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Research article

Techno-economic analysis of commercial-scale bioethanol production from oil palm trunk and empty fruit bunch

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Abstract

Importance of the work: Lignocellulosic ethanol production from oil palm trunk (OPT) and empty fruit bunch (EFB) is considered a promising approach because the materials are abundant and inexpensive.
Objectives: To produce ethanol 99.0 wt% and 10,000 L/d from OPT and EFB and to evaluate the economic feasibility using techno-economic analysis.
Materials & Methods: The Aspen Plus Software was used to simulate the bioethanol production model. The scheduling of the batch process was generated using Aspen Batch Process Developer. Process controllability of the purification section was studied using Aspen Plus Dynamics. For economic feasibility, a techno-economic analysis (TEA) was performed.
Results: The results indicated that bioethanol production with pervaporation was profitable when blended and sold as gasohol E20. In total, 48,000 kg/d of OPT and EFB were required to achieve the target capacity. Furthermore, increasing the OPT ratio provided a higher economic benefit. The best economic results in terms of the net present value, the internal rate of return and the payback period were at an OPT-to-EFB ratio of 75:25. The implemented control system in Aspen Plus Dynamics proved that the controllers could handle disturbances and control process variables to the desired specifications.
Main finding: Extractive distillation and pervaporation were successfully used for ethanol dehydration. The highest ethanol production rate was by feeding only EFB and using pervaporation as purification technology. TEA indicated that the total capital investment of bioethanol production using pervaporation was higher than for extractive distillation.

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Introduction

Energy demand has been increasing due to the global rise in population, while the availability of fossil fuels (crude oil, natural gas, or coal) is anticipated to decrease (Kaze, 2022). Bioethanol has become an alternative renewable fuel because of its potential for increasing the oxygen content in gasoline and proven lower carbon monoxide emission (Quintero et al., 2006). The first generation of bioethanol production used feedstock from food sources, such as corn, cassava and sugarcane; however, these raw materials competed with the need for food supply. Consequently, lignocellulosic biomass or bioenergy crops ((such as wood chips, palm residue, straw) are now used as an alternative raw material to produce the fuel through fermentation to produce what is known as "second-generation ethanol" (Robak and Balcerek, 2017). However, bioethanol production from lignocellulosic biomass requires a pretreatment step to facilitate accessing the cellulose, the source of sugar in yeast fermentation (Chiaramonti et al., 2012).

Several substances of lignocellulosic biomass have been studied for bioethanol production, including oil palm residues, such as the trunk, empty fruit bunch and frond, as these are abundant waste materials from the oil palm extraction process in Thailand. Various EFB components have been reported to have significance to production in Brazil (Raman and Gnansounou, 2014). Intensive studies in the laboratory proved that the oil palm trunk and oil palm empty fruit bunch are potential feedstock for bioethanol with a concentration of 3 wt% in fermentation. These wastes have received attention for utilization due to their component of 35–40 wt% cellulose. A pretreatment approach with a high efficiency has been examined to increase enzymatic and yeast accessibility (Pangsang et al., 2019; Tareen et al., 2020).

However, bioethanol production on a commercial scale could not be achieved without a separation process that must be conducted to make the concentration of bioethanol reach fuel grade (greater than 99.0 wt%). Unfortunately, fractional distillation cannot be utilized to obtain the desired grade easily. Because of the high azeotropic point between ethanol and water for a composition of 95.63 wt% ethanol and 4.37 wt% water, various separation techniques have been developed resolve this problem. Extractive distillation is a widespread technology in industry utilized to break the azeotrope between ethanol and water. It includes an entrainer or a solvent to change the low relative volatility of the key components without forming a new azeotrope and to improve the efficiency of this technique (Li and Bai, 2012). Another attractive separation technique is pervaporation, which uses a thermal membrane process to separate azeotropic mixtures. It is based on differences in diffusion rates through a solid membrane that combines permeation and vaporization (Jyoti et al., 2015). A simulation approach for residues was proposed and compared in terms of techno-economics (Quintero et al., 2013). The techno-economic study used in the work of Srinophakun et al. (2020) lacked full batch schedule and dynamics behavior in terms of controllability.

The current study aimed to evaluate the economic aspects, using techno-economic analysis, of two bioethanol production procedures, namely the integration of pretreatment and fermentation together with extractive distillation or pervaporation technologies. OPT and EFB were chosen as the feedstocks to produce fuel-grade bioethanol. Scheduling of pretreatment and fermentation (batch process) and process controllability of separation approaches (continuous process) were also investigated.

