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Research Article

Techno-Economic Analysis for Bioethanol Plant with Multi Lignocellulosic Feedstocks

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ABSTRACT. Oil palm empty fruit bunch and trunk are classified as primary lignocellulosic residues from the palm oil industry. They are considered to be promising feedstocks for bioconversion into value-added products such as bioethanol. However, using these lignocellulosic materials to produce bioethanol remains a significant challenge for small and medium enterprises. Hence, techno-economic and sensitivity analyses of bioethanol plant simultaneously treating these materials were performed in this study. The information based on preliminary experimental data in batch operations was employed to develop a simulation of an industrial-scale semi-continuous production process. Calculations of mass balance, equipment sizes, and production cost estimation of the production plant of various capacities ranging from 10,000 L/day to 35,000 L/day were summarized. The result based on 20 years of operation indicated that the net present value of the plant of lower capacities was negative. However, this value became positive when the plant operated with a higher capacity, 35,000 L/day. The highest ethanol yield, 294.84 $L_{EtOH}/ton_{feedstock}$, was produced when the plant treated only an empty fruit bunch generating 8.94% internal rate of return and US$ 0.54 production cost per unit. Moreover, the higher oil palm trunk ratio in the feedstock, the lower ethanol yield contributing to the higher production cost per unit.

Keywords: Bioethanol, Lignocellulosic, Techno-economic, Empty fruit bunch, Oil palm trunk

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1. Introduction

Fossil fuel depletion has become a severe issue. The reduction of fossil fuel sources is the main factor impacting on less energy fuel supply. Moreover, fossil fuel directly causes environmental impacts such as global warming and the greenhouse effect (Kang *et al.*, 2014).

Bioethanol is a form of renewable energy that can be produced from the fermentation of various feedstocks that contain sugars or carbohydrates, for example, rice, wheat, barley, potato, corn, and sugarcane. It is considered as alternative fuel energy to replace fossil fuel. This first-generation bioethanol has gained attention, but its production competes with the food supply and land utilization. The subsequent generation has been made for producing bioethanol from nonedible feedstocks, including lignocellulosic biomass such as cellulose and hemicellulose. However, it is not yet commercialized due to existing economic, technical and commercial barriers.

Lignocellulosic biomass is one of the most important renewable energy resources, not only avoid the competition of food sources but also solve the biomass disposal problems from agricultural industries especially, biomass produced from the palm oil industries. Oil palm empty fruit bunch (EFB) discarded from palm oil mills is a low-cost substrate containing about 38-70% cellulose (a linear unbranched polymer of hexose sugars), 10-35% hemicellulose (a group of polysaccharides) and 13-37% lignin (a very complex molecule of phenyl propane units) on a dry weight basis (Noorshamsiana *et al.*, 2017). Oil palm trunk (OPT) is a waste from the plantation of oil palm because oil palm trees are cut down for replanting at an interval of approximately 25 years. Its outer layers were often used for plywood manufacturing, but most of them tend to be discarded or burnt (Shahirah *et al.*, 2015).

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OPT contains around 22-44% cellulose, 12-41% hemicellulose, and 18-36% lignin on a dry weight basis (Noorshamsiana *et al*., 2017). Both EFB and OPT are considered as favorable lignocellulosic feedstocks due to their physicochemical characteristics. Moreover, they are highly abundant in Thailand, which is currently the third rank for palm oil producers in the world (Kang *et al*., 2014).

Bioethanol from lignocellulose is one of the most promising biofuels. It even plays an essential part in the 10 Year Alternative Energy and Development Plan (AEDP 2011-2021) from the Ministry of Energy Thailand to increase the alternative energy consumption by 25% in 2021 (Sutabutr, 2012).

Over the years, the technology to produce lignocellulosic ethanol has been under development. The unit steps in the process have been investigated through many researchers. For example, Baral and Shah (2017) studied the techno-economic of pretreatment technologies (steam explosion, dilute sulfuric acid, ammonia fiber explosion, and biological pretreatments), of which the models were developed by a commercial software. Their results suggested that the steam explosion pretreatment method is one of the good alternatives with low production cost.

Ballesteros *et al.* (2000) reported that the optimum particle size of softwood on a steam-explosion pretreatment was around 8 to 12 mm at 210 ºC and 4 min, from which a cellulose recovery was about 80%. In the same condition, 90% cellulose recovery was obtained from Ballesteros *et al.* (2002) and Negro *et al.* (2003).

Huang *et al.* (2008) reviewed various separation methods; Kunnakorn *et al.* (2013) compared azeotropic distillation with the hybrid system; Nagy *et al.* (2015) studied the stand-alone technology of distillation and pervaporation, and a hybrid process. All of them concluded that the hybrid technology using distillation and pervaporation not only saved energy demand but was also an environmentally friendly technology. The combination of distillation and pervaporation was suitable and efficient technology in terms of techno-economic analysis for ethanol production

Wingren *et al*. (2003) reported that SSF was a superior technique of the ethanol conversion process treating softwood as feedstock in terms of less capital investment and higher ethanol yield. They found that the total capital investment of SSF was 30% lower than that of SHF. Additionally, the production cost of SSF was 11% lower than that of SHF.

