Economic Analysis of an Organosolv Process for Bioethanol Production

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In a previous paper, conceptual process design, simulation, and mass and energy balances were presented for an organosolv process with a hardwood feed of 2350 metric tons (MT) per day and ethanol, lignin, furfural, and acetic acid production rates of 459, 310, 6.6, and 30.3 MT/day, respectively. In this paper, the investment and operating costs of the process and the minimum ethanol selling price (MESP) to make the process economically feasible were estimated. The total capital investment of the plant was approximately 720 million USD. Lignin price was found to affect the MESP considerably. With a base case lignin price of 450 USD/MT, the MESP was approximately 3.1 USD per gallon (gal). Higher lignin price of 1000 USD/MT was required to equal the MESP with the December 2013 ethanol market price (2.0 USD/gal). In addition to lignin price, the MESP was found to be strongly affected by feedstock, enzyme, and investment costs. Variations in feedstock and investment costs affected the MESP by approximately 0.2 and 0.5 USD/gal, respectively. Changing the enzyme dosage and price from base case estimate of 5270 USD/MT and 0.02 g/g cellulose to more conservative 3700 USD/MT and 0.06 g/g cellulose, respectively, increased the MESP by 0.59 USD/gal.

Keywords: Organosolv; Economic assessment; Pretreatment; Bioethanol; Lignin; Simulation

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INTRODUCTION

The production of bioethanol from lignocellulosic material has been seen as a potential way to reduce carbon dioxide emissions and dependence on fossil fuels (Hamelinck *et al.* 2005). In the production of this second-generation bioethanol, the natural recalcitrance of lignocellulosic material (Himmel *et al.* 2007) is overcome in a pretreatment step by disrupting the cell structure of the material, enabling the hydrolysis of cellulose to monomer glucose. The monomeric sugars are then fermented to ethanol, and the fermentation beer is dehydrated to a pure ethanol product. For the pretreatment step, several different methods, such as dilute acid, hot water, and steam explosion have been proposed (Hamelinck *et al.* 2005; Mosier *et al.* 2005). Since it has also been considered one of the most expensive processing steps (Mosier *et al.* 2005), the choice and development of a pretreatment method could have a considerable effect on the economics of bioethanol production.

One possible pretreatment method is organosolv pulping in which lignin is dissolved in an organic solvent. The process was originally considered as a pulping method for the production of paper pulp (Hergert 1998; Aziz and Sarkanen 1989), but it has recently gained interest as a potential pretreatment method in bioethanol production (Pan *et al.* 2005; 2006). Although it has generally been considered as an expensive pretreatment method (Zhao *et al.* 2009), the use of organic solvents enables the recovery of a pure lignin product and potentially other co-products (Pan *et al.* 2005; 2006). The production of co-products could possibly offset the perceived high cost of organosolv pulping. The operation of pilot or demonstration scale organosolv biorefineries by companies such as Chempolis, CIMV, and Lignol Innovations (Chempolis 2009; Lignol 2009; CIMV 2013a) can be considered as indications of the interest in organosolv technology.

Several techno-economic studies have been carried out to analyze the technical and economic feasibility of different second-generation bioethanol production processes and to compare different pretreatment methods (Eggeman and Elander 2005; Kazi *et al.* 2010; Humbird *et al.* 2011; Tao *et al.* 2011). These studies are typically based on simulation models, with their key parameters derived from laboratory or pilot scale experiments. The economic feasibility of bioethanol production is often characterized in these studies by the minimum ethanol selling price (MESP) (Humbird *et al.* 2011) covering the operating and capital costs of production and making a certain return on the invested capital.

A recent National Renewable Energy Laboratory (NREL) technical report (Humbird *et al.* 2011) presented a detailed techno-economic study on the production of bioethanol from corn stover based on dilute acid pretreatment and enzymatic hydrolysis. With a corn stover feed of 2000 dry metric tons (MT) per day and a price of 64.5 USD/dry MT, and an in-house enzyme production cost of 0.34 USD per gallon (gal) of ethanol produced, they arrived at an MESP of 2.15 USD/gal. Other techno-economic studies include the studies of Eggeman and Elander (2005), Kazi *et al.* (2010), and Tao *et al.* (2011), which compared the economics of different pretreatment technologies. All of these studies assumed a feedstock feed of 2000 MT/day. Depending on the pretreatment technology, the capital costs were in the range of 164 to 212 million US dollars (MUSD) (Eggeman and Elander 2005), 327 to 391 MUSD (in 2007 dollars) (Kazi *et al.* 2010), and 325 to 385 MUSD (in 2007 dollars) (Tao *et al.* 2011). The MESPs were in the range of 1.3 to 1.7 (Eggeman and Elander 2005), 3.4 to 4.5 (Kazi *et al.* 2010), and 2.7 to 4.1 USD/gal (Tao *et al.* 2011). A more extensive review of recent techno-economic studies was presented in the NREL study (Humbird *et al.* 2011). In this review, the MESPs were found to vary between different studies in the range of approximately 0.6 to 4.6 USD/gal of ethanol produced, with the variations being largely explained by differences in assumed feedstock prices, process yields, and co-product credits. In sensitivity analyses presented by Kazi *et al.* (2010) and Humbird *et al.* (2011), MESPs were found to be very sensitive to variations in capital costs (Humbird *et al.* 2011), enzyme cost, and cellulose to glucose conversion in enzymatic hydrolysis (Kazi *et al.* 2010; Humbird *et al.* 2011), and feedstock cost (Kazi *et al.* 2010).

A few economic assessments on organosolv processes have also been published. For example, van der Linden *et al.* (2012) reported in a conference presentation the income distributions, capital costs, operating costs, and payback times for an ethanol-organosolv biorefinery producing cellulose pulp, lignin, and furfural at assumed prices of 350 EUR/MT (455 USD/MT, using an EUR-USD exchange rate of 1.3 both here and elsewhere in this work), 750 EUR/MT (975 USD/MT), and 625 EUR/MT (813 USD/MT), respectively, from various feedstocks. At a feedstock feed of 150,000 MT/year, the total fixed capital was estimated to be 80 to 87 million EUR (104 to 113 MUSD) with the organosolv reactor section being the single most expensive section in the process. The payback time was found to be in the approximate range of 6 to 7 years. Dias *et al.* (2011) compared a first generation sugarcane juice bioethanol process to different integrated first and second-generation bioethanol processes (both sugarcane juice, bagasse, and trash). The second-generation process was based on different pretreatment methods. An organosolv-based process was found to result in investment and production costs and internal rates of return that were relatively similar to those of a steam explosion-based process. Lignin from the organosolv pulping process was assumed to be combusted. González Alriols *et al.* (2010) presented an organosolv biorefinery process where an ultrafiltration separation system was used to produce lignin fractions with different molecular weights. Capital and operating costs were calculated for the ultrafiltration system, resulting in a production cost of 52 EUR/MT (68 USD/MT) of lignin. Mabee *et al.* (2006) estimated the mass and energy balances and investment and operating costs for bioethanol production processes based on acid-catalyzed ethanol organosolv and sulfur dioxide-catalyzed steam explosion pretreatments of Douglas fir. They compared the processes and found the total production costs of the organosolv process to be lower. They further found the co-product lignin sales price to significantly affect the production costs. Only relative costs were, however, presented in these analyses. Absolute figures enabling the economic assessment of the organosolv process were not provided. Parajó and Santos (1995) presented a techno-economic study on the acid-catalyzed acetic acid pulping of *Eucalyptus globulus* wood for the production of paper pulp and co-products. Based on a proposed flowsheet and mass and energy balances, they calculated the investment and operating costs of the process and further assessed its profitability. They varied the lignin price and found the process to be clearly unprofitable and non-competitive with other paper pulp production processes if the lignin was sold at its fuel value. The process was closer to profitability if the lignin was sold at a price corresponding to its use as an asphalt extender and profitable at higher prices, corresponding to its use as a dispersing agent, intermediate in polymer manufacture or as a phenol extender in phenol-formaldehyde resins.

Although economic assessments of organosolv processes have been published (Parajó and Santos 1995; Mabee *et al.* 2006; González Alriols *et al.* 2010; Dias *et al.* 2011; van der Linden *et al.* 2012), no comprehensive studies analyzing the economic feasibility of ethanol production as well as the effect of co-products on the economics based on detailed flowsheets and balances are known. This paper is a continuation of a previous paper (Kautto *et al.* 2013) in which the technical and simulation aspects of an acid-catalyzed ethanol organosolv process for the production of bioethanol and co-products, lignin, acetic acid, and furfural, from lignocellulosic material were examined. Based on the detailed process flowsheets and mass and energy balances presented in the previous paper, the economics of the organosolv biorefinery will be analyzed in this paper. The estimated capital costs including the sizing and costing of the main equipment and the operating costs of the process will first be presented. The MESP for a base case scenario will then be calculated and the effect of variations in main technical and economic attributes on the MESP will be analyzed in sensitivity analyses. The effect of the recovery of co-products on the feasibility of the process will also be discussed. Whenever applicable, the technical analysis of the organosolv process presented earlier

(Kautto *et al.* 2013) followed closely the NREL dilute acid pretreatment study (Humbird *et al.* 2011). Assumptions adopted in this economic analysis will also be similar to those of the NREL study (Humbird *et al.* 2011), making the two processes comparable. The economic feasibility of the organosolv process will therefore be compared to the dilute acid process. Their relative differences will be discussed, and the conditions under which the organosolv process could be competitive will be analyzed.

