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Development and economic analysis of bioethanol production facilities using lignocellulosic biomass

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An integrated process for bioethanol production from *Miscanthus sacchariflorus* was used to construct a bench-scale plant constructed and an economic analysis was carried out to investigate the feasibility of its application to a commercial plant. The bench-scale plant was operated for 1 month and an economic analysis and sensitivity analysis was performed on the data acquired. In this study, 100,000 kL of bioethanol could be produced annually from 606,061 tons of *M. sacchariflorus* and the production cost was calculated to be US$1.76/L. However, the by-products of this process such as xylose molasses and lignin can be sold or used as a heat source, which can decrease the ethanol production costs. Therefore, the final ethanol production cost was calculated to be US$1.31/L, and is considerably influenced by the enzyme cost. The results and data obtained should contribute to the development of a commercial-scale lignocellulosic bioethanol plant.

[**Key words:** *Miscanthus sacchariflorus*; Bioethanol; Bench-scale plant; Scale-up; Economics]

Plant cell wall material (lignocellulosic biomass) is the most abundant renewable resource on Earth. Worldwide, plants fix $56x10^9$ tons of CO_2 per year and produce $170–200x10^9$ tons of biomass (1–3). Plant cell walls are composed of cellulose, hemicellulose and lignin, and these three components account for more than 70% of their composition. Cellulose and hemicellulose can be broken down into monomer sugars and can be converted to biofuels or other valuable chemicals by biological and chemical reaction (4,5). Lignocellulosic biomass-based bioethanol production has increased rapidly (6) as an alternative to fossil fuels, which are associated with environmental problems such as air pollution and climate change. Furthermore, bioethanol is compatible with current transportation infrastructure, so it could be integrated easily into existing fuel distribution networks. Additionally, it is not limited by the availability of natural resources (7). As well as having these advantages, bioethanol is emerging as a promising option for coping with climate change and diminishing oil supply (8).

Bioethanol is commercially produced by fermenting monomer sugar extracted from starch-based or sugar-based feedstock such as corn, cassava, grain or sugar cane. However, this has raised ethical and social concerns regarding food or feed usage (9). For this reason, a lot of research groups and companies have been attempting to develop the technology for producing bioethanol from lignocellulosic biomass commercially. Demonstration and commercial plants for producing bioethanol from lignocellulosic biomass have been researched and developed for a long time (10–13). However, acid-hydrolysis has a limited application to large-scale plants using integrated processes because of acid corrosion (14).

Worldwide, several lignocellulosic biofuel plants are under demonstration or operating in the U.S., Canada, China, and several European countries. All biofuel plants are trying to develop commercial-scale biofuel production, and all bioethanol companies located in the U.S. are demonstrating or operating to meet the federal renewable fuel standard (15). Despite the global recession, the biofuel industry now has facilities and projects under development in more than 20 U.S. states, representing billions of dollars in private investment. In Ottawa, Canada, the logen demonstration fuel production plant has produced 1 million gallons per year of bioethanol from cereal straw, bagasse, corn stover, and grasses since 2005. Beta Renewable began the operation of a commercial facility in Crescentino, Italy in 2012. This facility uses a mix of wheat straw, rice straw, bagasse, *Arundo donax*, corn stover, and poplar as feedstock and is producing 20 million gallons per year of ethanol (15).

The purpose of this work was to design a bench-scale bioethanol production plant using lignocellulosic biomass from *Miscanthus sacchariflorus*. All processes, such as pretreatment, hydrolysis, fermentation, and purification were included in the integrated process system. Moreover, possibility and operation result of the integrated process were investigated through the operation of the bench-scale bioethanol production plant for collecting scale-up factor. Finally, economic analysis was carried out to examine the potential to develop this process at commercial scale.