Despite moderately firm technologies, the feasibility and the economic viability of its production plant have been still unclear and even unknow for the plant treating OPT as feedstock. For example, Quintero *et al.* (2013) performed a techno-economic analysis of bioethanol production from four raw materials (sugarcane bagasse, coffee cut-stems, rice husk, and empty fruit bunches), intending to compare production cost for each raw material. Their results showed that EFB with high cellulose content and the low raw material cost was the highest potential feedstock to produce bioethanol with its production cost of 0.5779 US$/L. Nonetheless, the capacity of their plant model was fixed at 100,000 L/day, and only the chemical composition of the raw material was determined experimentally. Achinas *et al*. (2019) analyzed the feasibility of a bioethanol plant from lignocellulosic feedstocks using a simulation software. Their process consists of multiple cycles of size reduction process to produce 99.7 wt.% ethanol. Although they could reduce the purchase costs, the capital and operating costs were still high, resulting in a negative net present value.

As a result, this proposed work aimed to technically and economically investigate the feasibility of a bioethanol plant treating EFB, supposed to be the primary feedstock, and OPT, considered as seasonal raw material due to its fluctuating supply to provide techno-economic information to a broad audience and also proposes a flexible process working with multi feedstocks as a baseline study for further business investigations for an investment decision.

2. Materials and Methods

2.1 Material

The EFB residue was obtained from Chumporn Palm Oil Industry Public Co., Ltd. in Chumporn province, Thailand. The EFB was washed several times with running tap water and dehydrated overnight and then ground to a particle size of about 20×20×5 mm. Dried materials were kept inside plastic bags and kept in desiccators for 24 hours for removing moisture and other volatile impurities. The OPT residue was obtained from a farmer in Krabi province, Thailand. The preparation of OPT was the same as that of the EFB. The composition of these lignocellulosic feedstocks in the preliminary experiment is summarized in Table 1. The EFB contains higher cellulose and hemicellulose fractions; meanwhile, it has a lower lignin ratio.

2.2 Methods

2.2.1 Procedure

The preliminary experiment, including pretreatment (consisting of steam explosion (SE), hot water washing and hydrogen peroxide digestion (H_2O_2)), hydrolysis and fermentation of each feedstock, were performed separately in a laboratory scale.

The semi-continuous process was then designed in Aspen Plus based on the optimal conditions from the preliminary experiment. Process scheduling was performed to explore the size and number of units. Additionally, the distillation and pervaporation (PV) were also employed and optimized to produce fuel-grade ethanol (99.5 wt.%).

Then, the economic analysis was performed. Its economic indicators, including net present value (*NPV*), internal rate of return (*IRR*), payback period (*PB*), and production cost per unit, were analyzed to calculate the total capital and operating cost of the plant.

Table 1
Mass percent composition of lignocellulosic feedstocks

	Cellulose	Hemicellulose	Lignin	Ash	Other
EFB	38.85	26.14	11.62	1.4	21.99
OPT	38.67	23.3	23.76	1.62	12.65

Since the fluctuation in ethanol selling price and other expenditures, a sensitivity analysis was also applied. The critical variables for the sensitivity analysis in this proposed work were the change of ethanol selling price and influence variables contributed to total capital and operating cost.

2.2.2 Specific condition for mass balances

The process to produce bioethanol from lignocellulose composes of technological unit steps including *Pretreatment* to remove lignin, reduce the crystallinity of the cellulose and increase the porosity of lignocellulosic materials to make cellulose and hemicellulose more amenable to the following processes; *Hydrolysis* to turn cellulose and hemicellulose into sugars (complete hydrolysis of cellulose yields glucose); *Fermentation* to convert sugars into cellular energy, producing ethanol and carbon dioxide as by-products; *Distillation* to concentrate dilute aqueous solution of ethanol from the fermentation process; and *Dehydration* to separate azeotropic mixtures of ethanol-water.