MATERIALS AND METHODS

Overview of the Studied Organosolv Biorefinery Concept

As discussed in the *Introduction*, the economic analysis carried out in this paper was based on a technical analysis presented previously (Kautto *et al.* 2013). The detailed flowsheets and mass and energy balances of the previous paper were used to estimate the operating and capital costs and revenues of the process.

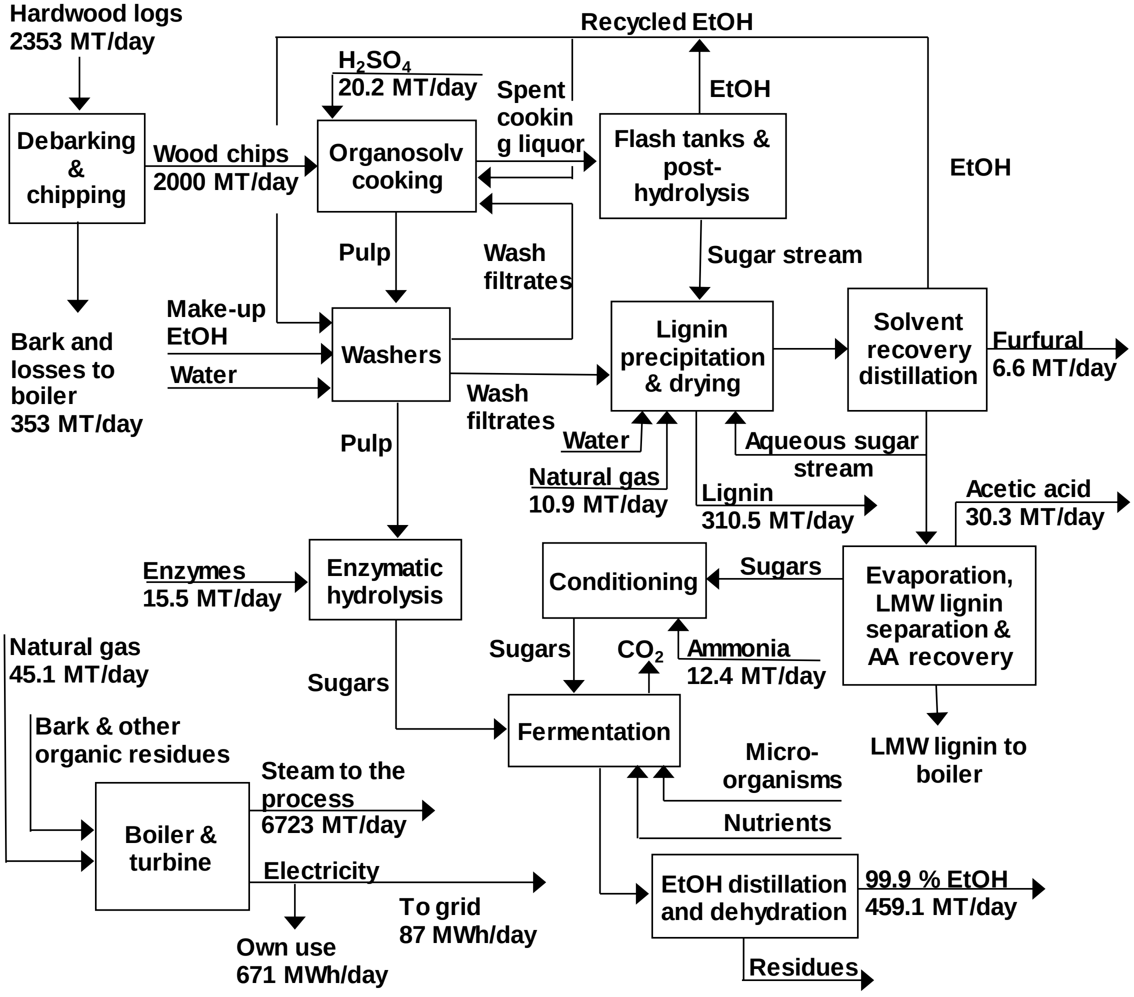

Fig. 1. Simplified block diagram of the modeled ethanol organosolv process with the main process steps and input and output flows

With an intake of 2350 dry MT/day of undebarked hardwood, resulting in 2000 MT/day of debarked chips fed to cooking, the production rates of ethanol, lignin, furfural, and acetic acid were 459.1, 310.5, 6.6, and 30.3 MT/day, respectively. A small amount of natural gas was needed to cover the steam balance of the process. The hardwood was assumed to be hybrid poplar. A simplified block diagram of the modeled process with the main process steps and input and output flows is presented in Fig. 1. More detailed process and balance information can be found in the previous study (Kautto *et al.* 2013). For the purposes of the economic assessment, the plant was assumed to be located in the USA.

Investment Costs

Due to the relatively high level of detail of the NREL study (Humbird *et al.* 2011) and significant similarities between that and our study in several process areas, most investment costs were assessed based on data of that study. The organosolv process plant was divided into process areas for investment cost analysis. The costs of areas assumed similar to the NREL study (Humbird *et al.* 2011) were estimated by directly scaling the costs from that study on a process area basis. These areas were wastewater treatment, storage, utilities, and boiler and turbine. The costs of all the other process areas were estimated on an individual equipment basis, using the equipment costs, installation factors, and scaling exponents of the NREL study (Humbird *et al.* 2011) as well as the Aspen Process Economic Analyzer software (AspenTech 2011), industry estimates, and a previous NREL study from 2002 (Aden *et al.* 2002). Where Aspen Process Economic Analyzer was used in estimating the equipment costs, the installed costs of these pieces of equipment were calculated using installation factors of the NREL study (Humbird *et al.* 2011).

Material of construction in most areas was either SS304 or SS316, depending on the process conditions. All prices were indexed to the year of 2013 using Chemical Engineering Plant Cost Index (CEPCI). A preliminary value of April 2013 was used as the value for 2013 (Chemical Engineering 2013). Table 1 reports the process areas, their descriptions, and the costing bases.

After assessing the installed equipment costs of different process areas, the costs of warehouse, site development, additional piping, different indirect costs, and other costs as well as the amount of working capital were calculated following the data and methodology presented in the NREL study (Humbird *et al.* 2011). The costs of warehouse, site development, and additional piping were calculated in the NREL study (Humbird *et al.* 2011) as a percentage of the inside battery limits (ISBL) investment cost. In this study the ISBL was set to consist of process areas of feed handling, pretreatment and lignin recovery, solvent and furfural recovery, conditioning of the hemicellulosic sugar stream, acetic acid recovery, enzymatic hydrolysis and fermentation, and ethanol product and solids recovery.

The indirect costs were calculated in the NREL study (Humbird *et al.* 2011) as a percentage of the total direct costs. The working capital was calculated as a percentage of the fixed capital investment (the sum of direct and indirect costs). All the percentages used in this study are presented in Table 4 in the subsection *Base case economic assessment*.

Table 1. Process Areas and Bases of Costing

Process area	**Description**	**Basis of costing**	**Source and scaling basis**
Feed handling	Wood yard (wood receiving, debarking, and chipping)	Process area	Industry estimate
Pretreatment and lignin recovery	Cooking, pulp washing, post hydrolysis, precipitation, and drying of lignin	Individual equipment	Industry estimate, (AspenTech 2011; Humbird *et al.* 2011;)
Solvent and furfural recovery	Solvent and furfural recovery columns	Individual equipment	(AspenTech 2011; Humbird *et al.* 2011;)
Conditioning of the hemicellulosic sugar stream	Evaporation, separation of low molecular weight lignin, ammonia treatment	Individual equipment	(Aden *et al.* 2002; AspenTech 2011; Humbird *et al.* 2011;)
Acetic acid recovery	Extraction of acetic acid from evaporator condensates, distillation to pure acetic acid product	Individual equipment	(AspenTech 2011)
Hydrolysis and fermentation	Enzymatic hydrolysis and fermentation to ethanol	Individual equipment	(Humbird *et al.* 2011)
Ethanol product and solids recovery	Distillation of pure ethanol product, filtration of distillation bottoms solids	Individual equipment	(Humbird *et al.* 2011)
Wastewater treatment	Anaerobic and aerobic treatment systems	Process area	(Humbird *et al.* 2011), assumed similarly sized
Storages	End-product and raw material storages	Process area	(Humbird *et al.* 2011), assumed similarly sized
Boiler and turbine	Combustion of organic residues and purchased natural gas and generation of steam and electricity	Process area	(Humbird *et al.* 2011), scaled on the amount of combustion heat
Utilities	Cooling and chilled water, plant/instrument air, and process water systems	Process area	(Humbird *et al.* 2011), cooling and chilled water systems scaled on cooling duty, everything else assumed similar

Annual Cash Flows

Cost of feedstock

Various types of biomass could be utilized in the production of second generation ethanol, including agricultural residues such as corn stover, straws and bagasse, forest biomass and woody residues (hardwoods and softwoods, various forest residues, and mill residues from the forest industry), and herbaceous grasses such as switchgrass (see Huang *et al.* 2009; U.S. Department of Energy 2011). Due to differences in composition (Santos *et al.* 2012) and delivered cost (Huang *et al.* 2009; Gnansounou and Dauriat 2010; Gonzalez *et al.* 2011), the type of feedstock has an effect on the economics of bioethanol production (Huang *et al.* 2009; Gnansounou and Dauriat 2010; Gonzalez *et al.* 2011; Santos *et al.* 2012). Moreover, various feedstocks have different availabilities and delivered costs depending on the exact location of the biomass processing facility.

