd Stage TOA Mixing Drum	3rd Stage TOA Settler	3rd Stage TOA Mixer	3rd Stage TOA Pump	4th Stage TOA Mixing Drum	4th Stage TOA Settler	4th Stage TOA Mixer
Height = 8.0 m	Height = 2.4 m	Power = 54 kW	Duty = 8.6 kW	Height = 8.0 m	Height = 2.4 m	Power = 54 kW
Diameter = 3.9 m	Diameter = 4.8 m			Diameter = 3.9 m	Diameter = 4.8 m	



Components/Stream	79	80	81	82	83	84	85	86	87	88	89
1-octanol	0	24,000	12,000	12,000	8,000	12,000	0	12,000	8,000	8,000	0
Acetic Acid	240	0	0	0	0	240	120	120	120	80	40
Calcium Sulfate	270	0	0	0	0	270	270	0	270	0	270
Formic Acid	240	0	0	0	0	240	120	120	120	80	40
Lactic Acid	1,900	0	0	0	0	1,900	950	950	950	630	320
Levulinic Acid	78	0	0	0	0	78	39	39	39	26	13
Trioctylamine	0	7,200	3,600	3,600	2,400	3,600	0	3,600	2,400	2,400	0
Water	150,000	0	0	0	0	150,000	150,000	0	150,000	0	150,000
Total	**150,000**	**31,000**	**16,000**	**16,000**	**10,000**	**320,000**	**150,000**	**18,000**	**250,000**	**10,000**	**150,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇# Stream Identification
Input/Output
■# Temperature (°C), gas
☐# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001 PFD 03-A-002
PFD 01-A-002 PFD 05-A-002
PFD 02-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

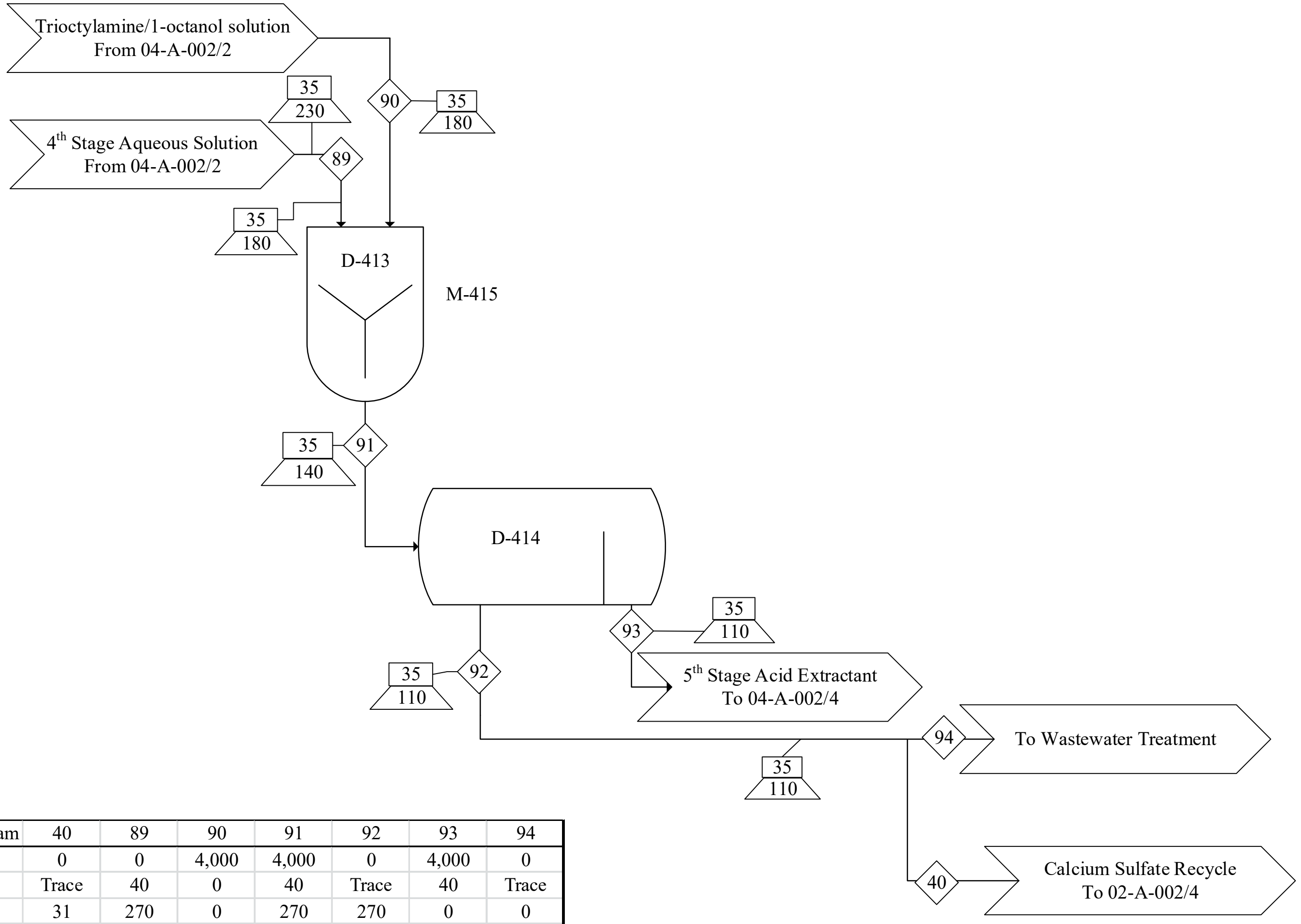
By/Date	DWG # Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/2	0	438

D-413
1st Stage TOA Mixing Drum
Height = 8.0 m
Diameter = 3.9 m

D-414
1st Stage TOA Settler
Height = 2.4 m
Diameter = 4.8 m

M-415
1st Stage TOA Mixer
Power = 54 kW

L-416 A/B
1st Stage TOA Pump
Duty = 13 kW



Components/Stream	40	89	90	91	92	93	94
1-octanol	0	0	4,000	4,000	0	4,000	0
Acetic Acid	Trace	40	0	40	Trace	40	Trace
Calcium Sulfate	31	270	0	270	270	0	0
Formic Acid	0	40	0	40	0	40	0
Lactic Acid	1.5	320	0	320	10	310	9
Levulinic Acid	0	13	0	13	0	12	0
Trioctylamine	0	0	1,200	1,200	0	1,200	0
Water	23,000	150,000	0	0	150,000	0	130,000
Total	**23,000**	**150,000**	**5,200**	**150,000**	**10,000**	**5,200**	**130,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #
〉
■ #
☐ #
▲ #
△ #

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

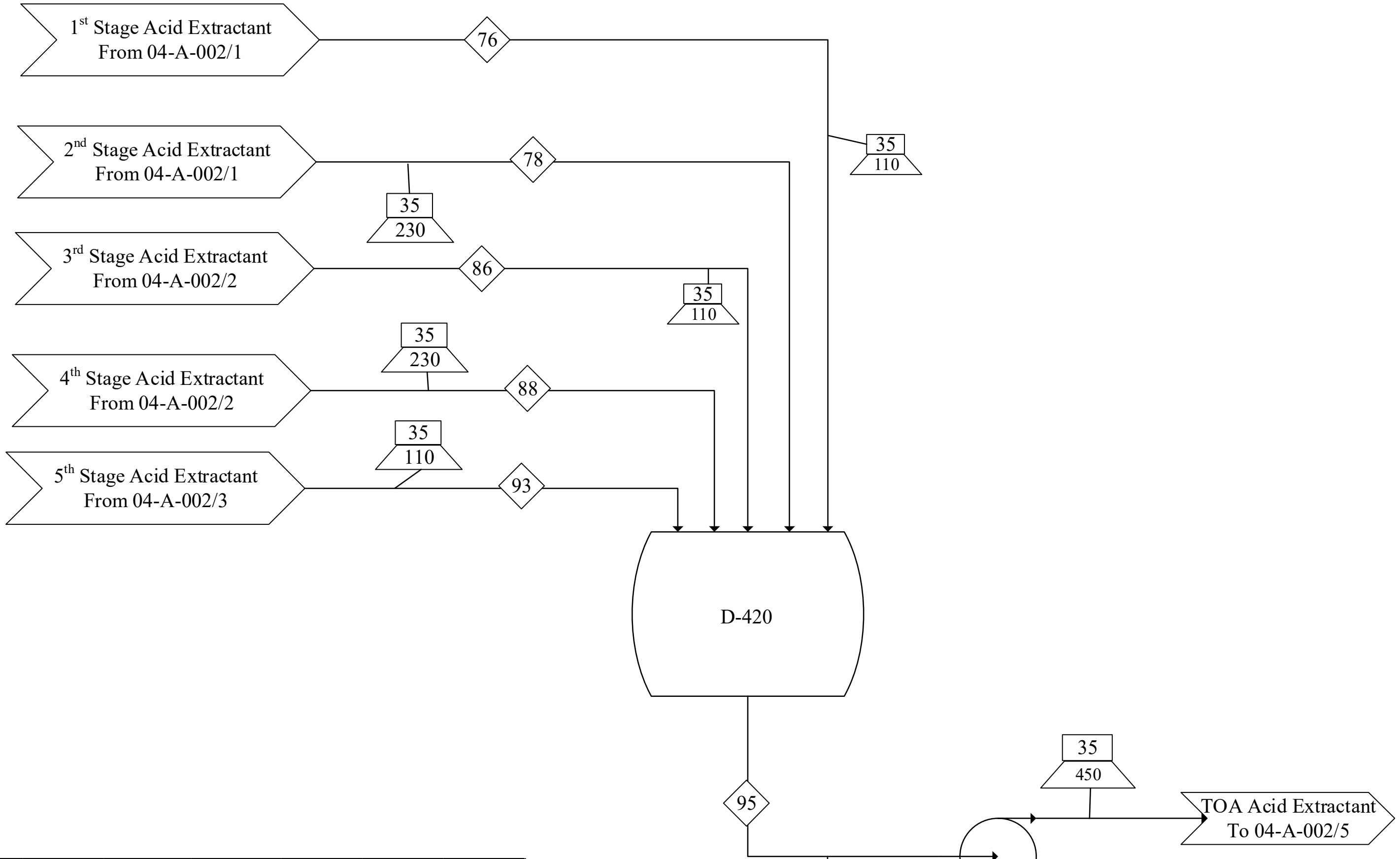
Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/3	0	439