Various conditions in each step of operations, excluding distillation and dehydration, were performed in the preliminary experiment. The results indicated that to produce the highest ethanol yield, the optimal conditions were as follows;

- The feedstocks were milled to small particle sizes of about 20×20×5 mm. Then, its moisture was removed by desiccators prior to pretreatment.
- In the first pretreatment step, the size-reduced feedstocks were added into an SE vessel. After that, the hot stream was injected. After the temperature and pressure inside the vessel reached 210 ºC and 18.6 bar, it was held for 4 min. After the explosion, the solid part was gathered by vacuum filtration.
- The hemicellulose residual was then removed by hot water washing (washing 80 °C, 1:8 by weight of treated feedstocks) in the second step.
- In the last step, the treated feedstocks were fed into a digestion vessel containing (by weight of treated feedstocks) 1:10 of water, 10,000:3.5 of NaOH, and 100:3 of H_2O_2 at 70 °C. All of the components was mixed for 30 min.
- After the end of the pretreatment, cellulose content of EFB and OPT increased from 38.85% and 38.67% to 70.57% and 67.86%, respectively.
- The treated feedstocks were neutralized with deionized water until its pH reached 7 and then mixed with growing media.
- The growing media consisted of 0.05 M buffer solution (10:1 of feedstock) at pH 4.8, 10 g of yeast extract, and 20 g of peptone per liter of the buffer solution.
- The mixed solution was sterilized at 121 °C for 20 min before they were fed into another unit.
- SSF and SHF were then compared. It was found that the SSF produced a higher yield. This result is agreeable with results from Wyman *et al.* (1992), Alfani *et al.* (2000), Öhgren *et al.* (2007), Dahnum *et al. (*2015) and Suttikul *et al.* (2016).
- The optimal conditions of the SSF were 40 ºC, 10% of inoculum (*Saccharomyces cerevisiae* TISTR 5606), 10 FPU of enzyme Ctec2 per gram of feedstock, and 60 h retention time.

From these conditions, the ethanol concentration in the broth after fermentation was 3.1 wt.%.

2.2.3 Conceptual design of an industrial plant

The operation model using batch-type units in the preliminary experiment, as illustrated in Fig. 1, was transformed to an industrial-scale semi-continuous production through a commercial software, Aspen Plus.

Aspen Plus program version 8.8 was used to model the bioethanol production process in the aspects of mass balances and design specifications. The configuration of the simulation model is illustrated in Fig. 2. In the simulation model, a scheduling approach was also employed in order to achieve high-efficiency production. Furthermore, in this kind of operation, size and numbers of equipment are determined to minimize waiting time, avoid the bottleneck in the process, and accomplish the desired capacity. The production plant was designed to be able to produce ethanol from two feedstocks (EFB and OPT) with various weight ratios (100:0, 80:20, 50:50, 20:80 and 0:100).

A mass balance calculation and scheduling were performed prior to equipment size evaluation. Since the upstream processes took a shorter time compared with an individual SSF unit, they could be designed to be a cyclic operation of small units to avoid the bottleneck and minimize the equipment purchasing costs. Additionally, Different weight ratios of feedstocks provided different mass transfers. In order to cover the maximum capacity of the plant and produce the desired bioethanol every day, pure EFB feedstock was assigned as a reference baseline to calculate the equipment size.

2.2.4 Computational method and model simulation of an industrial plant

The simulation of the bioethanol production process with multi lignocellulosic feedstocks was carried out by the non-random two-liquid (NRTL) model.

In the model, feedstocks were fed into a size reduction unit from which the size of feedstocks was reduced. Most of the moisture in the feedstocks was then removed by sun-drying. Afterward, the size-reduced feedstocks were conveyed to the pretreatment process. In this process, a specific initial pretreatment step was applied to obtain a high breakdown efficiency of their structure. The EFB and OPT were, afterward, treated by an SE unit.

Both feedstocks were fed into a high-pressure vessel where hot steam was fed in until the desired conditions (210 °C, 18.6 bar) was reached, and then the vessel was quickly depressurized. Both treated feedstocks were mixed and introduced to a hot water washing step to get rid of the hemicellulose residual. The last step was the digestion using 3 wt.% H_2O_2 to destroy the structure and remove the lignin.