In the technical part of this study, mass and energy balances over the organosolv process were calculated assuming a hybrid poplar feedstock (Kautto *et al.* 2013). Although organosolv pretreatment has been demonstrated to be suitable for a wide range of feedstocks (see Pan *et al.* 2005 for softwood, Pan *et al.* 2006 for hardwood, and Mesa *et al.* 2010 for sugarcane bagasse), the economic analysis is also carried out assuming the same feedstock. For bioenergy and biofuel production, various poplar prices have been reported in the literature. Table 2 lists some reported price estimates.

Table 2. Reported Poplar Prices

Reference	Poplar price/cost	Location and plant capacity
Phillips *et al.* 2007	Delivered cost target of 35 USD/dry short ton of wood chips (in 2005 dollars, approximately 39 USD/dry MT)	2000 dry MT/day
Huang *et al.* 2009	Delivered cost of cropland grown hybrid poplar chips estimated at 104.65 USD/dry MT (in 2005 dollars), with the cost consisting of land, production, harvest, chipping, fixed (independent of distance) transportation, and variable (dependent of distance) transportation costs of 20.50, 38.64, 16.0, 14.0, 4.55, and 10.96 (with a collection radius slightly below 60 km) USD/dry MT, respectively	Minnesota, 2000 dry MT/day
Jones *et al.* 2009	Hybrid poplar wood chip delivered price of 50.7 USD/dry short ton (in 2007 dollars, 55.9 USD/dry MT)	2000 dry MT/day
Gnansounou and Dauriat 2010	Delivered cost of poplar estimated at 140.4 USD/dry ton (in 2007 dollars), with the non-transport costs being 123.3 and transport costs 17.1 USD/dry ton (average collection radius 40 km)	1636 dry ton/day,
Khanna and Huang 2010	Of the studied three states, farm-gate break-even price of poplar chips was found to be the lowest in Michigan, being in the range 88 – 95 USD/dry MT (in 2007 dollars) when grown on marginal land and 106 – 115 USD/dry MT when grown on cropland. The break-even prices were found to be sensitive to changes in biomass yield and harvest cost both on marginal and cropland and to changes in the prices of corn and soybean on cropland (corn-soybean rotation was assumed to be the most profitable use of land and therefore determined the opportunity cost of cropland).	Michigan, Illinois, Oklahoma
Berguson *et al.* 2010	Delivered costs of hybrid poplar chips estimated to be in the range of 70 – 100 USD/dry short ton (approx. 77 – 110 USD/dry MT) in Pacific Northwest, with the exact value depending on the assumed land costs and discount rates. The costs consisted of 27% land, 9% establishment, 28% crop care and management, 24% harvest, and 12% transport costs. The prices were estimated to be in the same range also in Minnesota.	Pacific Northwest, Minnesota, average one-way hauls 30-65 miles

The study of Berguson *et al.* (2010) was followed in setting the feedstock price. Since hardwood feedstock was assumed to be debarked and chipped onsite in this organosolv study (Kautto *et al.* 2013) and Berguson *et al.* (2010) reported the cost of hybrid poplar as chipped, a minor conversion was made to the costs of Berguson *et al.* (2010). The base case price of the undebarked poplar wood was assumed in this study to be 85 USD/dry MT delivered to the plant gate, and the feedstock price was assumed to vary in the range of 70 to 100 USD/dry MT in sensitivity analyses. The effect of these changes on the MESP is discussed in more detail in subsection *Sensitivity analyses*.

Cost of enzymes

The cost contribution of enzymes in techno-economic second generation ethanol studies has typically been reported either on a very aggregated level as a cost contribution per gallon of ethanol produced or calculated based on assumed price (or production cost) and loading of enzymes. A wide range of enzyme cost contributions have been assumed in these studies, including a cost goal of 0.1 USD/gal (approximately 1.5 USD/kg cellulase at a loading of 0.02 g cellulase/g cellulose) in the 2002 NREL study (Aden *et al.* 2002), future price projection of 0.15 USD/gal in (Eggeman and Elander 2005), 0.25 USD/gal (Tao *et al.* 2011), and costs of 0.71, 0.35, and 2.09 USD/gal reflecting a loading of 0.0313 g protein/g cellulose in untreated biomass and a base case enzyme price of approximately 5.1 USD/kg derived from an enzyme production model and lower and higher price scenarios of 2.6, and 14.6 USD/kg, respectively (Kazi *et al.* 2010).

Some studies have presented more detailed analysis on the economics of enzyme production. Klein-Marcuschamer *et al.* (2011) carried out a techno-economic analysis of cellulase production from steam-exploded poplar wood by fungus *Trichoderma reesei*. At a cellulase production capacity enough to supply a cellulosic ethanol plant processing 1700 dry MT of corn stover per day (Klein-Marcuschamer *et al.* 2010), a poplar cost of 60 USD/MT, residence time of 192 h in aerobic fermentation, and a final enzyme concentration after filtration of 150 g/L, they arrived at a baseline cost of enzymes of 10.14 USD/kg (in 2010 dollars). This corresponded to a cost contribution of 1.47 USD/gal ethanol at a loading of 0.02 g enzyme/g cellulose. The enzyme production cost was found to be largely governed by capital (48% of total) and feedstock costs (28%). An on-site enzyme production process was also modeled in the NREL study (Humbird *et al.* 2011). Using glucose as the sugar source in fermentation, they arrived at a somewhat lower enzyme cost of 4.24 USD/kg (in 2007 dollars), which resulted in a cost contribution of 0.34 USD/gal ethanol at a loading of 0.02 g enzyme protein/g cellulose. The key cost contributors were found to be the sugar (57% of total) and capital costs (21%). The differences in enzyme production costs and cost structures between the NREL study (Humbird *et al.* 2011) and Klein-Marcuschamer *et al.* (2011) reflect at least partially different feedstocks (sugar sources) used in enzyme production.

Although the NREL enzyme cost could be considered rather optimistic, it was here used as a basis to calculate the base case enzyme cost to align this study with the NREL study. Based on the enzyme production cost breakdown presented in the NREL study (glucose, other nutrients, electricity, and fixed and capital costs), the cost was in this study indexed to the year 2013 using the plant cost index, the 2013 glucose price (USDA 2013), producer price index for industrial chemicals, and the 2013 electricity price. This resulted in an enzyme cost of 5270 USD/MT. Also, the enzyme loading (0.02 g enzyme/g cellulose) was assumed to be the same as in the NREL study. The enzyme was assumed to be supplied by a separate operator at the indexed enzyme production cost. Price estimates (3700 USD/MT) and dosing suggestions (0.06 g enzyme/g cellulose) from an enzyme provider from 2011 (Novozymes 2011) were used as a more conservative scenario in sensitivity analyses.

Other assumptions related to variable and fixed costs and revenues

There is no commercial production and therefore no existing market for the quantities of organosolv lignin that would be generated through such large scale production as studied in this paper. Furthermore, there is a wide range of products that could be assumed to be produced from organosolv lignin. Estimating the market price for the lignin is therefore difficult. The price of lignin was assumed to be 450 USD/MT in the base case scenario. This is in line with Gosselink (2011) who reported kraft (lignin separated from kraft pulp cooking liquor), soda (lignin from soda pulp cooking liquor), and organosolv lignin values in the range of 350 to 500 EUR/MT (455 to 650 USD/MT).

Higher price assumptions were reported by Baker (2010) in a presentation on the use of organosolv and kraft lignins in the production of carbon fibers (lignin price below 1100 USD/MT), by van der Linden *et al.* (2012) (975 USD/MT), and by Rushton (2012) in a presentation on Lignol's biorefining revenue scenarios (1900 USD/MT). Indication of a higher lignin price can also be drawn from the earlier activities of a company called Repap. The company operated a 30 MT/day organosolv pulping demonstration plant (so called Alcell process) intermittently from 1989 to 1996, producing more than 3700 MT of lignin and selling most of it for its use for example in phenolic resins (Pye 2010a). Repap further planned a larger, 450 MT/day organosolv pulping plant, and undertook studies to evaluate markets for the larger lignin output. According to Pye (2010a), the studies suggested that larger quantities of organosolv lignin could have been sold at an average price close to the price of phenolic resin (see subsection *Sensitivity analyses* for the current market price). Although some of the literature and earlier experience seem to suggest that relatively high prices could be obtained for organosolv lignin, the more conservative price assumption of 450 USD/MT, representing the lower end of the price range presented in the literature, was seen as justifiable in this paper due to lack of price data of current, larger scale sales of organosolv lignin. The pricing of lignin as well as the effect of changes to the assumed lignin price are discussed in more detail in subsection *Sensitivity analyses*.

The prices of chemicals were estimated mainly based on the NREL study (Humbird *et al.* 2011). Other sources were used for acetic acid (Kelley 2013), furfural (Arato *et al.* 2005), and kerosene (Dennis *et al.* 2013). The chemical prices were indexed to the year 2013 using CEPCI producer price index for industrial chemicals (Chemical Engineering 2013). The prices of natural gas and electricity represented price level in December 2013 and were based on data of U.S. Energy Information Administration (2013a,b). Other cost items, including salaries, labor burden, maintenance, and property insurance were estimated based on the NREL study (Humbird *et al.* 2011). The amount of personnel was assumed similar to that in the NREL study (Humbird *et al.* 2011), and the labor cost was indexed to 2013 using labor cost index for the chemical industry (Bureau of Labor Statistics 2013).