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MATERIALS AND METHODS

Materials *M. saccharifloru*s, genotype Geodaeuksae 1, grown and collected by the Bioenergy Crop Research Center (Muan, Korea), was used in this work. It was milled by cutter mill to a particle size of less than 2 mm, and then kept in one-ton bag at room temperature. The *M. saccharifloru*s consisted of 37.1% cellulose, 21.3% hemicellulose, 23.2% lignin, and 3.1% ash. The remaining 15.3% included some organic compounds, such as uronic acid and acetyl groups, and other trace components, including minerals, waxes, fats, starches, resins, and gums (14). Enzymes Cellic CTec 2 and Cellic HTec 2 (Novo Inc., Bagsværd, Denmark) were used in a simultaneous saccharification and fermentation (SSF) process. All reagents used in this study were of analytical grade, except for the sodium hydroxide (Duksan Chemical Co., Ltd., Ansan, Korea), which was of industrial grade. The yeast used in the present study was *Saccharomyces cerevisiae* CHY1011, which was isolated and identified by Changhae ethanol (Changhae Ethanol Co., Ltd., Jeonju, Korea) as an industrial strain (16).

Process description Fig. 1 presents a process flow diagram of the bench-scale plant for producing bioethanol from lignocellulosic biomass, and Table S1 shows the equipment list and specification. The bench-scale plant was scaled up through previous studies and its capacity was 100 kg (dry weight) of biomass per day (17,18). The bench-scale plant is composed of the following sections: (i) pretreatment process, (ii) hydrolysis and fermentation process, and (iii) purification process. In order to operate automatically, a programmable logic controller (PLC) system and human machine interface were installed.

Pretreatment process The pretreatment stage includes the biomass size reduction process, storage and feeding system, pretreatment process, and the dewatering, washing and supply system. A cutter mill was used in the size reduction process, and the storage and feeding process is composed of a two-stage system. The capacity of the first stage hopper is 250 kg of biomass and that of the second stage hopper is 5 kg, which allows 100 kg to be processed per day. A twin-screw extruder which is used normally in the food and polymer industry (Changhae ethanol multi extruder, CHEMET) was used in the pretreatment process and the temperature of the pretreatment reactor was controlled by steam. The CHEMET pretreatment was operated at the optimum condition: 95°C, 0.4 M sodium hydroxide, 80 rpm twin-screw speed, a flow rate of 120 mL/min, and solid:liquid ratio = 1:8 (17).

After the pretreatment process, solid/liquid (S/L) separation was conducted by a dewatering, washing and supply system (Changhae ethanol scant reagent feeder, CHESAF). The CHESAF system uses a single screw, and was divided into four stages as follows: the first dewatering stage, the washing stage, the second dewatering stage, and the supply stage. In the first dewatering stage, the pretreatment solvent was isolated from the pretreated biomass. Separated solvent was collected in the waste solvent tank, and then transferred to the solvent recovery system. The solid fraction was washed in the washing stage and the washing water was finally separated in the second dewatering step. CHESAF was connected to the bottom of the hydrolysis and fermentation tank.

The waste solvent from the pretreatment process was recovered by the solvent recovery system. After the pretreatment process, the waste solvent was separated by CHESAF and collected in the waste solvent tank. The waste solvent contained lignin, lipid, and a small amount of pretreated biomass. The waste solvent was then centrifuged and fine solids were removed using a 1 μm membrane filter. Finally, the solvent was recovered using a small pore ultrafiltration membrane and reused.

Hydrolysis and fermentation process The hydrolysis and fermentation process was adapted with a continuous feeding system (CHESAF process) and a particular type of impeller developed by our research team for producing high concentration bioethanol. A paddle and anchor-type agitator was included in the impeller and paddle was angled at 45° in order to stir the mixture well (18).

This bioethanol production facility included a hydrolysis/fermentation tank and buffer tank made from STS304. Volume of HF tank and buffer tank were 200 L and 800 L, respectively. In order to sterilize the reactor, a hydrolysis and fermentation tank (HF tank) was used. The HF tank could resist maximum temperature and pressure of 120°C and 3.0 kg/cm^3, respectively, and the pH of the HF tank was automatically controlled using an acid or a base. Pretreated biomass was continuously supplied to the bottom of the HF tank in order to load the high concentration of biomass in the SSF process. After completing hydrolysis and fermentation, the fermented mash was transported to a buffer tank using a pump.