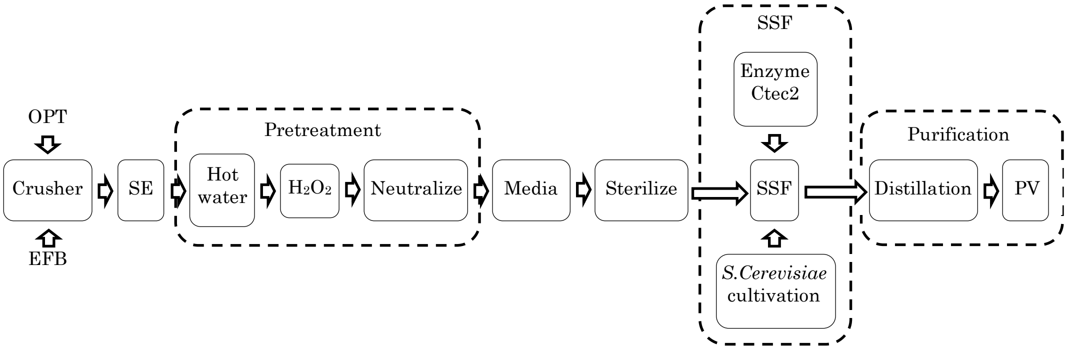

Fig. 1 Bioethanol production process in a batch-type operation

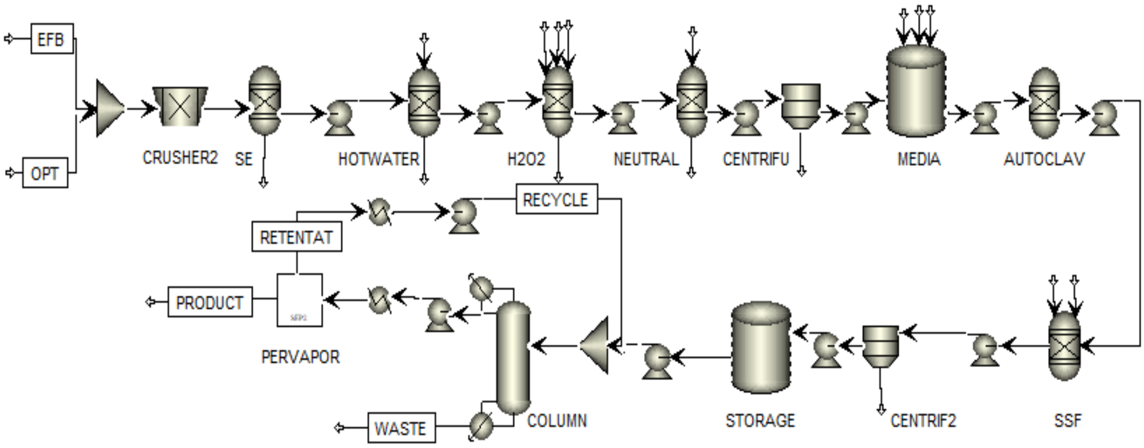

Fig. 2 Aspen Plus simulation model for a bioethanol process

This step in the pretreatment section could remove structural and compositional impediments to hydrolysis to improve the rate of enzyme hydrolysis and increase yields of fermentable feedstocks.

The pretreated feedstocks with less amount of lignin were then neutralized and transferred to a sterilizer unit before they were sent to the SSF unit where cellulose was converted to ethanol. In the SSF unit, enzyme Ctec2 and *Saccharomyces cerevisiae* TISTR 5606 were added simultaneously into the fermenter, where hydrolysis and fermentation occurred at the same time. This process took 60 h to produce 3.1 wt.% ethanol solution, called fermentation broth.

In order to produce fuel-grade bioethanol, it was required to increase ethanol concentration to 99.5 wt.% by the purification section. However, the purification section generated high production cost (Ebrahimiaqda and Ogden 2017). So, it was essential to select a suitable and available technology for purifying dilute ethanol with less energy consumption.

A combination of distillation and pervaporation to produce fuel-grade ethanol from diluted solution consumed energy demand lower than a 3-stage pervaporation of about 28.83% (Nagy *et al.*, 2015). Therefore, the combination of this technology was employed in this section.

The distillation column was performed using the RadFrac module with the equilibrium method. The operating conditions in a distillation column (number of stages, distillation to feed, molar reflux ratio, feed stage) were analyzed to determine the suitable parameters to produce ethanol at desired concentration before feeding to a pervaporation unit.

The energy consumption (Q, MJ/kg) in the pervaporation unit was carried out to calculate the energy needed to evaporate the permeate in the pervaporation model. The calculation method in this unit was conducted as recommended by previous work (Nagy *et al.*, 2015).

The fermentation broth was distilled near azeotrope point by the Radfrec unit and pumped to the pervaporation model to remove impurities. The permeated steam contained a high level of water content, but it still had a small amount of ethanol. So, it was condensed and recycled back to the distillation column to minimize the loss of ethanol. The polyimide 6FDA-NDA/DABA (2.7 $kg/m^2/h$; Le and Chung, 2014) was selected to dehydrate the top product of the distiller to fuel-grade ethanol as a final product.

2.2.5 Economic analysis

An economic analysis is a process that makes a clear picture of the existing economic feasibility. There are several tools for economic evaluation used to present a comprehensive view of the investment costs and benefits of the project. The fundamental tools for a measure of project worth including NPV, IRR, PB and production cost per unit (Achinas *et al.*, 2019), as illustrated in eq. (1) and (2).

The economic analysis evaluates the costs and benefits of a project. The *NPV* of a project reflects the worth of the project. *IRR* is used to reveal the attractiveness of the project and predict the possibility of generating profit. *IRR* is a discount rate that makes *NPV* of all cash flows equal to zero; it should be higher than the rate of return of a company's desire.

$$NPV = \sum_{i=0}^{T} \frac{C_i}{(1+r)^i} \quad (1)$$

where:

- r is the discount rate,
- T is the number of time periods,
- C_0 is an initial investment and
- C_i is cash flow at year i.