D-420
TOA Acid Extractant
Hold Up Drum
Height = 4.4 m
Diameter = 2.2 m

L-421 A/B
TOA Hold Up Drum
Pump
Duty = 18 kW



Components/Stream	76	78	86	88	93	95
1-octanol	80,000	16,000	12,000	8,000	4,000	120,000
Acetic Acid	800	160	120	80	40	1,200
Formic Acid	800	160	120	80	40	1,200
Lactic Acid	6,300	1,200	950	630	310	9,400
Levulinic Acid	260	52	39	26	12	390
Trioctylamine	37,000	4,800	3,600	2,400	1,200	49,000
Total	**120,000**	**22,000**	**18,000**	**10,000**	**5,200**	**180,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/4	0	440

D-430
Volatile Acid Recovery Column
Rectifying Diameter: 0.61 m
Stripping Diameter: 4.8 m
Height: 14 m

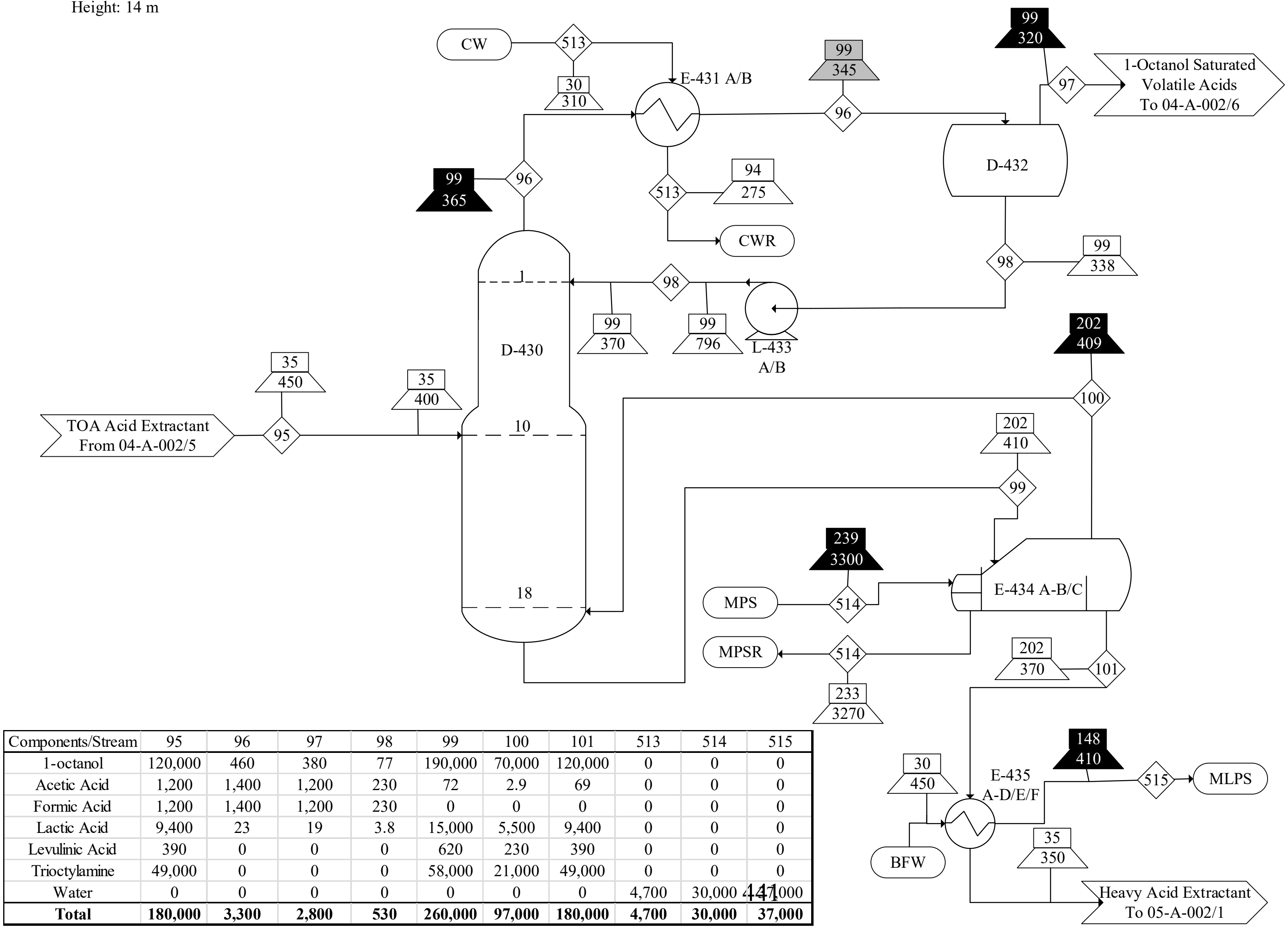
E-431 A/B
Volatile Acid Column Condenser
Duty: 1.7 GJ/hr
Area: 21 m^2

D-432
Volatile Acid Column Reflux Drum
Height: 1.5 m
Length: 0.76 m

L-433 A/B
Volatile Acid Column Reflux Pump
Duty: 0.74 kW

E-434 A-B/B
Volatile Acid Column Reboiler
Duty: 85 GJ/hr
Area: 900 m^2

E-435 A-D/E/F
Heavy Acid Extractant Cooler
Duty: 99 GJ/hr
Area: 840 m^2



Components/Stream	95	96	97	98	99	100	101	513	514	515
1-octanol	120,000	460	380	77	190,000	70,000	120,000	0	0	0
Acetic Acid	1,200	1,400	1,200	230	72	2.9	69	0	0	0
Formic Acid	1,200	1,400	1,200	230	0	0	0	0	0	0
Lactic Acid	9,400	23	19	3.8	15,000	5,500	9,400	0	0	0
Levulinic Acid	390	0	0	0	620	230	390	0	0	0
Trioctylamine	49,000	0	0	0	58,000	21,000	49,000	0	0	0
Water	0	0	0	0	0	0	0	4,700	30,000	37,000
Total	**180,000**	**3,300**	**2,800**	**530**	**260,000**	**97,000**	**180,000**	**4,700**	**30,000**	**37,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Trays have and efficiency of 70%

Special Symbols / Nomenclature

Stream Identification
Input/Output
Temperature (°F), gas
Temperature (°F), liquid
Pressure (kPa), gas
Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