Materials and Methods

Software for simulation

The simulation model of bioethanol production was carried out using the Aspen software version 11 (Aspen Technology, Inc.; Bedford, Massachusetts, USA), especially Aspen Plus. Aspen Custom Modeler was used to create a customized model to perform the continuous process in Aspen Plus. The scheduling of the batch process was generated using Aspen Batch Process Developer with the continuous feed into product separation condition. Furthermore, the process controllability of the purification section was studied using Aspen Plus Dynamics.

Design concept of bioethanol production

Techno-economic analysis of commercial-scale bioethanol production was conducted based on 10,000 L-ethanol/d capacity and 99.0 wt% ethanol purity. OPT and EFB were considered as feedstocks to produce bioethanol; the feed properties are presented in Table 1. This process was separated into three sections: feed preparation-pretreatment, fermentation and purification. For the feed preparation pretreatment, OPT and EFB were pretreated using a different method. OPT was treated using steam explosion (SE) and hot water (HW), whilst EFB was treated using hot-compressed water (HCW).

Table 1 Feed properties of oil palm trunk (OPT) and empty fruit bunch (EFB) raw materials

Composition (wt%)	OPT	EFB
Cellulose	38.67	38.85
Hemicellulose	30.22	26.14
Lignin	11.60	11.62
Ash	1.62	1.40
Other	17.89	21.99

After that, they were mixed together and treated using alkaline hydrogen peroxide (AHP). For fermentation, the selected approach to produce ethanol was simultaneous saccharification and fermentation (SSF) because it provided a high ethanol yield and used a single reactor to hydrolyze and ferment simultaneously. Laosiriwut et al. (2020) developed correlations for pretreatment units that were used to create the models in the current study, as shown in Table 2. This study only considered the simulation and basic design concept. Then, the product stream from fermentation, called broth, was purified to obtain the fuel-grade ethanol. This study anticipated comparing two purification techniques: a conventional method (extractive distillation using ethylene glycol) and an alternative method (pervaporation). The more suitable technology was selected using the best economic parameters as indicators. Since the two feedstocks were dependent on seasonal plantation harvest, the feed ratio between OPT and EFB was also studied: 75:25, 50:50, 25:75 and 100:0. The highest designed fraction of OPT was 75% since the processing plant aimed to operate normally using EFB and to reserve using OPT for specific seasons. A modified step with stillage recycling has been investigated recently (Puengprasert et al., 2020) but was not included in the current study.

Table 2 Configurations of bioethanol production from oil palm trunk (OPT) and empty fruit bunch (EFB) raw materials