PB is the length of time required to recover the cost of an investment, which can be calculated from eq. (2).

$$\text{PB} = \frac{\text{Cost of investment}}{\text{Annual net cash flow}} \quad (2)$$

Related parameters were considered to conduct the evaluate the economic feasibility of plant, consisting of:

- Operating period = 20 years; equal to 7,200 hours per year;
- The discount rate = 5.775% (minimum loan rate) (Krungthai Bank 2020);
- Tax rate = 20% (Thailand Corporate Tax Rate);
- Depreciation expense = 20% with straight-line depreciation method (Thailand Tax Depreciation Rates);
- Savage value and profit from land after the end of the plant's life = 20% (TerraBKK Research) and
- The escalation of products = 1% (Diopenes and Laptaned 2011).

The costs of conventional equipment such as mixing tanks, heated jacketed vessels, and pumps were estimated based on vendors' citations. Those of coolers, heaters and a distillation column were calculated by the Aspen process economic analyzer and summarized in supplementary Table S1.

Direct fixed capital (*DFC*) is represented as the total capital cost. It consists of total plant direct cost (*TPDC*), total plant indirect cost (*TPIC*), contractor's fee, and contingency. *TPDC* is estimated based on the total equipment purchasing cost (*PC*) in the plant. *TPIC* is based on *TPDC*. The contractor's fee and contingency are dependent on the summation of *TPDC* and *TPIC*. Table 2 illustrates the multiplying factor of all elements according to previous work (Petrides, 2000). Additionally, land price was estimated to be 4% of total equipment purchasing cost as Peters and Timmerhaus (1980) reported.

The operating costs consisted of raw material costs, labor costs, chemical costs, plant overhead costs, general and administrative (*G&A*) expenses, utility expenses, maintenance expenses and miscellaneous expense.

According to Goldthorpe *et al.* (2014), plant overhead costs accounted for 25% of a summation of labor costs and maintenance expenses. *G&A* expenses accounted for 4% of a summation of labor costs, plant overhead costs and maintenance expenses.

All of the maintenance expenses was estimated to be 10% of equipment purchasing costs (Aspen Plus). The miscellaneous expense was an expense, excluding the above elements. It was assumed to be 1,000 US$/year (Diopenes and Laptaned 2011). For the pervaporation model, the membrane needed to be replaced every 5 years (O'Brien *et al.*, 2000). The price of the membrane, on average, was 200 US$/m^2 with a US$ 100 replacement cost. The price of EFB was based on the average value of retail quotes, accounting for less than 1% of the total operating cost. The cost of OPT could not be clearly estimated, so it was assumed to be equal to that of oil palm fiber.

Table 2
Estimation of total capital cost

Total plant direct cost	Base cost	Multiplying
TPDC		
Equipment purchasing cost	PC	-
Installation	PC	0.5
Process piping	PC	0.4
Instrumentation	PC	0.35
Insulation	PC	0.03
Electrical	PC	0.15
Buildings	PC	0.45
Land	PC	0.04
TPIC		
Engineering	TPDC	0.25
Construction	TPDC	0.35
Contractor's fee	TPDC + TPIC	0.05
Contingency	TPDC + TPIC	0.1

3. Results and Discussion

3.1 Scheduling

The schedule of each unit operation in the upstream process before the SSF section is illustrated in Fig. 3.

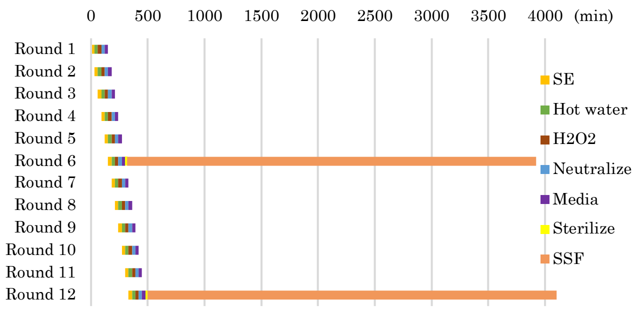

Fig. 3 Equipment utilization to fill up two consecutive SSF vessels

In order to fill up 1 SSF vessel, 6 operation rounds of 3 pretreatment units, 1 pH-adjusting unit, and 1 media preparing unit were required. Since the upstream processes were designed to be a cycling operation, a new round was initiated every 0.5 h. The time needed to operate from the preparation of feedstocks to the final product from the first SSF vessel was 65.33 h, and a next vessel finished in 3 h later. A set of 8 SSF vessels was required to produce 10,000 L of ethanol. The first 10,000 L of ethanol finished in 86.33 h and a next batch completed in 24 h. 21 SSF vessels were provided in the plant to avoid bottlenecks.