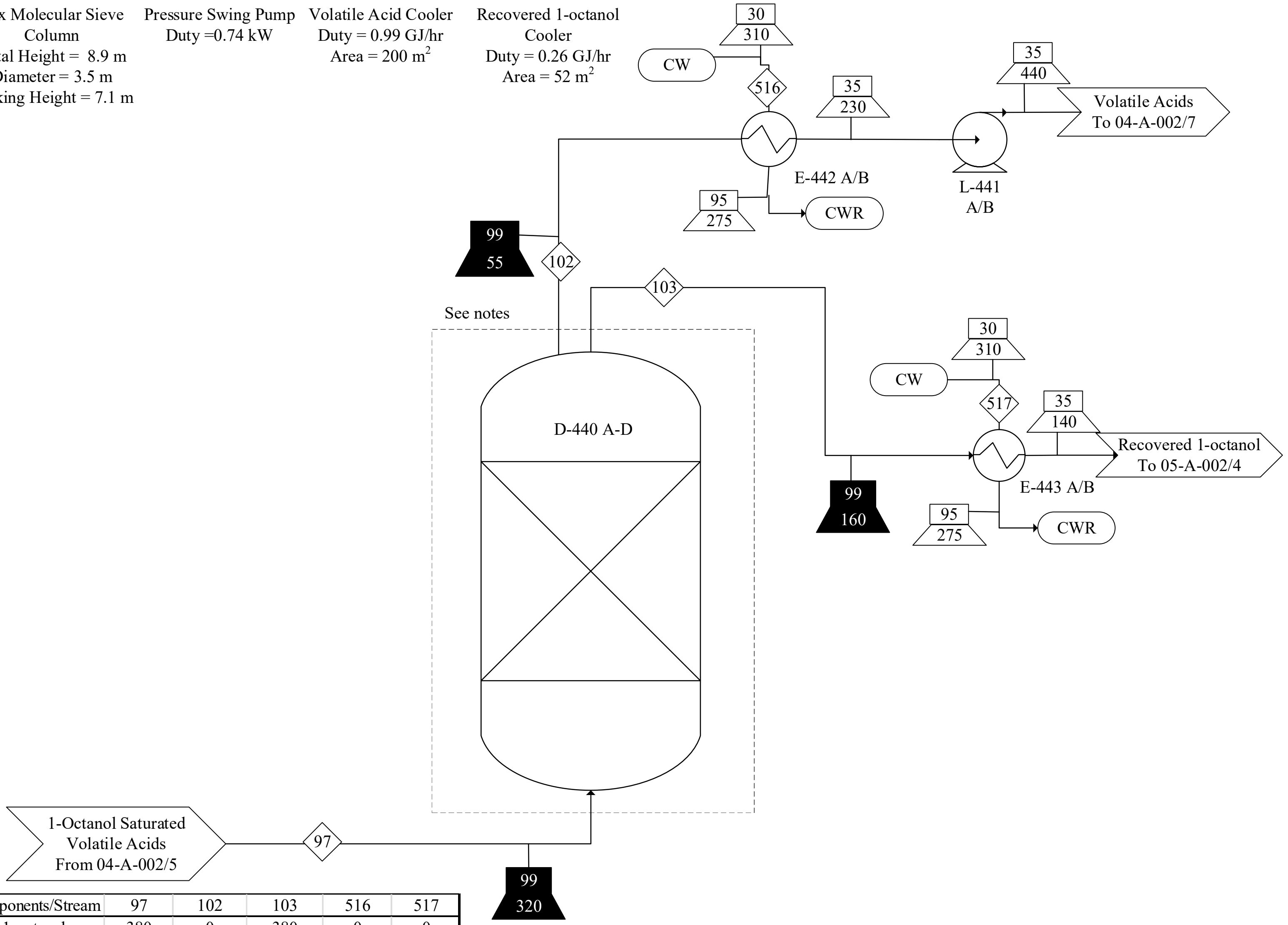
By/Date	DWG # / Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/5	0	441

D-440 A-D
13x Molecular Sieve Column
Total Height = 8.9 m
Diameter = 3.5 m
Packing Height = 7.1 m

L-441
Pressure Swing Pump
Duty =0.74 kW

E-442 A/B
Volatile Acid Cooler
Duty = 0.99 GJ/hr
Area = 200 m^2

E-443 A/B
Recovered 1-octanol Cooler
Duty = 0.26 GJ/hr
Area = 52 m^2



Components/Stream	97	102	103	516	517
1-octanol	380	0	380	0	0
Acetic Acid	1,200	1,200	0	0	0
Formic Acid	1,200	1,200	0	0	0
Lactic Acid	19	19	0	0	0
Water	0	0	0	2,800	950
Total	**2,800**	**2,400**	**380**	**2,800**	**950**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Compressors have an efficiency of 65%
5. D-440 is a semi-batch process, more information can be seen on sheet 04-A-003

Special Symbols / Nomenclature

◇ #	Stream Identification
⌦	Input/Output
■ #	Temperature (°F), gas
☐ #	Temperature (°F), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001	PFD 03-A-002
BFD 04-A-001	PFD 05-A-002
PFD 01-A-002	Supplemental
PFD 02-A-002	PFD 04-A-003

Drawing Type

Process Flow Diagram

Title

Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/6	0	442

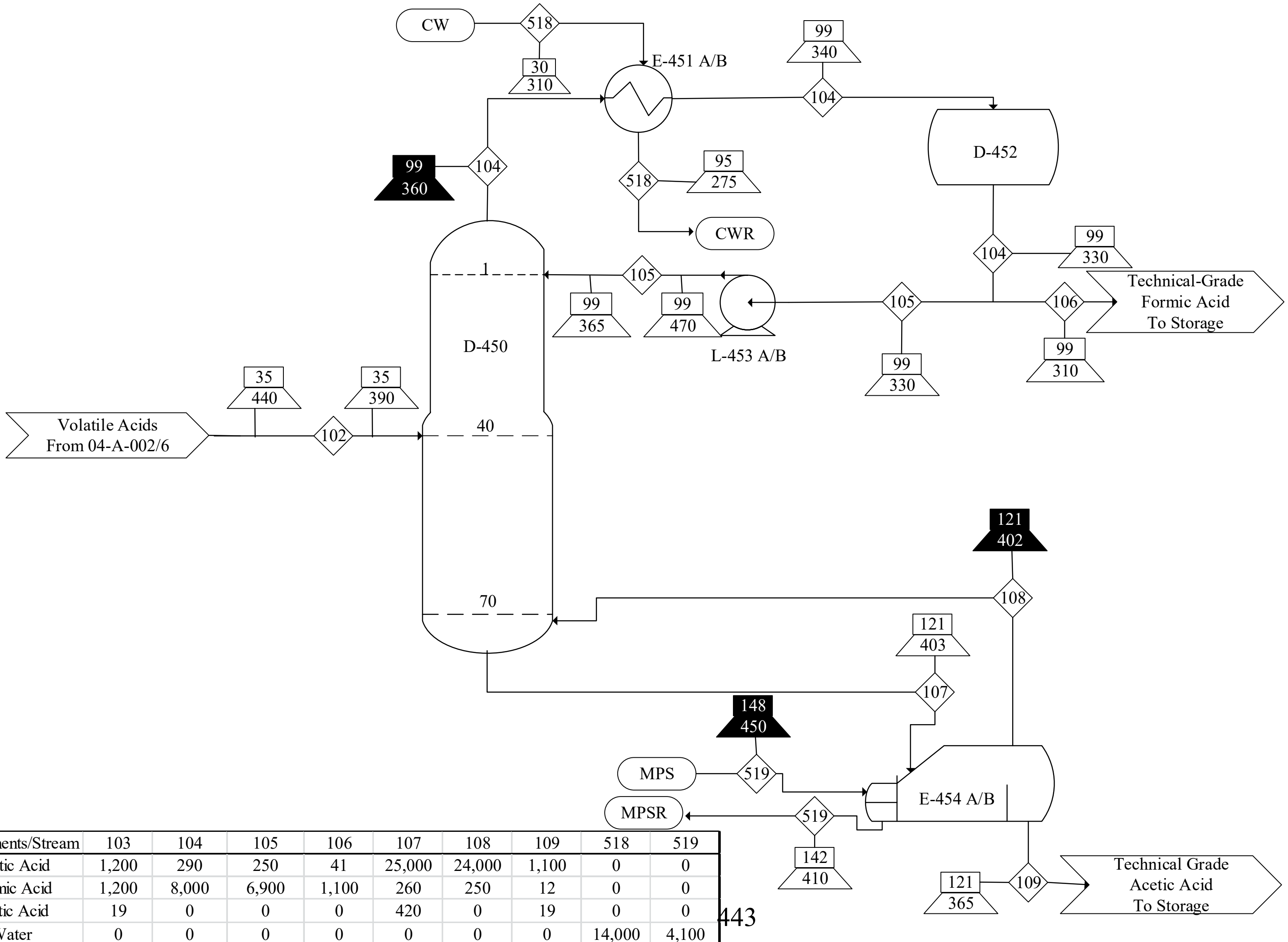
D-450
Formic/Acetic Column
Rectifying Diameter: 0.97 m
Stripping Diameter: 1.6 m
Height: 43 m