MESP Analysis

Following assumptions made in the NREL study (Humbird *et al.* 2011), plant on-line time of approximately 350 days per year was assumed in this study. The plant was assumed to be in full operation for 30 years. Table 3 presents the assumed time-table and cash flows for planning and engineering, construction, and start-up.

Table 3. Plant Planning and Engineering, Construction, and Start-up Times

Phase	Time, months	Cash flow
Planning and engineering	0 – 12	8% of project investment
Construction	12 – 24 24 – 36	60% of project investment 32% of project investment +working capital
Start-up	36 – 39	50% of production 75% of variable costs 100% of fixed costs
Full production	39 →	

Based on the estimated investment costs and annual cash flows, the economic feasibility of the organosolv plant was assessed using the MESP, which is widely used in the literature studying different bioethanol production concepts. The MESP is the selling price of ethanol that makes the net present value of the ethanol process equal to zero (Humbird *et al.* 2011) so that the ethanol sales revenue fully compensates for both the operating and capital costs of the project with a given discount rate. The MESP calculation was done following the recent and earlier NREL studies (Aden *et al.* 2002; Humbird *et al.* 2011). The discount rate was set to 10%, capital investment was assumed to be 40% equity financed, the interest on loan was 8% and it was paid back in 10 years, corporate tax rate was 35%, and a 150% declining base depreciation method with a 20-year recovery period was adopted for the steam and power production plant and a 200% declining base with a 7-year recovery for the rest of the equipment.

As either price information from 2013 or prices indexed to 2013 price level using CEPCI index values were used throughout this study, the dollars used in this study can be considered as 2013 dollars.

Technical Parameters Varied in Sensitivity Analyses

The economic assumptions described above and the technical assumptions described in the companion paper (Kautto *et al.* 2013) outline the base case scenario of the modeled organosolv process. Due to uncertainties inherent in a conceptual process simulation study like this, sensitivity analyses were run to estimate the effect of changes in some main parameters on the MESP of the studied organosolv concept. Both technical (internal) and exogenous (external, market-related) parameters were analyzed. The lignin sales price and feedstock, enzyme, and investment costs were selected as exogenous parameters, and parameters related to enzymatic hydrolysis, fermentation, and energy consumption were selected as technical parameters. Regarding the market-related parameters, variations in feedstock and enzyme costs were discussed in subsections *Cost of feedstock* and *Cost of enzymes*, investment costs were assumed to vary by 30%, and the sales price of lignin will be discussed separately in subsection *Sensitivity analyses*. The rationale for the selection of the technical parameters and their variation is discussed below.

Ethanol organosolv pulps have generally been found to exhibit good response to enzymatic hydrolysis (Pan *et al.* 2005; 2006). Pan *et al.* (2006) presented how low lignin content organosolv pulps exhibit significantly better response to enzymatic hydrolysis than higher lignin content steam-exploded wood. With approximately the same enzyme loading (20.9 mg cellulase and 5.7 mg β-glucosidase per g cellulose) and same hydrolysis time of 36 h (but with a significantly lower solids content), Pan *et al.* (2006) reported hybrid poplar pulp prepared at the same conditions adopted in this study to exhibit a conversion efficiency of cellulose to glucose of approximately 96%. The enzymatic conversion of organosolv-cooked hybrid poplar might therefore be higher than that of dilute acid-pretreated corn stover of the NREL study (Humbird *et al.* 2011) (cellulose to glucose conversion of 90%). The effect of improved enzymatic hydrolysis on the economics of the organosolv process was studied by setting sugar conversion yield 6 percentage points higher for all carbohydrates. On the other hand, due to enzyme inhibition and deactivation caused by soluble sugars, furfural, hydroxymethyl furfural, organic acids, and phenolics at higher solids contents (see Kim *et al.* 2011), the conversion rates in industrial scale enzymatic hydrolysis could be significantly lower than those reported by Pan *et al.* (2006). Therefore, a conversion of both cellulose and hemicelluloses to sugars of 80% was adopted as a more conservative scenario for the enzymatic hydrolysis.

In addition to assuming an improved enzymatic hydrolysis as described above, a higher overall sugar recovery could have also been achieved by assuming a higher carbohydrate recovery in the cooking section. As discussed in the previous paper (Kautto *et al.* 2013), the carbohydrate recovery was assumed to be 84% based on Pan *et al.* (2006), being considerably lower than that in the NREL study (Humbird *et al.* 2011) (approximately 98%). Mabee *et al.* (2006) reported a higher recovery in the organosolv cooking of softwood (90% for both glucose and hemicelluloses), suggesting that a higher carbohydrate recovery could be achievable. Improved carbohydrate recovery was, however, not specifically studied in the sensitivity analyses since the effect of such a change would be similar to that of the improved enzymatic hydrolysis.

Mannose and galactose were not assumed to be fermented by the fermenting organism *Zymomonas mobilis*, as outlined in the NREL study (Humbird *et al.* 2011). The fermentation of these sugars has, however, been assumed in several studies (see Aden *et al.* 2002). On the other hand, the fermentation of xylose and other hemicellulosic sugars has generally proven to be challenging (see Aden *et al.* 2002). In addition, due partially to relatively high temperature and long residence time in organosolv cooking (assumed to be 180 °C and 60 minutes (Kautto *et al.* 2013), based on Pan *et al.* (2006)), and the need to close the cooking mass balance, the aqueous stream from cooking was assumed to contain initially relatively high amounts of sugar degradation products (see discussion in the companion paper (Kautto *et al.* 2013)). Although the amount of these and other inhibitory compounds were assumed to be decreased to a level low enough for fermentation (Kautto *et al.* 2013), residual inhibitory compounds might affect the fermentation, and no experimental data were available on the fermentability of such a stream. The processing of the hemicellulosic sugar stream could also result in greater sugar losses than assumed; therefore, both the effect of fermentation of mannose and galactose and of a lower fermentation yield of hemicellulosic sugars on the economics of the organosolv process were considered.

Due to the preliminary nature of this study, several assumptions had to be made that would have a considerable effect on the steam and electricity consumption of the process. For example, the liquor-to-wood ratio (LTW) was assumed to be 5. With a higher or lower LTW, the amount of solvent and water recycled, and therefore the steam consumption in distillation, would be changed significantly. Also the extent of, and assumptions made in, heat integration would affect the steam consumption. Due to this inherent uncertainty, the effects of changes in the energy consumption of the process were studied. The steam and electricity consumption of the process were changed by 30%. The effects of these changes on the excess electricity production and/or the natural gas demand, and further on the MESP of the process were then analyzed.

RESULTS AND DISCUSSION

Base Case Economic Assessment

In the economic assessment of the organosolv process, the capital investment and variable and fixed costs as well as the co-product credits were assessed first. Based on these, the MESP was calculated to analyze the economic feasibility of the organosolv process. Table 4 presents the investment costs of installed equipment as well as the other direct costs, indirect costs, the cost of land, and working capital. The sizing and costing of individual pieces of equipment and process areas as well as the bases of costing are presented in more detail in the supplementary material (Supplementary Table 1).

Table 4. Investment Cost Analysis

Investment cost	**MUSD**	**Investment cost**	**MUSD**
Installed equipment cost		*Indirect costs*	
Process areas		Prorateable expenses (10% of TDC)	43
Feed handling	42		
Pretreatment and lignin recovery	88	Field expenses (10% of TDC)	43
Solvent and furfural recovery	31		
Conditioning of the hemicellulosic stream	19	Home office and construction (20% of TDC)	86
Acetic acid recovery	12		
Enzymatic hydrolysis and fermentation	28	Project contingency (10% of TDC)	43
Ethanol product and solids fermentation	18		
Wastewater treatment	54	Other costs (10% of TDC)	43
Storages	5	**Total indirect costs**	**257**
Boiler and turbine	82		
Utilities	7	**Fixed capital investment (FCI)**	**685**
Total installed equipment cost	**386**		
Other direct costs		Land	3
Warehouse (4% of ISBL)	10	Working capital (5% of FCI)	34
Site development (9% of ISBL)	21		
Additional piping (4.5% of ISBL)	11	**Total capital investment**	**722**
Total direct costs (TDC)	**428**		

As can be seen in Table 4, the estimated total capital investment was calculated to be 722 MUSD, resulting in a total capital investment per annual ethanol production of 13.4 USD/gal (4490 USD/MT). The organosolv process presented by van der Linden *et al.* (2012) apparently included size reduction, organosolv cooking, solvent, furfural, and lignin recovery as well as pulp washing but no major auxiliary processes. Scaling their capital cost (see *Introduction*) by a scaling exponent of 0.6 to the feedstock flow used in this study resulted in a total capital investment that is well in line with the direct and indirect capital costs (using the percentages for other direct costs and indirect costs presented in Table 4) of the sections feed handling, pretreatment and lignin recovery, and solvent and furfural recovery of this study (both approximately 300 MUSD).

The NREL dilute acid process (Humbird *et al.* 2011) was used in this work as a reference point for the studied organosolv process. The total capital investment of the NREL process indexed to 2013 from their cost year of 2007 would be approximately 458 MUSD. Not unexpectedly, a more complicated flowsheet of the organosolv process with the recovery of solvent and co-products would therefore make the investment costs of this process higher.