3.2 Size and number of units

The equipment utilization from scheduling exhibited the number of operation rounds and set of units. These data and the desired mass throughput were applied to estimate required equipment sizes in the pretreatment and SSF sections, as summarized in supplementary Table S2. These equipment sizes were based on pure EFB feedstock because this feedstock generated the highest mass throughput. The units in the purification section, namely distillation and pervaporation units, were optimized simultaneously based on the desired concentration and flow rate of the final product. The optimized sizes and conditions of these units are summarized in supplementary Table S3.

It is worth noticing that the fermentation broth released from the SSF vessels in each plant capacity had different mass throughput; therefore, the purification process had to operate under various conditions to produce the same quality of the final product.

3.3 Effect of feedstock ratio

The feedstock ratio directly affected the bioethanol production capacity of the plant. EFB had a high potential to produce bioethanol due to higher ethanol yield compared with that from the combinations of EFB and OPT, which had more lignin fraction. An example of this effect is summarized in Table 3.

3.4 Economic results

The elements of the total capital cost are summarized in supplementary Table S4. Moreover, the components used to calculate the operating costs are summarized in supplementary Table S5. The most significant part of the operating costs is the utility costs consisting of electric city, steam, and cooling water.

As a result, the *NPV* of the 10,000 L/day plant is negative for all weight ratios of EFB:OPT, as summarized in Table 4. This result is compatible with that of Table 3 in view of the lower ethanol yielded from that plant (or the higher OPT ratio in the feedstock), the lower *NPV* of the plant.

Table 3
Ethanol yield from the 10,000 L/day plant treating multi feedstocks at different weight ratios

EFB:OPT	100:0	80:20	50:50	20:80	0:100
Ethanol (L/ton)	294.97	284.16	268.26	252.55	242.17

Table 4
Economic results of the 10,000 L/day plant treating multi feedstocks at various weight ratios

	EFB:OPT				
	100:0	80:20	50:50	20:80	0:100
NPV ($\times 10^6$ US$)	-0.595	-1.571	-2.920	-4.382	-5.364
IRR (%)	5.40	4.79	3.90	2.93	2.26
PB (years)	13.89	14.8	16.3	18.27	n/a
Cost/unit (US$)	0.666	0.686	0.716	0.752	0.778

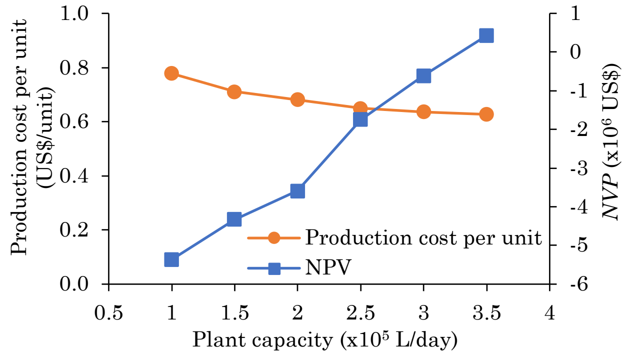

Fig. 4 NPV and production cost per unit for the plant treating pure OPT feedstocks of various capacities

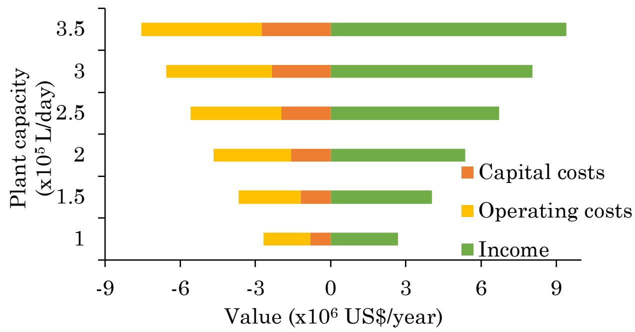

Fig. 5 Investment costs and revenue per year of the plant treating pure OPT feedstock of various capacities

The capacity of the plant was adjusted to take advantage of economies of scale to overcome this issue. Moreover, from the scenario of higher OPT ratio in feedstock yielding lower *NPV*, The *NPV* of the plant treating pure OPT feedstock can be a bottom line of all *NPVs*. The results of the plant of various capacities are illustrated in Fig. 4. Increasing plant capacity creates higher equipment costs (the most significant part of the investment costs) and operating costs (utilities, chemicals, etc.). Even though the plant capacity negatively affects total capital and operating cost significantly. The *NPV* of the plant treating pure OPT feedstock increases when the plant capacity increases and becomes a positive value when the plant capacity is 35,000 L/day. Meanwhile, the production cost per unit continually decreases with the rise of the plant capacity. These phenomena occur as a consequence of that the increase of total investment costs is lower than that of annual income, as illustrated in Fig 5. Increasing plant capacity provides a larger quantity of the final product. Therefore, the increased total investment costs can distribute over a more massive output. Furthermore, the NPVs of the 35,000 L/day plant treating multi feedstocks at other weight ratios are positive (as previously described). An economic result of the plant is summarized in Table 5.