E-451
Formic/Acetic Column
Condenser
Duty: 3.9 GJ/hr
Area: 61 m^2

D-452
Formic/Acetic Column
Reflux Drum
Diameter: 0.95 m
Length: 1.9 m

L-453
Formic/Acetic Column
Reflux Pump
Duty: 0.74 kW

E-454
Formic/Acetic Column
Reboiler
Duty: 11 GJ/hr
Area: 86 m^2



Components/Stream	103	104	105	106	107	108	109	518	519
Acetic Acid	1,200	290	250	41	25,000	24,000	1,100	0	0
Formic Acid	1,200	8,000	6,900	1,100	260	250	12	0	0
Lactic Acid	19	0	0	0	420	0	19	0	0
Water	0	0	0	0	0	0	0	14,000	4,100
Total	**2,800**	**8,300**	**7,200**	**1,100**	**26,000**	**25,000**	**1,100**	**14,000**	**4,100**

443

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Pumps have an efficiency of 65%

5. Trays have and efficiency of 70%

Special Symbols / Nomenclature

#	Stream Identification
〉	Input/Output
#	Temperature (°C), gas
#	Temperature (°C), liquid
#	Pressure (kPa), gas
#	Pressure (kPa), liquid

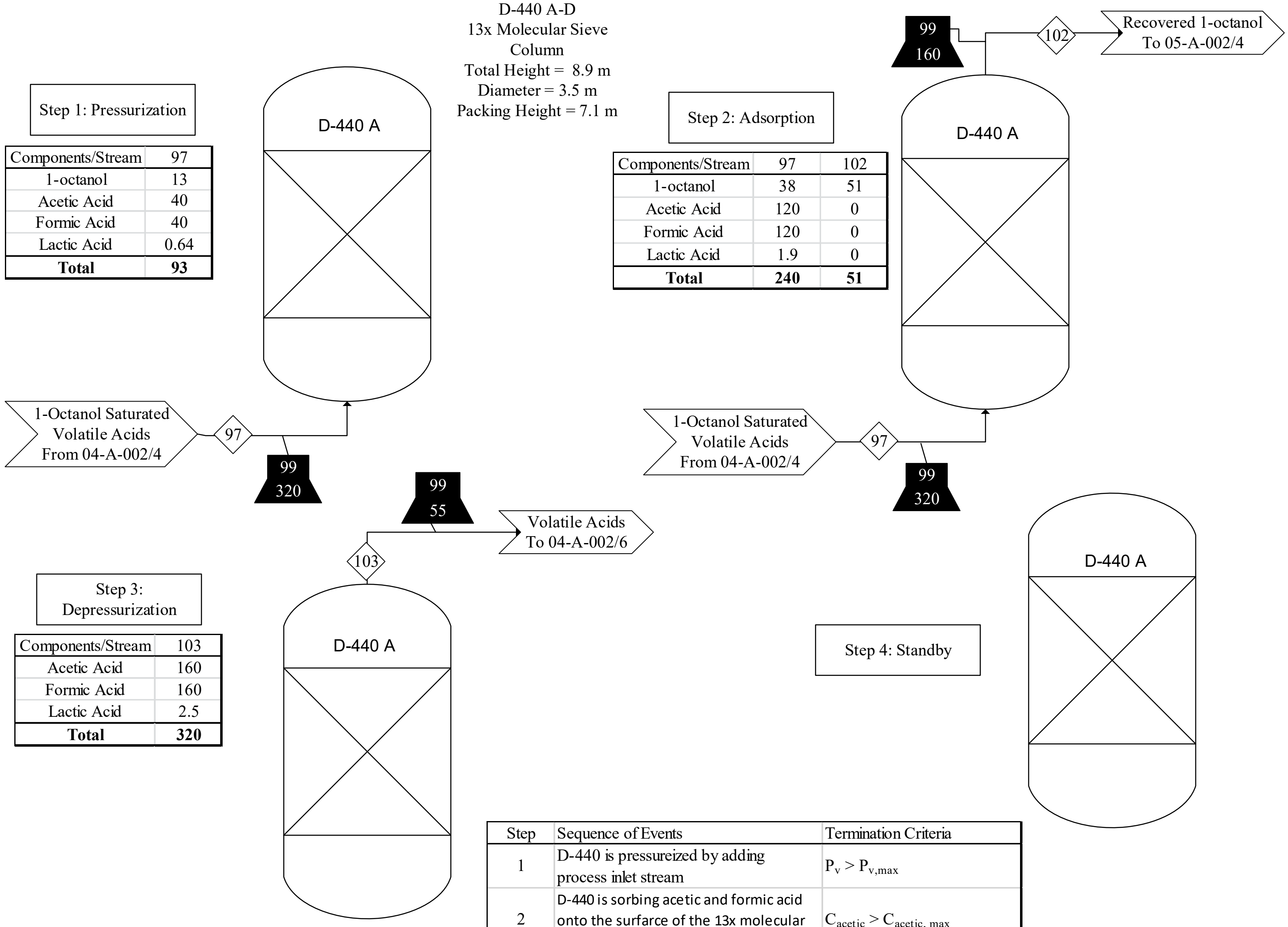
Reference Drawings:
I/O 00-A-001
BFD 04-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 05-A-002

Drawing Type
Process Flow Diagram

Title
Process Area 04: Acetic and Formic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	04-A-002/7	0	443

D-440 A-D
13x Molecular Sieve
Column
Total Height = 8.9 m
Diameter = 3.5 m
Packing Height = 7.1 m



Step 1: Pressurization

Components/Stream	97
1-octanol	13
Acetic Acid	40
Formic Acid	40
Lactic Acid	0.64
Total	**93**

Step 2: Adsorption

Components/Stream	97	102
1-octanol	38	51
Acetic Acid	120	0
Formic Acid	120	0
Lactic Acid	1.9	0
Total	**240**	**51**

Step 3: Depressurization

Components/Stream	103
Acetic Acid	160
Formic Acid	160
Lactic Acid	2.5
Total	**320**

Step 4: Standby

Step	Sequence of Events	Termination Criteria
1	D-440 is pressureized by adding process inlet stream	$P_v > P_{v,max}$
2	D-440 is sorbing acetic and formic acid onto the surface of the 13x molecular sieve	$C_{acetic} > C_{acetic, max}$
3	D-440 is being depressurized, releasing sorbed components.	$P_v < P_{v,min}$
4	D-440 is on standby	$C_{acetic} > C_{acetic, max}$ of previous vessel

Notes

1. Drawing not to scale
2. Mass flow rates are in kg/step
3. Mass balances may not add up due to rounding.
4. Step 1 is 2 minutes
5. Step 2 is 6 minutes
6. Step 3 is 24 minutes
7. Step 4 is 4 minute
8. Four adsorbers used in parallel for continual adsorption

Special Symbols / Nomenclature

Symbol	Meaning
◇ #	Stream Identification
⟩	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 00-A-002
PFD 00-A-003