This study used prehydrolysis at the start of the reaction and simultaneous saccharification and fermentation to produce bioethanol from *M. saccharifloru*s. Operating conditions were 35°C, pH 4.8 (controlled using sulfuric acid), and agitation at 60 Hz. Thirty FPU/g cellulose of enzyme cocktail (30 FPU/g cellulose Cellic CTec 2 and 15% Cellic HTec 2 based on the amount of added Cellic CTec 2) was continuously supplied to the HF tank. After prehydrolysis, a 7% seed culture was added to the HF tank. SSF samples were taken periodically to analyze the concentration of glucose, xylose, and ethanol. All experiments were conducted in duplicate, and then the average value was calculated and presented.

Purification process Purification processes such as distillation and dehydration are essential for producing fuel-grade bioethanol from lignocellulosic biomass. Distillation is a process for separating an ethanol from a liquid mixture on the basis of the selective vaporization and condensation of ethanol. It is generally used for the enrichment of bioethanol because the fermented broth contains impurities that must be removed. The distillation facility was divided into a mash column and a rectification column. Solid and liquid elements of the fermented broth were separated. The ethanol was enriched prior to distillation using a mash column, which consisted of stacked trays inside a column. The mash column was operated at a temperature of 100°C at the bottom and 82–84°C at the top. The first concentrated ethanol produced in the mash column was further enriched in the rectification column. Fuel-grade anhydrous ethanol could not be produced by distillation, because of the azeotropic point of water and ethanol. Therefore, the produced ethanol was transferred to a dehydration facility. This study adopted the pressure swing adsorption process in the dehydration step. The pressure swing adsorption process consisted of adsorption, depressurization, purge, and pressurization steps. Anhydrous ethanol was produced through the continuous application of these four steps using the human machine interface program (19).

Analytical methods Composition of pretreated and untreated materials was analyzed according to the National Renewable Energy Laboratory laboratory analytical procedure (20). After the reaction was completed, glucose and ethanol concentration was determined using high-performance liquid chromatography (HPLC, Waters, Milford, MA, USA) with an Aminex HPX-87P column (Bio-Rad, Hercules, CA, USA) and a refractive index detector. Deionized water was used as the mobile phase, and the flow rate was 0.6 mL/min at 85°C.

Economic analysis The net present value (NPV), internal rate of return (IRR), benefit-cost ratio (B/C ratio), and discount payback period (DPP) were calculated in the economic analysis (21). NPV is the difference between the present value of cash inflows and the present value of cash outflows. NPV is used in capital budgeting to analyze the profitability of an investment or project. The following is the equation for NPV:

$$\mathrm{PVB} = \sum_{t=0}^{T}\left(\frac{B_t}{(1+r)^t}\right),\ \mathrm{PVC} = \sum_{t=0}^{T}\left(\frac{C_t}{(1+r)^t}\right),\ \mathrm{NPV} = \sum_{t=0}^{T}\left[\frac{B_t - C_t}{(1+r)^t}\right] \quad (1)$$

In Eq. 1, PVB is the present value benefit, PVC is the present value cost, B_t is the benefit, C_t is the cost, r is the discount rate, and t is the number of time periods. The advantage of NPV as an indicator is that the evaluation method is simple. However, there are simplifying assumptions and arbitrary interpretations in the computation process, so this method does not reflect the complexity of markets, which can change and involve risk.

The B/C ratio (below) is calculated as the ratio of the present value benefit and the present value cost. The feasibility of a project is determined by the deducted value. In other words, if the B/C ratio is less than 1, the investment would be considered nonproductive.

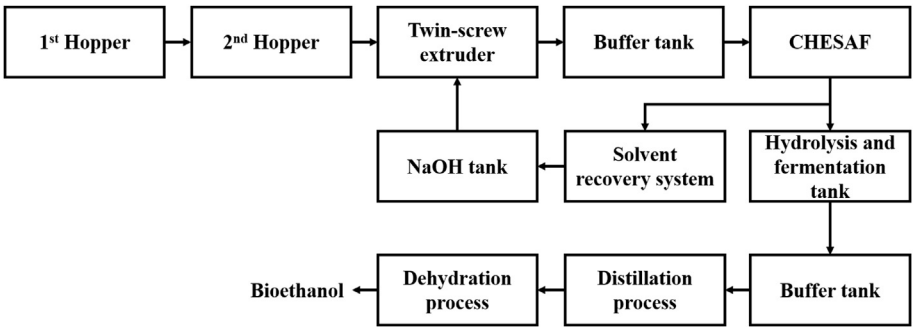

FIG. 1. Process flow diagram of the bench-scale plant for producing bioethanol from lignocellulosic biomass.