The production cost of lignocellulosic derived ethanol with a variety of feedstocks under different conditions in pretreatment and fermentation are illustrated in Table 6. Most of the works in this table reported that not only ethanol yield but also feedstock, enzyme and utility costs were the key influencing factors of the ethanol production cost. On the contrary, the feedstock costs in this proposed work played an unimportant role, even almost negligible.

Table 5
Economic results of the 35,000 L/day plant treating multi feedstocks at various weight ratios

	EFB:OPT				
	100:0	80:20	50:50	20:80	0:100
NPV (x10^6 US$)	17.5	14	8.97	3.93	0.426
IRR (%)	8.94	8.32	7.42	6.51	5.85
PB (years)	10.07	10.59	11.45	12.45	13.26
Cost/unit (US$)	0.540	0.555	0.579	0.606	0.627

Table 6
The summarization for the production cost of ethanol from biomass

Raw material	Pretreatment	Fermentation	Production cost (US$/L)	Ref.
EFB:OPT				
100:0	SE 210 °C, 18.6 bar, 4 min. Hot water washing 80 °C, 30 min, H_2O_2 digestion 3 wt.% H_2O_2, 70 °C, 30 min	SSF Enzyme Ctec2 10 FPU/g feedstock, *S.cerevisiae* at 40 °C, 60 h	0.540	Proposed work
80:20			0.555	
50:50			0.579	
20:80			0.606	
0:100			0.627	
Spruce	SO_2 steam pretreatment SO_2 loaded 2.4 kg/100 kg dry wood, at 215 °C, 3 min	SHF Enzyme activity of 19 FPU/g cellulose, 38 °C, 96 h	0.57	Wingren *et al.*, 2003
		SSF Enzyme activity of 32 FPU/g cellulose, 38 °C, 48 h	0.63	
Salix	SO_2 steam pretreatment 195 °C, SO_2 conc. 1 %	SSF Yeast conc. 3 g/L, Enzyme 15 FPU/g water-insoluble solid, 37 °C, 72 h	0.871	Sassner *et al.*, 2008
Corn stover	190 °C, SO_2 conc. 1 %	Yeast conc. 1.8 g/L, Enzyme 15 FPU/g water-insoluble solid, 35 °C, 72 h	0.865	
Spruce	205 °C, SO_2 conc. 1.25 %	Yeast conc. 2.5 g/L, Enzyme 15 FPU/g water-insoluble solid, 37 °C, 72 h	0.693	
Corn stover	Dilute sulfuric acid	SHF Cellulase enzymes, at 65 °C, 36 h Fermentation at 41 °C, 72 h by *Zymomonas mobilis*	0.641	Aden and Foust 2009
Sugarcane bagasse	Dilute acid 135 °C, 4 h	SHF Cellulase enzyme, at 50 °C, 96 h Fermentation at 33 °C, 30 h by *Zymomonas mobilis*	0.7662	Quintero *et al.*, 2013
EFB			0.5779	
Rice Husk			0.6393	
Coffee cut-stems			0.6807	
Rice straw	Hydrothermal 180 °C, 3 Mpa	SHF Cellular enzyme 10 FPU/g-dry straw, at 45 °C, 72 h Fermentation at 30 °C, 24 h	1.19	Diep *et al.*, 2015
Dried cassava	Liquefaction with enzyme	SSF Unreported conditions	0.4695	Quintero *et al.*, 2015
Fresh cassava			0.4203	
Miscanthus sacchariflorus	95 °C, 0.4 M NaOH	SSF 30 FPU/g cellulose of enzyme cocktail (Cellic CTec 2 and Cellic Htec2), *S. cerevisiae* at 35 °C, pH 4.8,	1.76	Kang *et al.,* 2019

The economic results of the proposed study were based on a small industrial scale plant. For future studies, it would be highly recommended to develop a medium to large industrial scale, since the economies of scale were proved to be critical.

3.5 Sensitivity analyses

Sensitivity analysis of the 35,000 L/day plant was carried out for all EFB:OPT weight ratios to determine how the uncertainty in the economic feasibility of the plant can be divided and allocated to different sources of uncertainty in its inputs. The *NPV* of the plant was determined to be a function of equipment costs as a major variable of the total capital cost; chemical and utility costs as the two most substantial elements in the operating costs; and the ethanol selling price in Thailand (0.781 US$/L as a base price). It was assumed that all sources of uncertainty would increases or decreases to the extent of 20%.

The slope of the criterion lines in Fig. 6 indicates how sensitive the plant economics are to changes in these criterion – the steeper the line, the more sensitive the plant economics are to that sources of uncertainty.

It can be noticed from the sensitivity analysis that the plant is exceptionally robust and is mostly insensitive to the operating costs (chemical and utility costs). As expected, the plant economics are most sensitive to the ethanol selling price.