Drawing Type

Supplemental Process Flow Diagram

Title

13x Molecular Sieve Column Steps

By/Date	DWG # / Sheet #	Rev	Page
AJK 7/05/20	04-A-003/1	0	444

D-500
1st Stage TMA Mixing Drum
Height = 7.8 m
Diameter = 3.9 m

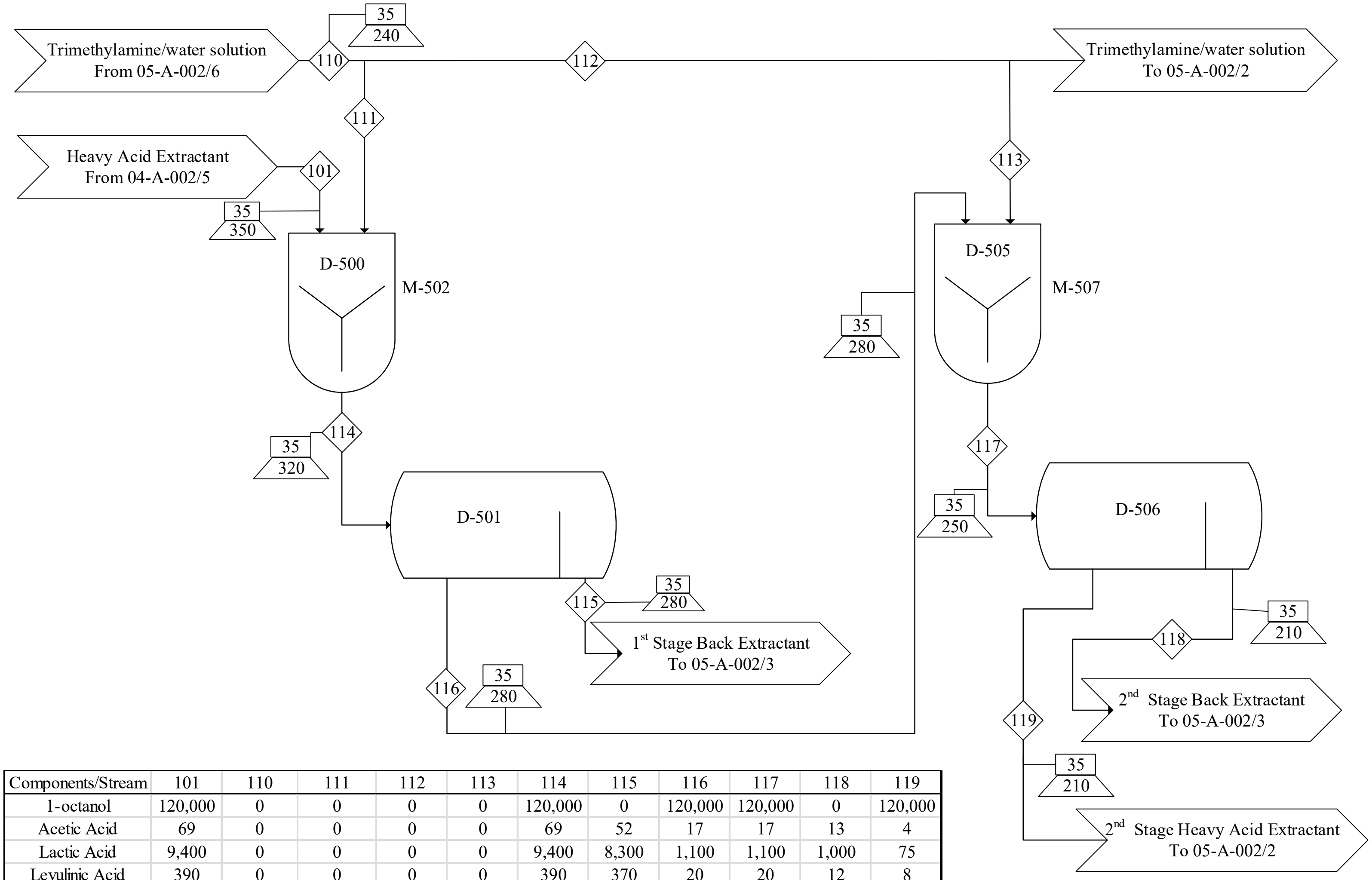
D-501
1st Stage TMA Settler
Height = 2.1 m
Diameter = 4.2 m

M-502
1st Stage TMA Mixer
Agitation Power = 32 kW

D-505
2nd Stage TMA Mixer
Height = 7.8 m
Diameter = 3.9 m

D-506
2nd Stage TMA Settler
Height = 2.1 m
Diameter = 4.2 m

M-507
2nd Stage TMA Mixer
Agitation Power = 32 kW



Components/Stream	101	110	111	112	113	114	115	116	117	118	119
1-octanol	120,000	0	0	0	0	120,000	0	120,000	120,000	0	120,000
Acetic Acid	69	0	0	0	0	69	52	17	17	13	4
Lactic Acid	9,400	0	0	0	0	9,400	8,300	1,100	1,100	1,000	75
Levulinic Acid	390	0	0	0	0	390	370	20	20	12	8
Trimethylamine	0	10,000	8,900	1,100	920	8,900	8,900	0	920	920	0
Trioctylamine	49,000	0	0	0	0	49,000	0	49,000	49,000	0	49,000
Water	0	9,600	8,600	1,000	840	8,600	8,600	0	840	840	0
Total	**180,000**	**20,000**	**18,000**	**2,100**	**1,800**	**200,000**	**26,000**	**170,000**	**170,000**	**2,800**	**170,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/1	0	445

D-510
3rd Stage TMA Mixing Drum
Height = 7.8 m
Diameter = 3.9 m

D-511
3rd Stage TMA Settler
Height = 2.1 m
Diameter = 4.2 m

M-512
3rd Stage TMA Mixer
Agitation Power = 32 kW

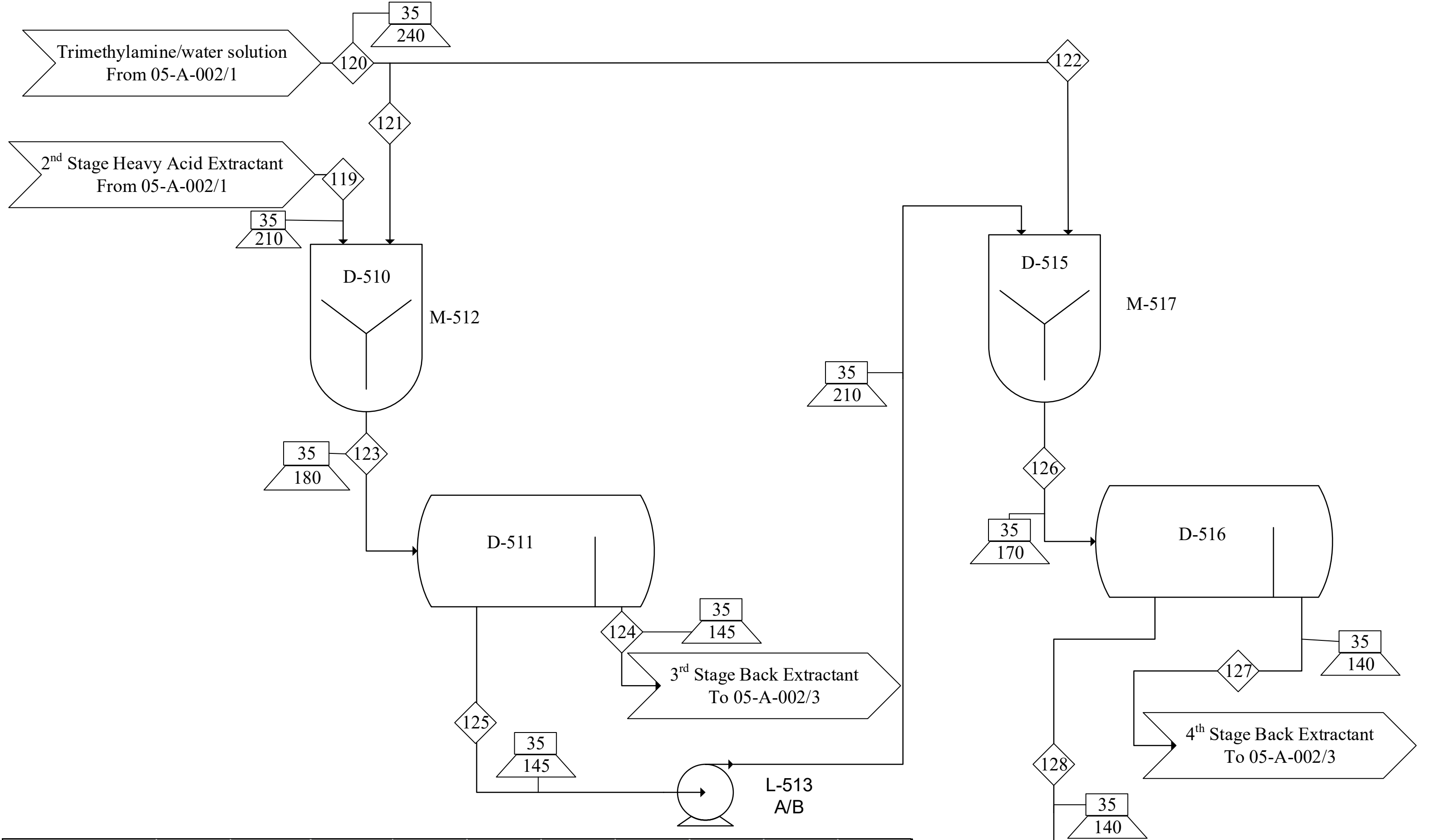
L-513 A/B
3rd Stage TMA Pump
Duty = 5.9 kW