$$B/C \text{ ratio} = \frac{PVB}{PVC} \quad (2)$$

RESULTS AND DISCUSSION

Bioethanol production by the bench-scale plant In order to produce fuel-grade anhydrous bioethanol, an integrated process system from biomass pretreatment to purification was performed, as developed in previous studies (17–19). Bench-scale bioethanol production was divided into three major processes: the pretreatment; hydrolysis and fermentation; and purification processes (distillation and dehydration).

As mentioned above, the pretreatment process used a twin-screw extruder that is generally used in the polymer and food industry. A twin-screw extruder can be utilized in a continuous process, and would be practicable for large-scale production processes. It can easily control temperature and achieve high-efficiency pulverization because it has a high shearing force and a high throughput; further, it can be adapted to many different processes (20,22). The high shearing force of the CHEMET pretreatment with sodium hydroxide homogenizes the biomass and reduces the particle size so the pretreatment chemical can react more readily with the biomass surface (17). After pretreatment, the *M. sacchariflorus* contained 57.3% cellulose, 19.6% hemicellulose, 12.4% lignin, and 2.0% ash. Cellulose and hemicellulose were very promoting substrates for the ethanol production. The pretreated biomass was removed chemically in the first dewatering stage of the CHESAF process and washed in the washing stage. Washed samples were then dehydrated and supplied to the hydrolysis and fermentation tank. The moisture content of pretreated biomass supplied to the hydrolysis and fermentation tank was 50–60%. CHESAF used only one system with a single screw and a driving motor, thus it could simplify the process between the pretreatment and the hydrolysis steps.

For this study, prehydrolysis and SSF processes were used for producing bioethanol from *M. sacchariflorus*. Prehydrolysis was required to treat the biomass initially supplied because the glucose concentration was too low for the activation of the seed culture. After securing the required glucose concentration, 7% *S. cerevisiae* CHY1011 (7.0 ± 0.2 g/L cell density) was added to the reactor and then the SSF process was conducted at 33°C. As shown in Fig. 2A, the SSF process produced 46.5 g/L bioethanol, which included ethanol produced from enzyme and seed culture from a 20% biomass concentration. 41.9 g/L bioethanol was produced from only biomass. The theoretical amount of produced bioethanol was 58.2 g/L and the saccharification and ethanol conversion rates were 84.8% and 72.0%, respectively. In order to enhance the efficiency of bioethanol conversion, the impeller was improved as detailed in a previous study (18). As a result, 45.5 g/L of bioethanol was produced from the biomass-only mixture and the ethanol conversion rate was increased to 78.2% (Fig. 2B).

After fermentation, a purification process was performed to concentrate the ethanol to 99.5 wt%. The mash column was operated for purifying the fermented broth. When initially operated, the temperature at the top of the column was 94–95°C and the ethanol concentration was 6–8 wt%. When the ethanol was collected in the mash column reflux tank, ethanol was transferred to the top portion of the column. The temperature of the top portion was gradually decreased and the ethanol concentration increased with the increasing reflux ratio. The mash column could concentrate ethanol to 73.7 wt% (approximately 80 vol%) when column conditions were as follows: 1/8 reflux ratio, 82–83°C top portion temperature, and 100.6°C bottom portion temperature (Table S2).

After it was supplied to the rectification column, the ethanol could be concentrated to 93.3 wt%. The column was operated under a 1/10 reflux ratio, 77–78°C top portion temperature, and 100.5°C bottom portion temperature (Table S2).