The annual variable and fixed costs, co-product revenues as well as the minimum ethanol sales revenue and the according MESP are presented in Table 5 (see Supplementary Tables 2 and 3 for more detailed breakdowns of the variable operating costs and revenues from by-products as well as of the MESP calculation). The minimum annual ethanol sales revenue was 166 MUSD/year. With an ethanol production of 459 MT/day (Kautto *et al.* 2013), this converted to an MESP of 3.07 USD/gal (approximately 1030 USD/MT).

Table 5. Annual Variable and Fixed Costs, Revenues, and MESP

Annual costs	**MUSD/a**	**Annual revenues**	**MUSD/a**
Variable costs		*Revenues from co-products*	
Feedstock	70.1	Organosolv lignin	49.0
Chemicals		Furfural	3.7
Cellulase	28.7	Acetic acid	6.4
Others	9.3	Electricity	1.6
Natural gas	4.1	**Total co-products revenue**	**60.6**
Ash disposal	0.8		
Fixed costs		*Revenues from ethanol*	
Salaries	2.7	**Minimum ethanol sales**	**165.5**
Labor burden (90% of salaries)	2.4		
Maintenance (3% of ISBL)	7.1	**MESP**	**3.07**
Property insurance	4.8		**USD/gal**
Total variable and fixed costs	**130.0**		

The major cost factors of the process were feedstock and enzyme costs, lignin co-product credit, and capital costs (Table 6).

Table 6. Breakdown of MESP to Cost Contributors

Cost contributor	**USD/gal**
Feedstock	1.30
Enzyme	0.53
Other chemicals & ash disposal	0.17
Natural gas	0.08
Co-products	
Organosolv lignin	-0.91
Furfural	-0.07
Acetic acid	-0.12
Electricity	-0.03
Fixed costs	0.32
Capital depreciation	0.42
Tax	0.24
Average return on investment	1.14
Total	**3.07**

Due to their smaller production rates, the sales of the other co-products, furfural and acetic acid, had a smaller effect on the MESP than lignin. By allocating the direct investment costs of the pieces of equipment directly related to the production of furfural and acetic acid as well as the indirect investment costs proportional to these direct investment costs, the direct variable operating costs, and the direct revenues to the furfural and acetic acid co-products, the payback period for these products would be approximately 0.2 and 5.2 years, respectively. This would indicate that the recovery of furfural as a side-draw from the ethanol recovery distillation columns and the further purification of furfural by distillation (Kautto *et al.* 2013) could be economically feasible. Due to the relatively low concentration (approximately 0.7 wt-%) of acetic acid in the evaporator condensates and its rather low price, the economic feasibility of recovering the acetic acid through extraction and distillation would be rather uncertain, especially taking into account that no TOPO (trioctyl phosphine oxide, a solvent used in the extraction of acetic acid) was assumed to be lost in the extraction (Kautto *et al.* 2013).

Aligning feedstock costs of the NREL dilute acid study (Humbird *et al.* 2011) with the current organosolv study and indexing other operating costs and investment costs to the year 2013 resulted in an MESP of the NREL process (approximately 2.6 USD/gal) that is relatively close to the base case MESP of the organosolv process (approximately 3.1 USD/gal). This demonstrates that the organosolv pulping process for the production of bioethanol could be competitive, with either a slightly higher price of lignin or lower investment or operating costs. The US market price of ethanol in December 2013 was approximately 2.0 USD/gal (Chicago Mercantile Exchange 2013), indicating that the MESP of the base case organosolv process would need to be somewhat lower or the market price of ethanol higher to make the process economically attractive (not taking into account any policy instruments to support lignocellulosic ethanol). The economic viability of the organosolv process will be further analyzed in the next section.

Sensitivity Analyses

Parameters used in the sensitivity analyses covered both technical and market-related aspects of the studied organosolv process. The technical parameters (steam and electricity consumption of the process, conversion rates in the enzymatic hydrolysis stage, and conversion rates of hemicellulosic sugars in the fermentation stage) and their expected variations are discussed in section *Technical parameters varied in sensitivity analyses*. The sensitivity of changes of these factors on the MESP is presented in Table 7. C5 sugars in the table refer to hemicellulosic pentoses (xylose and arabinose) and C6 sugars to hexoses (mannose and galactose). Also the effect of changes in the total capital investment and in the discount rate used in MESP calculation are presented in the table. The effects of lignin price and feedstock and enzyme costs are shown separately in Fig. 2.

Table 7. Sensitivity of MESP to Technical Assumptions

Design parameter	Base case	Value in sensitivity analysis	MESP, USD/gal
Steam and power consumption	6720 MT/day (steam), 670 MWh/day (power)	8740 MT/day, 870 MWh/day (+30%) 4710 MT/day, 470 MWh/day (-30%)	3.25 (+0.18) 2.91 (-0.17)
Conversion rate in the enzymatic hydrolysis	90% for cellulose, 82% for hemicelluloses	96% for cellulose, 88% for hemicelluloses (high) 80% for cellulose and hemicelluloses (low)	2.92 (-0.15) 3.38 (0.30)
Conversion rate of hemicellulosic sugars in the fermentation	85% for C5 sugars, 0% for C6 sugars	85% for C5 and C6 sugars (high) 50% for C5 sugars, 0% for C6 sugars (low)	2.88 (-0.19) 3.23 (0.16)
Total capital investment	722 MUSD	939 MUSD (+30%) 505 MUSD (-30%)	3.60 (+0.53) 2.54 (-0.53)
Discount rate used in MESP calculation	10 %	20 %	4.30 (+1.23)

As can be seen in Table 7, the MESP was very sensitive to changes in investment costs. Decreasing the total capital investment by 30% decreased the MESP by 0.53

USD/gal, illustrating that decreases in capital costs could improve the economics of the process considerably. In the base case scenario the plant was assumed to be a greenfield project with no existing infrastructure or utility systems. Integrating the plant into an existing industrial facility could be a way to decrease the investment costs considerably, for example by making investments in the boiler and turbine units and wastewater treatment unnecessary. Smaller capital savings could be gained by omitting the investment in acetic acid recovery, which was found to have a relatively long payback period.

Similarly to the absolute level of investment, also the required rate of return on the invested capital (discount rate) had a considerable effect on the MESP. Doubling the discount rate from 10% to 20% increased the MESP by 1.23 USD/gal. Taking into account the risk associated with novel technology and a novel product (organosolv lignin) that have not been demonstrated in large scale, the discount rate might need to be higher than 10%.

Also, changes in the studied technical parameters had an effect on the MESP, with changes in the MESP being in the range of -0.19 to 0.30 USD/gal. The conversion rate both in hydrolysis and fermentation had a relatively considerable effect on the MESP, demonstrating the importance of the overall conversion rate from carbohydrates to ethanol on the economics of the process.

As discussed in the *Introduction*, one of the main arguments for an organosolv pretreatment of lignocellulosic biomass prior to enzymatic hydrolysis is that a relatively pure lignin fraction can be produced. As seen in Tables 5 and 6, with an assumed base case lignin price of 450 USD/MT, the lignin sales are a significant source of revenue and, consequently, a major contributor in the MESP.

The effect of lignin sales on the overall economics of the organosolv process depends considerably on the assumed lignin price. Some price estimates have been presented in the literature (Baker 2010; Pye 2010a; Gosselink 2011; Rushton 2012; van der Linden *et al.* 2012). As there is however no existing commercial, larger scale production of organosolv lignin, the estimates presented in the literature have to be considered as only indicative. The base case lignin price was set in this study at the lower end of the price range presented in the literature. In addition to literature estimates, some indication of the price level could be found from the possible end-products produced from the lignin.

A wide scale of potential application areas have been identified for lignins in general. Holladay *et al.* (2007) classified the various uses into three categories: power-fuel-gasification (utilizing lignin as a carbon source in energy applications such as heat and power and synthesis gas derived fuels), macromolecules (utilizing lignin’s macromolecular nature for example in carbon fibers, adhesives, and resins), and aromatic chemicals (breaking up lignin’s macromolecular structure to aromatic monomers, including BTX chemicals, (benzene, toluene, xylene) and phenol). Ragauskas *et al.* (2014) reviewed challenges and recent advancements in the processing and use of lignin in materials (carbon fibers, plastics, and composites) and fuels and chemicals. Gosselink (2011) ranked lignin applications based on their value and market volume, from very high-volume, low-value applications (such as energy and fuels) to medium-scale, higher-valued applications (such as aromatic chemicals and phenolic resin) and further to very low-volume, very high-value applications (carbon fibers, fine chemicals). End-product prices ranging from a few hundred dollars per ton and below for energy applications to several thousands for vanillin and phenol derivatives (Gosselink 2011) illustrate the wide price range that could be adopted also for the organosolv lignin.

For organosolv lignins, uses especially in various resin systems such as phenolic and epoxy resins and polyurethanes have been suggested (Pye 2010b; CIMV 2013b). Within these resin systems, applications demonstrated either in pilot or larger scale include the partial replacement of phenol-formaldehyde resin as well as methylene diphenyl diisocyanate (used in polyurethane manufacture) by organosolv lignin in wood products (oriented strand board and medium-density fiberboard), partial replacement of phenol-formaldehyde resin in friction binders and rigid foam insulation, and the partial replacement of epoxy resins in coatings (Rushton 2012). These resins are relatively high-valued products, with the July 2014 prices being approximately 1800 USD/MT for phenolic, 2300 USD/MT for epoxy, and 2500 USD/MT for polyurethane resins (Plastics News 2014). The market volumes of phenolic and polyurethane resins in turn are large with above 10 million MT (MarketsandMarkets 2011; Research and Markets 2012), while the epoxy resin markets is somewhat smaller at approximately 2 million MT (Research and Markets 2013). The larger phenolic and polyurethane markets specifically, both of which are expected to grow in the near term future (MarketsandMarkets 2011; Research and Markets 2012), could be assumed large enough to absorb the amounts of lignin produced in the studied organosolv process (109,000 MT/year) with no major negative effect on the resin prices.