D-515
4th Stage TMA Mixer
Height = 7.8 m
Diameter = 3.9 m

D-516
4th Stage TMA Settler
Height = 2.1 m
Diameter = 4.2 m

M-517
4th Stage TMA Mixer
Agitation Power = 32 kW

L-518 A/B
4th Stage TMA Pump
Duty = 5.8 kW



Components/Stream	119	120	121	122	123	124	125	126	127	128
1-octanol	120,000	0	0	0	120,000	0	120,000	120,000	0	120,000
Acetic Acid	4	0	0	0	4	3	1.0	1.0	1	Trace
Lactic Acid	75	0	0	0	75	64	11	11	8	3.0
Levulinic Acid	8	0	0	0	8	7.5	0.5	0.5	7.5	Trace
Trimethylamine	0	180	100	80	100	100	0	80	100	0
Trioctylamine	49,000	0	0	0	49,000	0	49,000	49,000	0	49,000
Water	0	160	89	71	89	89	0	71	71	0
Total	**170,000**	**340**	**190**	**150**	**170,000**	**250**	**170,000**	**170,000**	**190**	**170,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

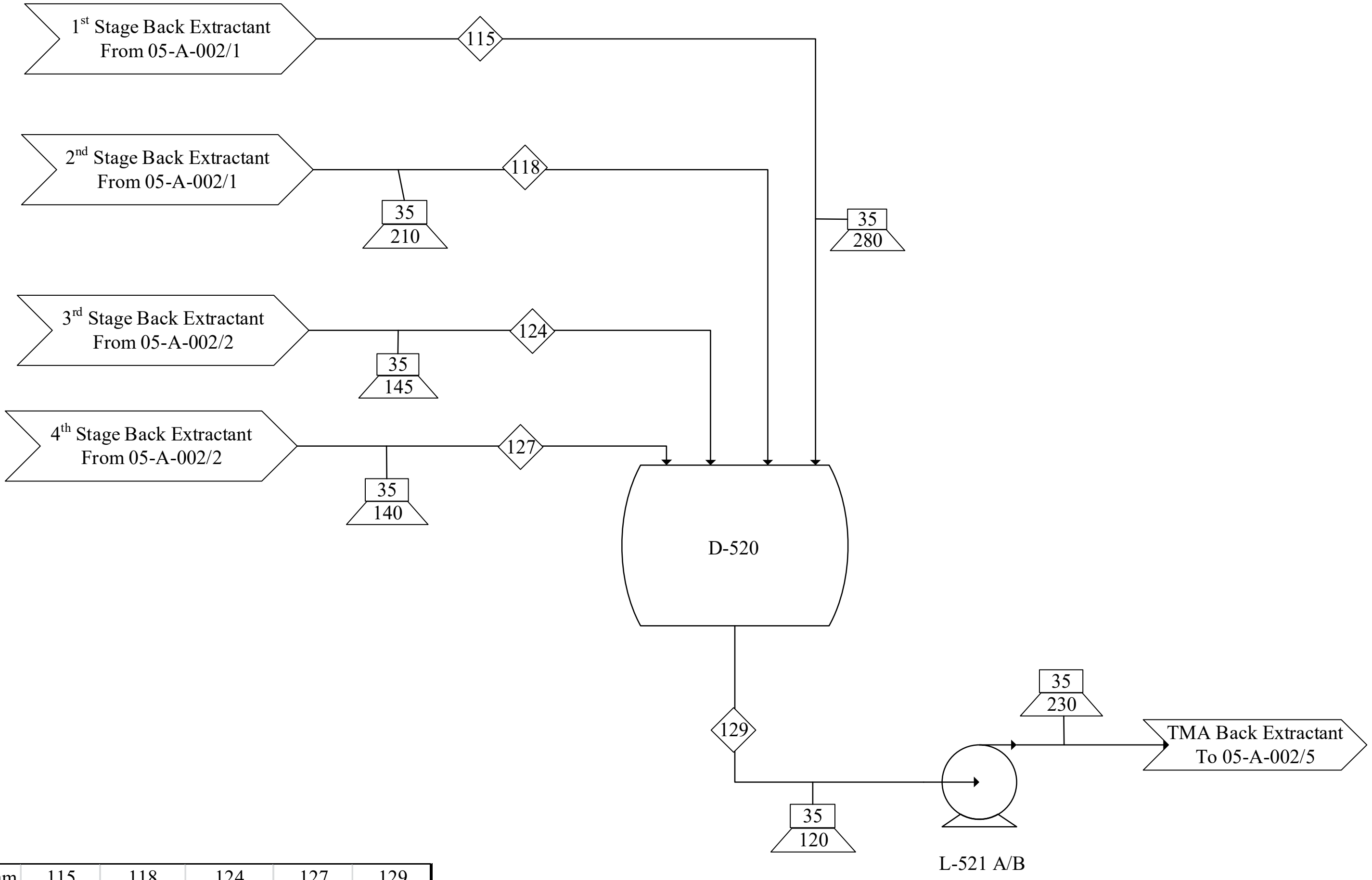
Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/2	0	446

D-520
TMA Acid Extractant
Hold Up Drum
Height = 2.4 m
Diameter = 4.8 m

L-521 A/B
TMA Hold Up Drum
Pump
Duty = 27 kW



Components/Stream	115	118	124	127	129
Acetic Acid	52	13	3	1	69
Lactic Acid	8,300	1,000	64	8	9,400
Levulinic Acid	370	12	7.5	0.5	390
Trimethylamine	8,900	920	100	80	10,000
Water	8,600	840	89	71	9,600
Total	**26,000**	**2,800**	**250**	**160**	**29,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

Stream Identification
Input/Output
Temperature (°C), gas
Temperature (°C), liquid
Pressure (kPa), gas
Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

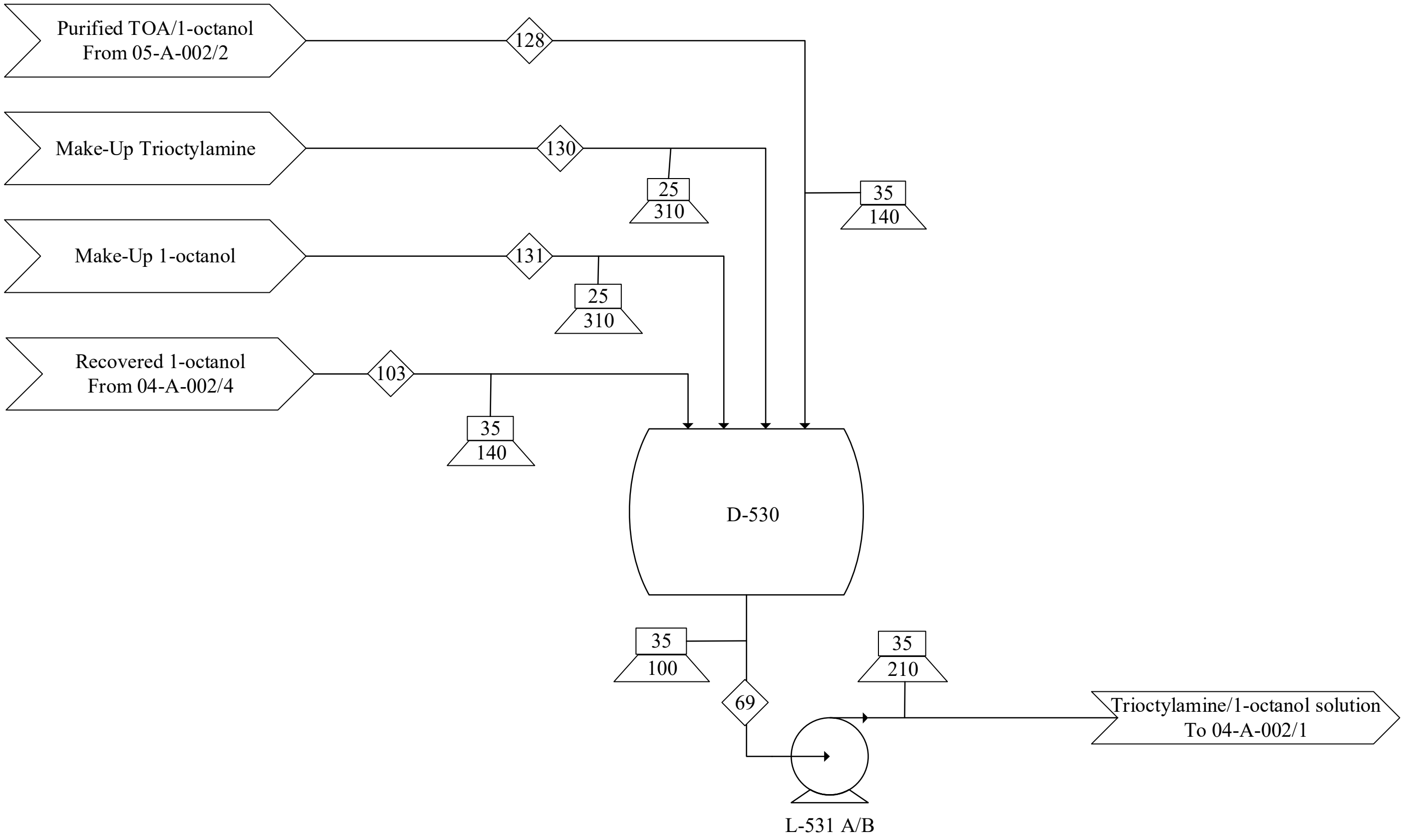
Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/3	0	447