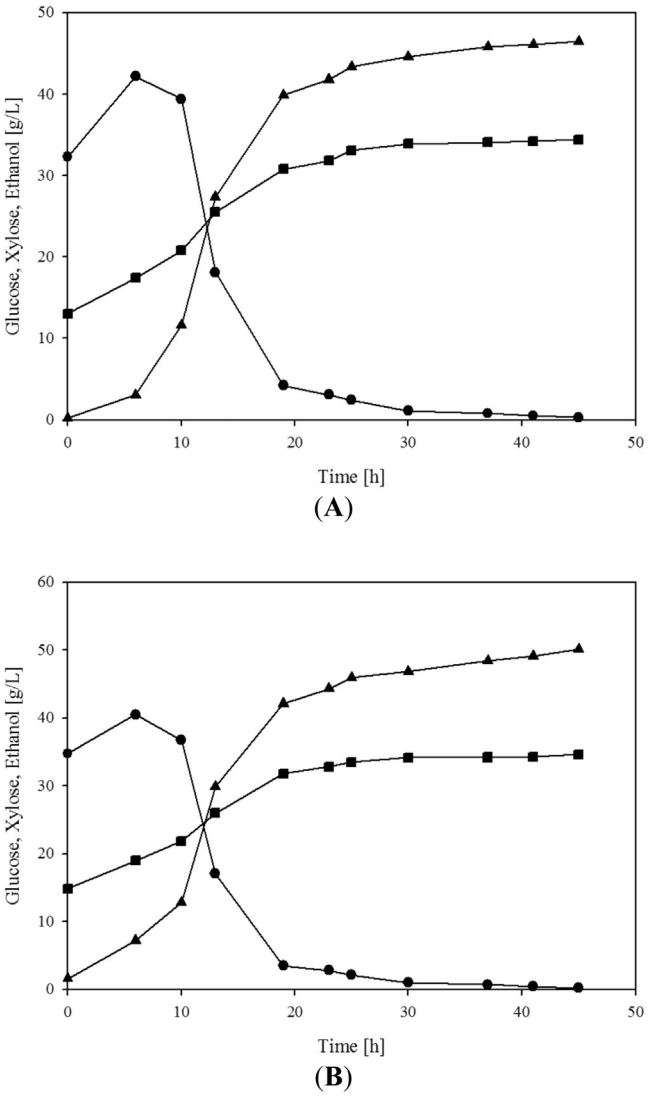

FIG. 2. Results of ethanol production from *M. sacchariflorus*. (A) Bench-scale plant with the normal-type impeller, and (B) bench-scale plant with the angled-type impeller. Circles, glucose concentration; squares, xylose concentration; triangles, ethanol concentration.

Finally, in order to produce fuel-grade anhydrous ethanol, dehydration equipment was used. The ethanol concentration varied from 99.2 wt% to 99.8 wt% depending on the purge ratio. For high purge ratios, the concentration of anhydrous ethanol increased but productivity of anhydrous ethanol decreased. The dehydration process was operated at a feed/purge ratio of 7/3 for the stable production of anhydrous ethanol. As a result, 99.6 wt% ethanol could be produced (Table S3).

A bench-scale bioethanol plant was designed, constructed, and operated for this study, and bioethanol was produced using *M. sacchariflorus* as lignocellulosic biomass. Ethanol (99.6 wt%) of 165 L was produced from one ton of *M. sacchariflorus* by the integrated plant.

Solvent recovery system Waste solvent from the pretreatment is harmful to the environment and solvent recovery can reduce the ethanol production cost. After pretreatment, the sodium hydroxide concentration was 0.21 M. Waste solvent of 700 L was continuously recovered by the solvent recovery system at

1.0–1.2 kg/cm^3 for 4 h and the flux was kept at 2.7–3.9 L/m^2/h (LMH). This means the system can recover one ton of waste sodium hydroxide in one day. After operating the solvent recovery system, 675 L of 0.207 M-solvent filtrate and 75 L of 0.241 M-solvent concentrate were recovered. The recovery rates were 98% based on the sodium hydroxide concentration of solvent and 90% based on the solvent volume.

In order to investigate the efficiency of the recycled solvent, pretreatment for the SSF process was performed under the same conditions with recycled solvent. The concentration of the solvent was adjusted to 0.4 M by adding 25% sodium hydroxide. The SSF process produced 45.8 g/L bioethanol, which included ethanol from the enzyme and seed culture from a 20% biomass concentration. The conversion rate was 70.8%, which signifies that the recovered solvent is no less effective than the original chemical. The ability to recover and reuse the pretreatment solvent was confirmed, and the solvent recovery system was effective at lowering the cost of bioethanol production.