To study the effect of lignin price on the MESP further, the lignin price was varied within a price range. The minimum price of lignin can be considered to be its value as a fuel. As a very conservative fuel pricing scenario, the price of lignin can be calculated from the price of coal. Taking into account differences in the heating values (approximately 25 MJ/kg for lignin and 29 MJ/kg for coal), a coal price of 62.0 USD/MT (U.S. Energy Information Administration 2013c) converts to a lignin fuel value of approximately 53 USD/MT. More favorable fuel price references for lignin would be pellets and forest residues. The maximum price of lignin in the sensitivity analysis was set based on its use in phenolic resins. This use was highlighted by Stewart (2008), who argued that the application of lignin as a substitute for phenol in phenolic resins is a credible option due to the large and growing size of the phenol market and legislative reasons. In phenolic resins lignin can act both as a direct replacement of the resin itself (when lignin is blended with the resin) or as a replacement of the phenol chemical (when lignin is used as a phenolic component in the manufacture of the resin) (Pye 2010a). In this sensitivity analysis the reference product was assumed to be the phenol chemical. Although using organosolv lignin as a substitute for phenol would not directly mean that lignin would be sold for the same price as phenol, not least because modification may be required to enhance the reactivity of the lignin (Stewart 2008; Gosselink 2011), the price of phenol was used here as the optimistic, upper end of the price range used in the sensitivity analysis. In February 2013 phenol sold for approximately 1500 USD/MT (Dietrich 2013). Although higher-value uses than phenolic resins could be found, the average sales price of larger scale lignin production as outlined in this study was assumed to fall within the price range presented above.

Figure 2 shows the results of this analysis. The lignin prices corresponding to the indexed and feedstock cost adjusted MESP of the NREL process (Humbird *et al.* 2011) and the December 2013 market price (Chicago Mercantile Exchange 2013) are presented in the figure. In addition to the effect of lignin price on the base case MESP, also the effects of feedstock and enzyme costs are presented in the figure.

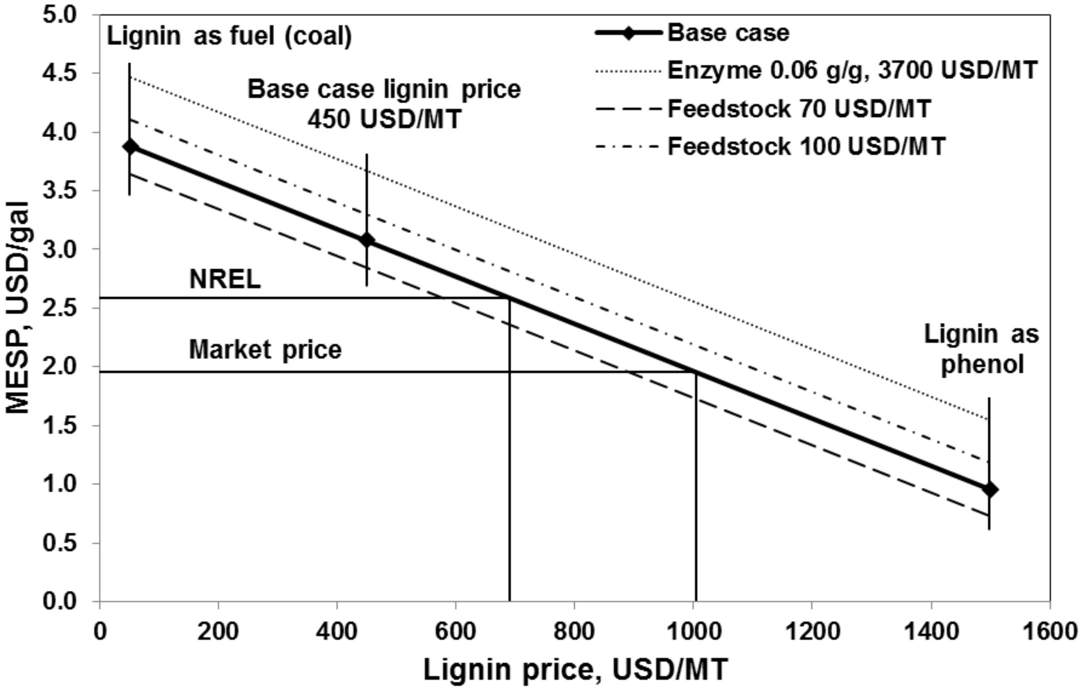

Fig. 2. The effect of lignin selling price on the MESP. Market price in the figure refers to the December 2013 market price of ethanol (Chicago Mercantile Exchange 2013) and NREL to the MESP of the NREL study (Humbird *et al.* 2011), indexed to 2013 and feedstock cost adjusted. The figure also presents the effect of enzyme and feedstock costs on the MESP.

It is apparent from Fig. 2 that under a very optimistic scenario with lignin sold at the 2013 phenol price, the MESP would be slightly below 1.0 USD/gal, making the process very attractive. Under the most pessimistic scenario, with lignin priced at its fuel value (with coal as the very conservative price reference), the MESP would be approximately 3.9 USD/gal, considerably higher than the MESP of the NREL dilute acid process (Humbird *et al.* 2011), with feedstock cost aligned with the current study and other costs indexed to 2013 as well as the December 2013 market price of ethanol (Chicago Mercantile Exchange 2013). This result, unsurprisingly, indicates that unless higher-valued applications than combustion can be found for lignin, the organosolv process is not competitive with the NREL dilute acid pretreatment process (Humbird *et al.* 2011) nor could it be used to produce ethanol profitably at the market price (not taking into account any policy instruments to support lignocellulosic ethanol). The lignin prices corresponding to the MESP of the NREL dilute acid process and December 2013 market price are 690 USD/MT and 1000 USD/MT, respectively. The lignin price of 1000 USD/MT is close to the prices assumed by Baker (2010) (below 1100 USD/MT) and van der Linden *et al.* (2012) (975 USD/MT). These higher prices could be realizable if, for example, part of the produced lignin could be sold as a phenol substitute.

In addition to presenting the effect lignin price on the base case MESP, Fig. 2 also presents the effect of variations in feedstock and enzyme costs on the base case MESP. Feedstock and enzyme made a major contribution to the MESP (Table 6). This also resulted in a relatively high sensitivity of the MESP to variations in these cost items. Changing the feedstock price from the base case value of 85 USD/dry MT by 15

USD/dry MT (70 or 100 USD/dry MT) changed the MESP by 0.23 USD/gal. Changing the enzyme loading and price from the base case values of 0.02 g enzyme/g cellulose and 5270 USD/MT to a more conservative 0.06 g enzyme/g cellulose and 3700 USD/MT (Novozymes 2011) increased the MESP by 0.59 USD/gal.

In commercializing the organosolv process, the sensitivity of the MESP to changes in the lignin price presented in Fig. 2 makes a strong case for market creation for the lignin product. The figure also demonstrates the importance of enzyme development and securing of a low-cost feedstock on the viability of this or any other enzymatic bioethanol production process.

CONCLUSIONS

1. In the previous study (Kautto *et al.* 2013), the organosolv process was found to have higher energy consumption and lower ethanol yield than the dilute acid process due to the recovery of solvent and lower carbohydrate recovery in pretreatment. In the current study, the increased number of processing steps was further found to increase the investment cost.
2. A major advantage of the organosolv process is that it enables the production of a relatively pure organosolv lignin fraction. Also other co-products, namely furfural and acetic acid can be produced. A larger fraction of the incoming raw material is therefore converted to products than in the dilute acid process.
3. The price of lignin defines to a considerable degree whether the revenue from the recovery of the co-products offsets the decreased ethanol yield, poorer energy economy, and higher investment cost. As there is no commercial scale production of organosolv lignin, estimating its possible market price is difficult.
4. A wide range of possible lignin prices was therefore studied in sensitivity analyses. At a base case lignin price of 450 USD/MT, the MESP of the organosolv process was found to be 3.1 USD/gal, being higher than that of the dilute acid process, which was used as a reference.
5. A lignin price of 690 USD/MT was required to make the ethanol production of the organosolv process cost competitive with that of the dilute acid process, illustrating the importance of finding higher value markets for lignin to make the organosolv process competitive.