Additionally, the change of EFB:OPT weight ratio does not affect the slope of the criterion lines in case of the equipment costs (Fig. 6(a)), the chemical costs (Fig. 6(b)) and the utility costs (Fig. 6(c)). However, there is an interaction between the EFB:OPT weight ratio and the ethanol selling price (Fig. 6(d)). Changes of the EFB:OPT weight ratio deviates slope of the criterion lines, meaning that the effect of the ethanol selling price on *NPV* depends on the level of the EFB:OPT weight ratio in several different ways.

4. Conclusion

The techno-economic evaluation confirms that the plant treating multi feedstocks using proposed technology to produce 35,000 L/day of bioethanol is feasible. However, EFB is a better promising feedstock to provide the high ethanol yield, which is the primary parameter impacting the production cost per unit. The economic results indicate that pure EFB feedstock can provide the highest ethanol yield with the lowest production cost (0.54 US$/unit). The *NPV* of this scenario is 17.5x10^6 US$ and 8.94 % *IRR* with a payback period of 10.07 years. On the other hand, a higher ratio of OPT produces a lower quantity of final product reflecting on high production cost per unit. However, OPT is a real waste after replantation; therefore, the cost of this raw material can be neglected, and higher profit can be achieved.

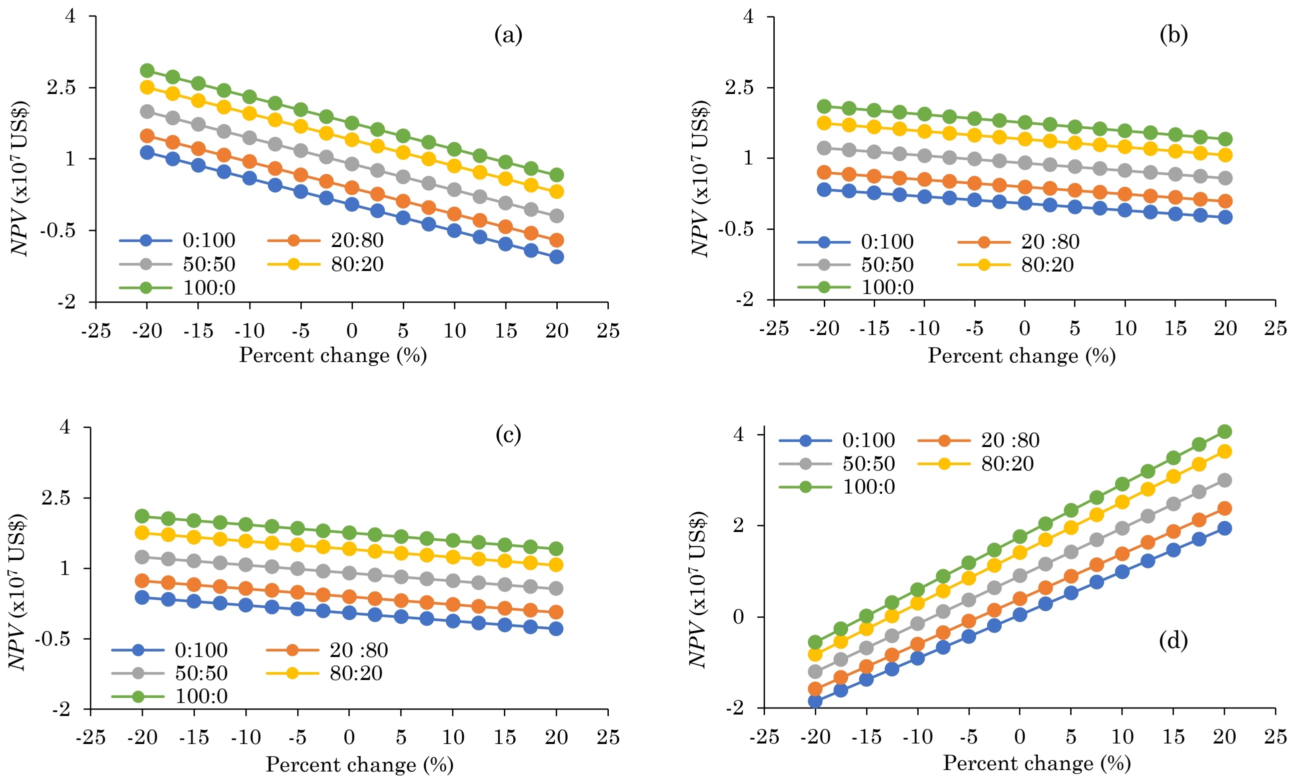

Fig. 6 Effect of (a) equipment costs, (b) chemical costs, (c) utility costs and (d) ethanol selling price on *NPV* of the 35,000 L/day plant treating multi feedstocks at various EFB:OPT weight ratios

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