Economic analysis The results of the economic analysis are summarized in Tables 1 and 2 (see also Table S4). The individual components (capital cost, operating cost, by-product credit, and ethanol production cost) are discussed in the following section.

Table 1 shows a summary of the process evaluation and the factors used for calculating the fixed capital investment. The capital investment of a bioethanol plant depends on the type of feedstock and the location and scale of the plant. In this study, the construction and operation period were assumed to be one year and twenty years, respectively. There supposed that 100,000 kL of bioethanol for transportation-fuel was produced by commercial plants from 606,061 tons of *M. sacchariflorus* in economic analysis. NREL estimated that the total capital investment of bioethanol plants, which produced 2,300,000 kL of bioethanol from corn stover, was US$422.5 million in 2007 (23). The largest percentage of the total capital investment was spent on the boiler and turbogenerator. The wastewater treatment and land accounted for 20% and 12% of the cost, respectively. The capital investment required for a cellulosic bioethanol plant utilizing *M. sacchariflorus* producing 100,000 kL of bioethanol per year is US$163 million.

The largest portion of operating costs is spent on feedstock, which also has an effect on the price of bioethanol. However, forecasting the price of cellulosic biomass is difficult in a volatile market environment, because the market for cellulosic biomass is unstable and has not settled yet. The price of *M. sacchariflorus* was estimated using the cost of harvest/baling and transportation, which were US$35.6/ton and US$58/ton, respectively. The costs of utilities (water, electricity, and steam), and chemicals for producing the bioethanol were estimated by examining the results of the bench-scale plant. These values and the calculations used are shown in Table 3 (detailed in Table S4).

The traditional production process of ethanol from corn or grain generates two by-products: carbon dioxide and distillers dried grains with solubles (DDGS). In this plant, xylose molasses and lignin were produced as by-products. Xylose molasses can be converted to xylitol, and lignin could be burned in a black liquor boiler and then a used as a heat reservoir. These kinds of by-product could decrease the bioethanol cost because they have additional marketable uses.

In this case, the production of cellulosic bioethanol was influenced by the ethanol cost, biomass price, enzyme price, and capital interest. Therefore, a sensitivity study was performed to examine how these kinds of factors would affect the ethanol production cost.

TABLE 1. Capital cost for producing bioethanol from cellulosic biomass.

Item	Cost (US dollar)
Land	20000000
Structure and housing	10000000
Pretreatment	12100000
Saccharification and fermentation	4070000
Chemical recovery	12100000
Distillation and solid recovery	14300000
Wastewater treatment	33000000
Storage	3300000
Boiler and turbogenerator	49500000
Utilities	4950000

TABLE 2. Estimating the production cost of cellulosic bioethanol.

Item	Cost (US dollar per kilo liter)
Raw material	561.2
Enzymes	328.0
Chemical	64.0
Wastewater treatment	72.0
Cooling water	7.3
Electricity	15.2
Steam	200.9
Labor	284.1
Depreciation	97.0
Sub-total	134.0
By-product credit	−458.1
Total production cost	1305.6

Sensitivity analysis Sensitivity studies were performed at various ethanol costs, biomass costs, enzyme costs and capital interests.

Fig. 3 shows the effect of ethanol cost on the NPV, B/C ratio, and IRR. When NPV is greater than zero, the ethanol price should be US$1.25/L. In order to be equal to the total net present value of US$163.9 million (the investment required to finance a plant producing 100,000 tons per year), the ethanol price should be US$1.38/L. Ethanol price should be US$1.25/L and then the B/C ratio to 1.0 (100%) and IRR to 0%. However, the loan rate is 5.22%, ethanol price should be US$1.33/L, higher than the loan rate.

The variability of enzyme costs has a considerable effect on the ethanol production cost, because the enzyme cost is 25% of the ethanol production total cost. If the enzyme cost was reduced by 50%, the NPV, B/C ratio, and IRR would increase to US$364.5 million, 1.21, and 8.53%, respectively. On the other hand, if the enzyme cost increased by 50%, the NPV, B/C ratio, and IRR would be reduced to −US$36.7 million, 0.94, and −3.16%, respectively (Table 3).