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APPENDIX: SUPPLEMENTARY MATERIAL

Supplementary Table 1. Equipment Sizing and Costs (EC Refers to Equipment Cost, IF to Installation Factor, IC to Installed Cost, CB to Costing Basis, SE to Scaling Exponent, [1] to Industry Estimate, [2] to Aspen Process Economic Analyzer (AspenTech 2011), [3] to Humbird *et al.* (2011), [4] to Aden *et al.* (2002)). All Costs in MUSD. The Names of the Pieces of Equipment Refer to the Names Used in the Process Flowsheets Presented in the Technical Analysis of the Process (Kautto *et al.* 2013)

Name	Material	Qty	Size	EC	IF	IC	CB	SE	Remarks
Feed handling						**41.7**	[1]		
Pretreatment and lignin recovery									
Digester & washing						73.8	[1]		
Flash tank I	SS316	1	D 2.6 m, h 2.6 m, V 41.0 m^3	0.4	2.0	0.9	[2]		5 min liquid hold-up
Flash tank II	SS316	1	D 2.4 m, h 7.5 m, V 34.9 m^3	0.1	2.0	0.3	[2]		5 min liquid hold-up
Flash tank III	SS316	1	D 2.4 m, h 7.2 m, V 33.5 m^3	0.1	2.0	0.2	[2]		5 min liquid hold-up
Post-hydrolysis reactor	SS316	1	D 5.5 m, h 16.6 m, V 393 m^3	1.2	2.0	2.4	[2]		Costed as a vertical tank, residence time 60 min
Agitator	SS316	1		0.1	1.5	0.1	[3]	0.5	
Dilution & mixing tank	SS316	1	D 4.0 m, h 13.7 m, V 169 m^3	0.6	2.0	1.2	[2]		Costed as an agitated tank, residence time 10 min
Lignin filter	SS316	1		2.2	1.7	3.8	[3]	0.8	Pressure filter, scaled based on solids flow
Filtration auxiliary equipment				0.6	1.5-3.1	1.3	[3]	0.5-0.8	Pumps, tanks, compressors, agitator, cake screw & conveyor
Lignin spray dryer	CS	2		1.3	1.7	2.3	[2]		
Screen	SS	1		0.7	1.7	1.2	[2]		Costed as a screen and dewatering unit
Heat exchanger I	SS316;SS	1	Area 170 m^2	0.1	2.2	0.1	[2]		Shell and tube (S&T), overdesign factor (OF) 1.15
Heat exchanger II	SS316;SS	1	Area 80 m^2	0.0	2.2	0.1	[2]		S&T, OF 1.15
Heat exchanger III	SS304;SS316	1	Area 530 m^2	0.2	2.2	0.4	[2]		S&T, OF 1.15
Pumps		7				0.4			OF 1.10
Total process area						**88.5**			

Name	Material	Qty	Size	EC	IF	IC	CB	SE	Remarks
Solvent and furfural recovery									
Recovery column I									
Tower	SS316	1	D 5.5 m, h 25.0 m, actual number of trays 35	2.2	2.4	5.3	[2]		Tray efficiency 70%
Condenser	SS316; SS316	1	Area 3900 m^2	1.8	2.4	4.4	[2]		
Reboiler	SS316; SS316	1	Area 1910 m^2	1.3	2.4	3.2	[2]		
Recovery column II									
Tower	SS316	1	D 8.2 m, h 25.0 m, actual number of trays 35	3.8	2.4	9.1	[2]		Tray efficiency 70%
Condenser	SS316;CS	1	Area 10100 m^2	2.5	2.4	6.1	[2]		
Reboiler	SS316; SS316	1					[2]		Condenser of Recovery column I
Furfural distillation									
Tower	SS316	1	D 0.3 m, h 13.4 m, actual number of trays 16	0.1	2.4	0.2	[2]		Tray efficiency 70%
Reboiler & condenser	SS316;CS	1	Condenser 1.5 m^2, reboiler 6.1 m^2	0.1	2.4	0.2	[2]		
Decanter I	SS304		D 0.9 m, h 2.7 m, V 1.8 m^3	0.0	2.0	0.05	[2]		Costed as a vertical tank, residence time 5 min
Heat exchanger IV	SS304;SS316	1	Area 1250 m^2	0.5	2.2	1.1	[2]		S&T, OF 1.15
Heat exchanger V	SS316;CS	1	Area 10 m^{2*}	0.0	2.2	0.05	[2]		S&T, OF 1.15
Heat exchanger VI	SS304;CS	1	Area 5 m^{2*}	0.0	2.2	0.05	[2]		S&T, OF 1.15
Heat exchanger VII	SS304;SS316	1	Area 10 m^{2*}	0.0	2.2	0.05	[2]		S&T, OF 1.15
Heat exchanger VIII	SS316;SS	1	Area 850 m^2	0.3	2.2	0.6	[2]		S&T, OF 1.15
Pumps		5				0.2	[2]		OF 1.1
Total process area						**30.7**			
Conditioning of the hemicellulosic sugar stream									
Evaporation									
Effects	SS304	4	Total area 15600 m^2	4.3	2.1	9.1	[4]	0.7	Scaled based on area
Auxiliaries	SS304			0.8		1.9	[4]		Pumps, condenser

Name	**Material**	**Qty**	**Size**	**EC**	**IF**	**IC**	**CB**	**SE**	**Remarks**
Separation of tarry lignin	SS304	1	Decanter 1.5 m, L 6.2 m, V 23.5 m^3	0.1		0.2	[2], [3]	0.8	Decanter (costed as a horizontal vessel), screw, conveyor
LMW lignin separation									
Mixers	SS304	4	D 1.7 m, h 1.7 h, V 3.8 m^3	0.2	1.7	0.4	[2]		Agitated vessel (5 min)
Settlers	SS304	4	D 1.0 m, L 3.9 m, V 2.8 m^3	0.1	1.7	0.2	[2]		Horizontal vessel (5 min)
Furfural vacuum distillation									
Tower	SS316	1	D 3.4 m, h 6.7 m, actual number of trays 6	0.4	2.4	1.1	[2]		Tray efficiency 70%
Condenser	SS316;CS	1	Area 870 m^2	0.5	2.4	1.2	[2]		
Reboiler	SS316;CS	1	Area 1240 m^2	0.6	2.4	1.3	[2]		
Steam stripping	SS316	1	D 2.9 m, h 9.1 m, actual number of trays 9	0.4	2.4	0.9	[2]		Tray efficiency 70%
Neutralization	SS304;SS	1		0.1		0.2	[3]	0.5-0.7	Tank, agitator, mixer, scaled on mass flow
Decanter II	SS304	1	D 1.5 m, h 4.4 m, V 8.1 m^3	0.0	2.0	0.1	[2]		Costed as a vertical tank, residence time 5 min
Heat exchanger IX	SS304;CS	1	Area 30 m^2	0.0	2.2	0.1	[2]		S&T, OF 1.15
Heat exchanger X	SS316;SS316	1	Area 1020 m^2	0.5	2.2	1.0	[2]		S&T, OF 1.15
Heat exchanger XI	SS304;CS	1	Area 70 m^2	0.0	2.2	0.1	[2]		S&T, OF 1.15
Heat exchanger XII	SS316;SS304	1	Area 980 m^2	0.4	2.2	0.9	[2]		S&T, OF 1.15
Pumps		4				0.1	[2]		OF 1.1
Total process area						**18.7**			
Acetic acid recovery									
AA extraction column									
Mixers	SS316	4	D 3.6 m, h 3,6 m, V 37.1 m^3	1.0	1.7	1.7	[2]		Agitated vessel (5 min)
Settlers	SS316	4	D 3.0 m, L 12.1 m, V 86.2 m^3	0.5	1.7	0.9	[2]		Horizontal vessel (5 min)
AA distillation column I									
Tower	SS316L	1	D 1.4 m, h 6.7 m, actual number of trays 5	0.1	2.4	0.3	[2]		Tray efficiency 70%
Reboiler & condenser	SS316L;CS	1	Condenser 6 m^2, reboiler 180 m^2	0.1	2.4	0.1	[2]		Includes a decanter

Name	Material	Qty	Size	EC	IF	IC	CB	SE	Remarks
AA distillation column II									
Tower	SS316L	1	D 6.6 m, h 11.6 m, actual number of trays 13	1.5	2.4	3.5	[2]		Tray efficiency 70%
Reboiler & condenser	SS316L;CS	1	Condenser 610 m^2, reboiler 180 m^2	0.4	2.4	0.9	[2]		Includes a decanter
AA distillation column III									
Tower	SS316L	1	D 1.2 m, h 46.3 m, actual number of trays 70	1.2	2.4	3.0	[2]		Tray efficiency 70%
Reboiler & condenser	SS316L;CS	1	Condenser 33 m^2, reboiler 56 m^2	0.1	2.4	0.2	[2]		
Decanter III	SS304	1	D 0.8 m, h 2.4 m, V 2.4 m^3	0.0	2.0	0.1	[2]		Costed as a vertical tank (5 min)
Heat exchanger XIII	SS316;SS316	1	Area 780 m^2	0.4	2.2	0.9	[2]		S&T, OF 1.15
Heat exchanger XIV	SS316;SS316	1	Area 10 m^2*	0.0	2.2	0.1	[2]		S&T, OF 1.15
Heat exchanger XV	SS316;SS316	1	Area 20 m^2*	0.0	2.2	0.1	[2]		S&T, OF 1.15
Pumps		3				0.2	[2]		OF 1.1
Total process area						**11.9**			
Enzymatic hydrolysis and fermentation									
Enzyme-hydrolyzate mixer	SS316	1				0.2	[3]	0.5	Scaled on the hydrolyzate mass flow
Hydrolysis tanks	SS304					6.4	[3]	0.7	Same scaling basis as above
Ethanol fermentation tanks						15.1	[3]	0.7	Scaled on beer flow leaving fermentation, includes coolers and agitators
Beer storage tank	SS316	1				1.2	[3]	0.5-0.7	Scaled on beer leaving beer tank, includes an agitator
Seed fermentation tanks	SS304					3.5	[3]	0.5-0.7	Scaled on inoculum flow, includes coils and agitators.
Seed hold tank	SS316	1				0.8	[3]	0.5-0.7	Scaled on inoculum flow, includes an agitator.
Heat exchanger XVI	SS;CS	1	Area 230 m^2	0.1	2.2	0.1	[2]		S&T, OF 1.15
Pumps		5				0.3	[3]	0.8	OF 1.1
Total process area						**27.7**			