L-531 A/B
TOA Hold Up Drum
Pump
Duty = 11 kW

D-530
TOA Acid Extractant
Hold Up Drum
Height = 4.4 m
Diameter = 2.2 m



Components/Stream	70	103	128	130	131
1-octanol	120,000	380	120,000	0	Trace
Acetic Acid	Trace	0	Trace	0	0
Lactic Acid	3.00	0	3.00	0	0
Levulinic Acid	Trace	0	Trace	0	0
Trioctylamine	49,000	0	49,000	Trace	0
Total	**170,000**	**260**	**170,000**	**Trace**	**Trace**

Notes

1. Drawing not to scale

2. Flow rates are in kg/hr

3. Mass balance may not close due to rounding.

4. Pumps have an efficiency of 65%

5. Make-up 1-octanol and trioctylamine flows are normally zero.

Special Symbols / Nomenclature

◇ # Stream Identification
〉 Input/Output
■ # Temperature (°C), gas
☐ # Temperature (°C), liquid
▲ # Pressure (kPa), gas
△ # Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/4	0	448

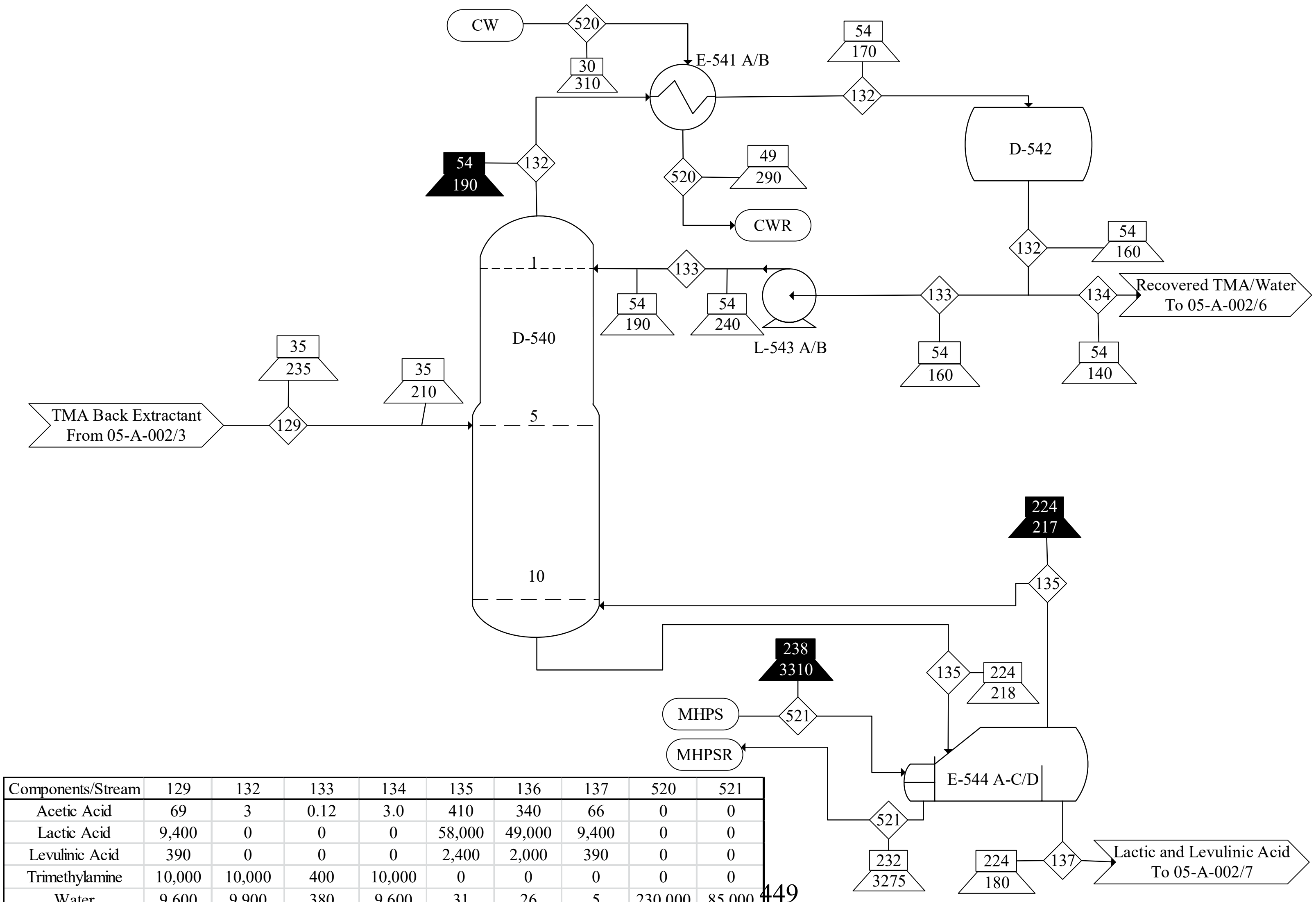
D-540
Back Extractant Solvent Recovery Column
Rectifying Diameter: 1.3 m
Stripping Diameter: 2.0 m
Height: 7.5 m

E-541 A/B
Back Extractant Recovery Condenser
Duty: 29 GJ/hr
Area: 480 m^2

D-542
Back Extractant Recovery Reflux Drum
Height: 2.7 m
Length: 1.3 m

L-543 A/B
Back Extractant Recovery Reflux Pump
Duty: 3.7 kW

E-544 A-C/D
Back Extractant Recovery Reboiler
Duty: 41 GJ/hr
Area: 300 m^2



Components/Stream	129	132	133	134	135	136	137	520	521
Acetic Acid	69	3	0.12	3.0	410	340	66	0	0
Lactic Acid	9,400	0	0	0	58,000	49,000	9,400	0	0
Levulinic Acid	390	0	0	0	2,400	2,000	390	0	0
Trimethylamine	10,000	10,000	400	10,000	0	0	0	0	0
Water	9,600	9,900	380	9,600	31	26	5	230,000	85,000
Total	**29,000**	**20,000**	**780**	**20,000**	**60,000**	**51,000**	**9,900**	**230,000**	**85,000**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Trays have and efficiency of 70%

Special Symbols / Nomenclature

◇ #	Stream Identification
〉	Input/Output
■ #	Temperature (°C), gas
☐ #	Temperature (°C), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

Title

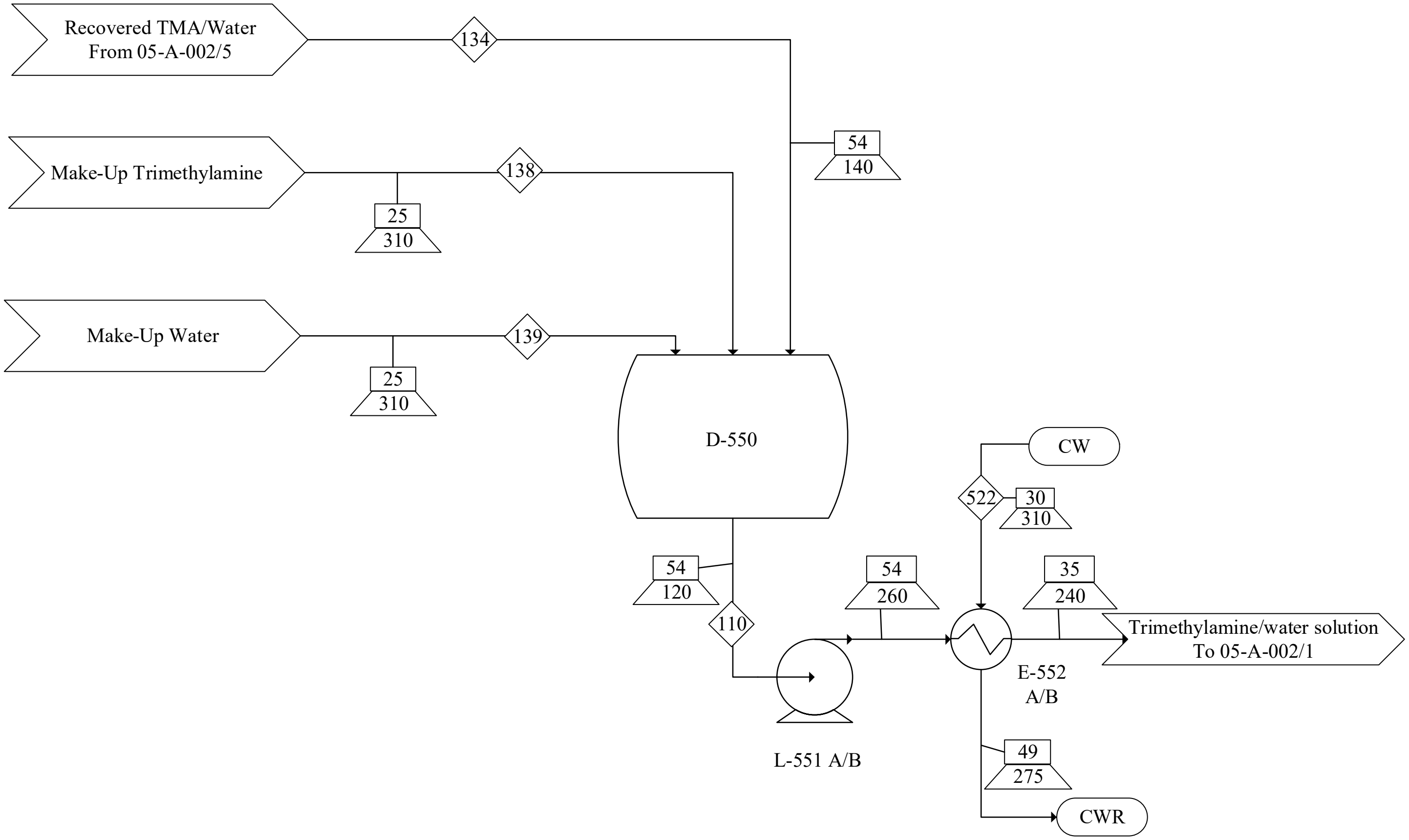
Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/5	0	449