Biomass cost is of great importance as 43% and has a big impact on the bioethanol production cost. Table 3 shows the effect of biomass cost and loan rate on the economic feasibility of the project. If the biomass cost was decreased by 50%, the NPV would be raised to US$507.2 million. The result of increasing IRR to 12.79% was exceeding the lending interest rate. On the other hand, if the enzyme cost increased by 50%, the NPV and B/C ratio would be reduced to −US$179.4 million and 0.87 (Table 3). As shown in

TABLE 3. The effect of enzyme cost, biomass cost, and capital interest on economic feasibility.

		Cost changes (%)	
		−50	50
Enzyme	Total PVP (million US dollar)	364.5	−36.7
	B/C ratio	1.21	0.94
	IRR (%)	8.53	−3.16
Biomass	Total PVP (million US dollar)	507.2	−179.4
	B/C ratio	1.34	0.87
	IRR (%)	12.79	None
Capital interest	Total PVP (million US dollar)	325.1	55.6
	B/C ratio	1.11	1.00
	IRR (%)	5.32	0.22

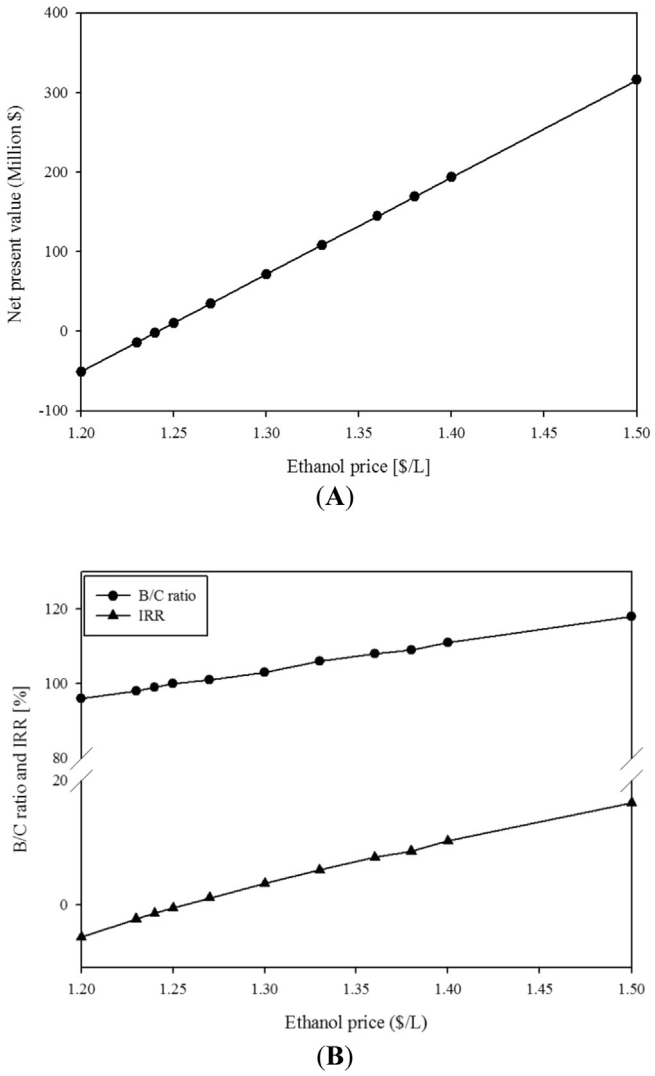

FIG. 3. The net present value (A), B/C ratio, and IRR (B) according to the ethanol price fluctuations.

Table 3, the variability of capital interests no significant fiscal effect on the overall business value than biomass and enzyme costs.

In summary, an economic analysis and sensitivity analysis were conducted to test the results gained from the operation of the bench-scale plant. As a result, the bioethanol production cost was calculated to be US$1.76/L, though reuse of the by-products could reduce the costs to US$1.31/L. Lastly, the enzyme cost and biomass cost had a considerable effect on the overall lignocellulosic bioethanol production cost.

Supplementary data to this article can be found online at https://doi.org/10.1016/j.jbiosc.2019.04.004.

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