Name	Material	Qty	Size	EC	IF	IC	CB	SE	Remarks
Ethanol product and solids recovery									
Beer & rectification columns	SS316;CS	1				8.2	[3]	0.6	Includes reboilers, beer column condenser, and feed heat exchanger, scaled on EtOH flow to molecular sieves
Water scrubber	SS304;PP	1				0.5	[3]	0.6-0.8	Scaled on the total vents flow, includes a pump
Molecular sieve package	SS	1				4.7	[3]	0.6	Scaled on the EtOH product stream
Insolubles filter	SS316	1				3.7	[3]	0.8	Scaled on the insolubles stream
Auxiliary equipment						1.3	[3]	0.6-0.8	Auxiliaries include pumps, tanks, compressors, agitator, cake screw, and conveyor, scaled on the insolubles stream
Total process area						**18.4**			
Wastewater treatment						**53.5**	[3]		Not scaled, assumed similar to [3]
Storage						**5.4**	[3]		Not scaled, similar to [3]
Boiler and turbine						**82.3**	[3]	0.6	Scaled on heat load
Utilities									
Cooling tower system						3.4	[3]	0.6	Scaled on cooling duty
Chilled water package						1.8	[3]	0.6	Scaled on cooling duty
Plant/instrument air						0.1	[3]		Not scaled, similar to [3]
Process water system						0.6	[3]		Not scaled, similar to [3]
Sterile water & CIP/CS systems						1.4	[3]		Not scaled, similar to [3]
Total process area						**7.3**			
Total installed equipment cost						**386**			

*****Small heat exchangers that could possibly be omitted

Supplementary Table 2. Variable Operating Costs and Revenues from By-products

	MT/hr	**Price, USD/MT**	**Source of the price**	**Cash flow, MUSD/a**	**Remarks**
Variable costs					
Feedstock (as dry)	98.0	85	Berguson *et al.* 2010	70.1	Debarked hardwood 83.3 MT/hr, bark and losses 14.7 MT/hr
Chemicals					
Cellulase	0.6	5270	Novozymes 2011	28.7	
Sulfuric acid	0.8	130	Humbird *et al.* 2011	0.9	
Ammonia	0.5	590	Humbird *et al.* 2011	2.6	
Corn steep liquor	1.1	75	Humbird *et al.* 2011	0.7	
Diammonium phosphate	0.1	1300	Humbird *et al.* 2011	1.5	
Sorbitol	0.03	1490	Humbird *et al.* 2011	0.4	Scaled based on flow to the seed train (Humbird *et al.* 2011)
Caustic	1.1	200	Humbird *et al.* 2011	1.9	Scaled based on ammonia usage (Humbird *et al.* 2011)
Flue gas desulfurization lime	0.4	270	Humbird *et al.* 2011	0.9	Scaled based on sulfuric acid usage (Humbird *et al.* 2011)
Cooling tower chemicals	0.002	4040	Humbird *et al.* 2011	0.1	Usage assumed similar to Humbird *et al.* (2011)
Make-up water	147.1	0.22	Humbird *et al.* 2011	0.3	
Kerosene	0.005	1090	Dennis *et al.* 2013	0.05	
Total chemicals				37.9	
Ash disposal	2.6	36	Humbird *et al.* 2011	0.8	
Natural gas	2.3	4.17 (USD/GJ)		4.1	Source U.S. Energy Information Administration 2013a
Total variable costs				**112.9**	
Revenues from by-products					
Organosolv lignin	12.9	450	Estimate	49.0	
Furfural	0.3	1600	Arato *et al.* 2005	3.7	Net furfural production, taking the make-up furfural required in the furfural extraction into account
Acetic acid	1.3	600	Kelley 2013	6.4	
Electricity	3.6 (MW)	52.2 (USD/MWh)		1.6	Source U.S. Energy Information Administration 2013b
Total by-product revenues				**60.6**	

Supplementary Table 3. Minimum Ethanol Selling Price (MESP) Calculation (in MUSD)

Year	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**
Fixed capital investment	21.9	164.3	87.6														
Land	3.1																
Working capital			34.2														
Loan payment				61.2	61.2	61.2	61.2	61.2	61.2	61.2	61.2	61.2	61.2				
Loan interest payment	2.6	22.3	32.9	32.9	30.6	28.1	25.5	22.6	19.6	16.2	12.6	8.7	4.5				
Loan principal	32.9	279.3	410.8	382.4	351.8	318.7	283.0	244.4	202.8	157.8	109.2	56.7	0.0				
Ethanol sales				144.8	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5
Lignin sales				42.8	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0
Other sales				10.2	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7
Total sales				**197.9**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**
Annual manufacturing costs																	
Feedstock				61.3	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1
Cellulase				26.9	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7
Natural gas				3.8	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1
Baghouse bags				0.7					0.7					0.7			
Other variable costs				9.4	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0
Fixed costs				17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1
Total product costs				**119.2**	**130.0**	**130.0**	**130.0**	**130.0**	**130.6**	**130.0**	**130.0**	**130.0**	**130.0**	**130.6**	**130.0**	**130.0**	**130.0**
Annual depreciation																	
General plant																	
DDB				172.1	122.9	87.8	62.7	44.8	29.9	14.9							
SL				86.0	71.7	61.5	54.9	52.3	52.3	52.3							
Remaining value				602.3	430.2	307.3	219.5	156.8	104.5	52.3							
Actual				**172.1**	**122.9**	**87.8**	**62.7**	**52.3**	**52.3**	**52.3**							
Steam plant																	
DDB				6.2	5.7	5.3	4.9	4.5	4.2	3.9	3.6	3.3	3.0	2.8	2.5	2.2	1.9
SL				4.1	4.0	3.9	3.8	3.8	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7

Remaining value				82.3	76.2	70.5	65.2	60.3	55.8	51.6	47.7	44.0	40.4	36.7	33.0	29.4	25.7
Actual				**6.2**	**5.7**	**5.3**	**4.9**	**4.5**	**4.2**	**3.9**	**3.7**	**3.7**	**3.7**	**3.7**	**3.7**	**3.7**	**3.7**
Net revenue				-132.5	-63.1	-25.1	3.1	16.7	19.5	23.8	79.9	83.8	87.9	91.8	92.5	92.5	92.5
Losses forward					-132.5	-195.5	-220.6	-217.5	-200.8	-181.3	-157.5	-77.6	0.0	0.0	0.0	0.0	0.0
Taxable income				-132.5	-195.5	-220.6	-217.5	-200.8	-181.3	-157.5	-77.6	6.1	87.9	91.8	92.5	92.5	92.5
Income tax				0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	2.1	30.8	32.1	32.4	32.4	32.4
Annual cash income				17.4	34.9	34.9	34.9	34.9	34.3	34.9	34.9	32.8	4.2	63.4	63.8	63.8	63.8
PV of annual CF				13.1	23.9	21.7	19.7	17.9	16.0	14.8	13.5	11.5	1.3	18.4	16.8	15.3	13.9
Total capital investment + interest	27.6	169.7	127.9														

Year	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**	**28**	**29**	**30**	**31**	**32**	**33**
Fixed capital investment																
Land																-3.1
Working capital																-34.2
Loan payment																
Loan interest payment																
Loan principal																
Ethanol sales	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5	165.5
Lignin sales	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0	49.0
Other sales	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7	11.7
Total sales	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**	**226.1**
Annual manufacturing costs																
Feedstock	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1	70.1
Cellulase	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7	28.7
Natural gas	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1	4.1
Baghouse bags		0.7					0.7					0.7				

Other variable costs	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.0
Fixed costs	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1	17.1
Total product costs	**130.0**	**130.6**	**130.0**	**130.0**	**130.0**	**130.0**	**130.6**	**130.0**	**130.0**	**130.0**	**130.0**	**130.6**	**130.0**	**130.0**	**130.0**	**130.0**
Annual depreciation																
General plant																
DDB																
SL																
Remaining value																
Actual																
Steam plant																
DDB	1.7	1.4	1.1	0.8	0.6	0.3										
SL	3.7	3.7	3.7	3.7	3.7	3.7										
Remaining value	22.0	18.4	14.7	11.0	7.3	3.7										
Actual	**3.7**	**3.7**	**3.7**	**3.7**	**3.7**	**3.7**										
Net revenue	92.5	91.8	92.5	92.5	92.5	92.5	95.5	96.2	96.2	96.2	96.2	95.5	96.2	96.2	96.2	96.2
Losses forward	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Taxable income	92.5	91.8	92.5	92.5	92.5	92.5	95.5	96.2	96.2	96.2	96.2	95.5	96.2	96.2	96.2	96.2
Income tax	32.4	32.1	32.4	32.4	32.4	32.4	33.4	33.7	33.7	33.7	33.7	33.4	33.7	33.7	33.7	33.7
Annual cash income	63.8	63.4	63.8	63.8	63.8	63.8	62.1	62.5	62.5	62.5	62.5	62.1	62.5	62.5	62.5	62.5
PV of annual CF	12.6	11.4	10.4	9.5	8.6	7.8	6.9	6.3	5.8	5.2	4.8	4.3	3.9	3.6	3.3	3.0
Total capital investment + interest																

Discount rate 10%, interest on loan 8%, payback period for loan 10 years, income tax 35%, net present value (of annual CF) 0.0, MESP 3.07 USD/ga