D-550
TMA Acid Extractant
Hold Up Drum
Height = 4.2 m
Diameter = 2.1 m

L-551 A/B
TMA Hold Up Drum
Pump
Duty = 24 kW

E-552 A/B
TMA/Water Cooler
Duty = 0.91 GJ/hr
Area = 160 m^2



Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%
5. Make-up trimethylamine and water flows are normally zero.

Special Symbols / Nomenclature

◇# Stream Identification
▷ Input/Output
■# Temperature (°C), gas
□# Temperature (°C), liquid
▲# Pressure (kPa), gas
△# Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/6	0	450

Components/Stream	110	134	138	139	522
Acetic Acid	3.0	3.0	0	0	0
Trimethylamine	10,000	10,000	Trace	0	0
Water	9,600	9,600	0	Trace	820
Total	**20,000**	**20,000**	**Trace**	**Trace**	**820**

D-560
Lactic/Levulinic Acid Column
Rectifying Diameter: 1.1 m
Stripping Diameter: 0.39 m
Height: 26 m

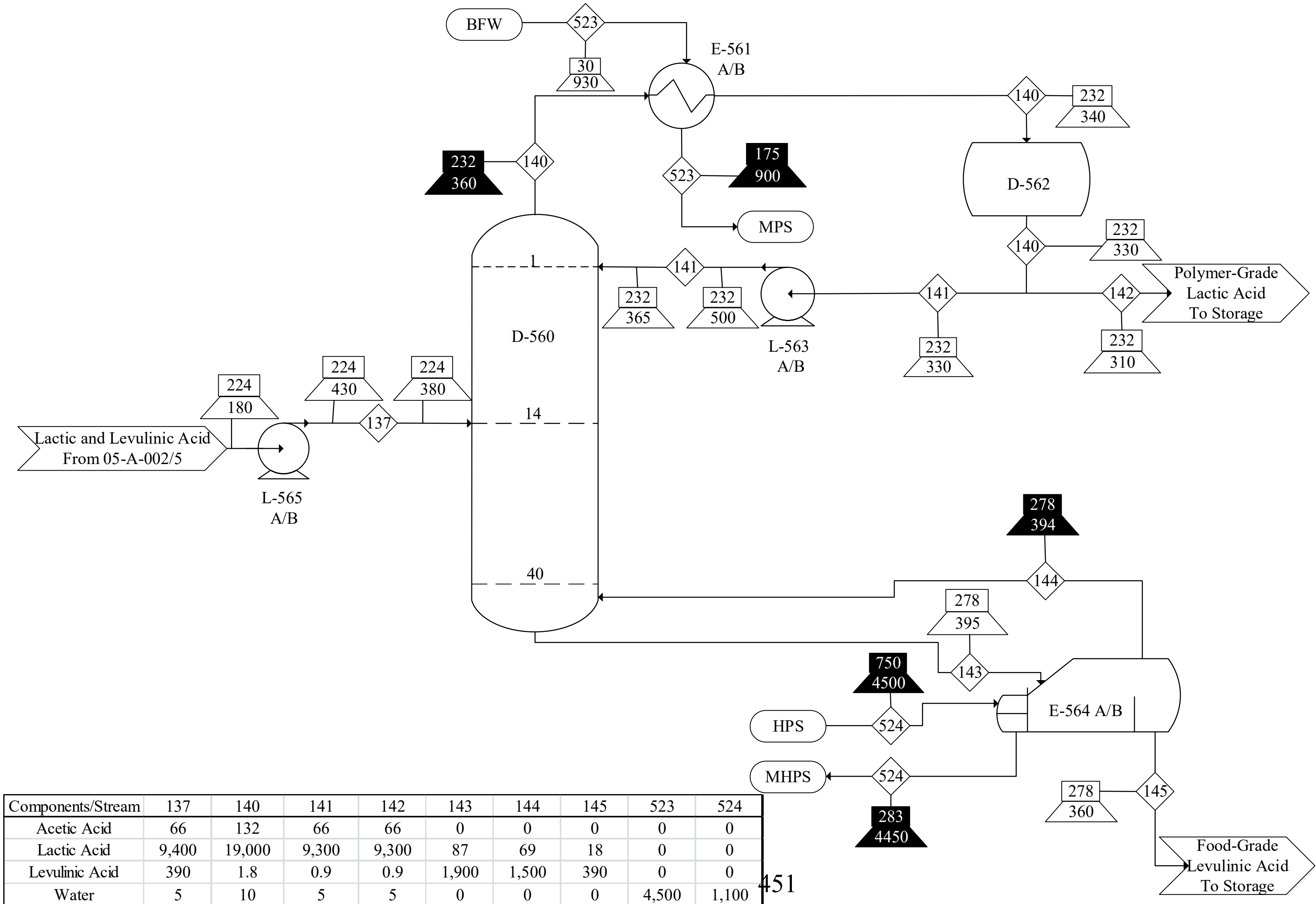
E-561 A/B
Lactic/Levulinic
Condenser
Duty: 13 GJ/hr
Area: 45 m^2

D-562
Lactic/Levulinic Reflux
Drum
Height: 1.6 m
Length: 0.79 m

L-563 A/B
Lactic/Levulinic
Reflux Pump
Duty: 0.92 kW

E-564 A/B
Lactic/Levulinic Reboiler
Duty: 0.84 GJ/hr
Area: 3.7 m^2

L-565 A/B
Lactic/Levulinic
Feed Pump
Duty: 0.74 kW



Components/Stream	137	140	141	142	143	144	145	523	524
Acetic Acid	66	132	66	66	0	0	0	0	0
Lactic Acid	9,400	19,000	9,300	9,300	87	69	18	0	0
Levulinic Acid	390	1.8	0.9	0.9	1,900	1,500	390	0	0
Water	5	10	5	5	0	0	0	4,500	1,100
Total	**9,900**	**19,000**	**9,300**	**9,300**	**2,000**	**1,600**	**410**	**4,500**	**1,100**

Notes

1. Drawing not to scale
2. Flow rates are in kg/hr
3. Mass balance may not close due to rounding.
4. Pumps have an efficiency of 65%

Special Symbols / Nomenclature

◇ #	Stream Identification
▷	Input/Output
■ #	Temperature (°F), gas
☐ #	Temperature (°F), liquid
▲ #	Pressure (kPa), gas
△ #	Pressure (kPa), liquid

Reference Drawings:

I/O 00-A-001
BFD 05-A-001
PFD 01-A-002
PFD 02-A-002
PFD 03-A-002
PFD 04-A-002

Drawing Type

Process Flow Diagram

Title

Process Area 05: Lactic and Levulinic Acid Purification

By/Date	DWG #/ Sheet #	Rev	Pg
AJK 7/05/20	05-A-002/7	0	451

APPENDIX K - REFERENCES

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