Unit	Value
Feed preparation-pretreatment	
Crusher (C-001/C-002)	
Desired particle size	20–40 mm
Efficiency	90%
Steam Explosion (SE)	
Temperature	210 °C
Residence time	4 min
Pressure	18.6 bars
Equation for modeling	$190 \le T \le 210,\ 4 \le t \le 15$
Solid recovery	$276 - 1.036T - 39.60t - 0.206t^2 + 0.199t$
Cellulose	$921 - 8.632T + 6.623t + 0.021182T^2 - 0.5497t^2 - 0.00372T \times t$
Hemicellulose	$-1734 + 17.18T - 23.12t - 0.04202T^2 + 0.662t^2 + 0.0887T \times t$
Lignin	$1894 - 18.28T + 3.52t + 0.04433T^2 - 0.170t^2 + 0.0095T \times t$
Ash	$0.4 - 0.007T + 1.252t + 0.00005T^2 - 0.0089t^2 - 0.00532T \times t$
Moisture content (after process)	37%
Hot-compressed water (HCW)	
Temperature	200 °C
Residence time	15 min
Pressure	30 bars
Equation for modeling	$150 \le T \le 200,\ 5 \le t \le 25$
Solid recovery	$188.5 - 0.678T - 2.70t - 0.0460t^2 + 0.0199T \times t$
Cellulose	$32.0 + 0.0794T + 1.786t - 0.02310t^2 - 0.00419T \times t$
Hemicellulose	$63.05 - 0.2397T - 7.820t + 0.026523t^2 + 0.03656T \times t$
Lignin	$53.6 - 0.217T - 0.61t - 0.00810t^2 + 0.0046T \times t$
Ash	$1.55 - 0.0032T + 0.121t - 0.00053t^2 - 0.00044T \times t$
Moisture content (after process)	37%
Alkaline hydrogen peroxide (AHP)	
Temperature	70 °C
Residence time	3 min
Pressure	1 bar
$[H_2O_2]$	3.0 wt.%
Equation for modeling	$50 \le T \le 90,\ 30 \le t \le 90,\ 1 \le C \le 5,\ m_{cellulose}$: inlet cellulose fraction
Solid recovery	$52.9 + 1.176T - 0.351t - 1.42C - 0.00966T^2 + 0.00143t^2 + 0.111C^2 + 0.00173T \times t + 0.0195T \times C - 0.0069t \times$
Cellulose of OPT	$C45.9 + 0.851T - 0.199t - 0.96C - 0.00689T^2 + 0.000611t^2 - 0.08C^2 + 0.00104T \times t + 0.0315T \times C + 0.001t \times$
Cellulose of EFB	$C(0.789 \times m_{cellulose}) + 0.851T - 0.199t - 0.96C - 0.00689T^2 + 0.000611t^2 - 0.08C^2 + 0.00104T \times t + 0.0315T \times C + 0.001t \times C$
Hemicellulose	$3.354 - 0.056T - 0.0019t + 0.123C + 0.000471T^2 + 0.000104t^2 + 0.0163C^2 - 0.000065T \times t - 0.0036T \times C - 0.000403t \times C$
Lignin	$43.49 - 0.886T + 0.1653t - 1.49C + 0.00674T^2 - 0.000744t^2 + 0.056C^2 - 0.001003T \times t + 0.00883T \times C + 0.00372t \times C$
Ash	$-0.2824 + 0.025056T - 0.012875t + 0.01528C - 0.000196T^2 + 0.000065t^2 + 0.004167C^2 + 0.00004T \times t - 0.000417T \times C$
Moisture content (after process)	10%

Table 2 Continued

Unit	Value
Autoclave (AUTOCLAVE)	
Temperature	120 °C
Residence time	20 min
Pressure	1 bar
Substrate: Buffer	1:10
Fermentation (SSF)	
Temperature	40 °C
Residence time	3,600 min
Pressure	1 bar
Ethanol yield, OPT:EFB	%w/v
75:25	3.333
50:50	3.356
25:75	3.378
0:100	3.400
Purification	
Centrifuge (CFUGE)	Ideal separation
Extractive distillation	Solvent: ethylene glycol
Beer column (T-101)	
Number of stages	22
Reflux ratio (mole)	3.17
Distillate to feed (mass)	0.038
Extractive column (T-102)	
Number of stages	18
Reflux ratio (mole)	1.60
Distillate to feed (mass)	0.54
Recovery column (T-103)	
Number of stages	11
Reflux ratio (mole)	0.25
Distillate to feed (mass)	0.73
Pervaporation	
Beer column (T-201)	
Number of stages	22
Reflux ratio (mole)	3.17
Distillate to feed (mass)	0.038
Membrane (PV-201-203)	3 units
Material	Polyvinyl chloride
Support	Polyacrylonitrile
Area	300 m^2

Data sourced from Laosiriwut et al. (2020)

Process description

Feed preparation-pretreatment

The first section of this process was feed preparation-pretreatment. Crushers (C-001/C-002) were used to reduce the size of OPT and EFB, respectively. Then, the fine particles of OPT and EFB were transferred separately to hydrothermal and chemical pretreatment units. As mentioned in the previous section, OPT and EFB were pretreated using different methods due to their different characteristics, details of which are available elsewhere. The pretreatment process aimed to increase the fraction of cellulose as the source of glucose. The yeast in this process was *S. cerevisiae*, which converts only glucose into ethanol. OPT was fed into the SE unit to destroy the rigid structure and improve enzymatic and yeast accessibility. After that, the sludge was sent to the HW unit to eliminate hemicellulose and lignin. Separately, the EFB was fed into the HCW unit, where the hemicellulose in EFB was dissolved into a liquid phase. Next, the AHP unit was used to delignified the different mixtures of OPT and EFB using hydrogen peroxide. After draining and neutralizing, the sludge was transferred to the Autoclave unit (AUTOCLAVE). The substrate was mixed with a buffer and sterilized.

Fermentation

The treated OPT and EFB were delivered to the fermentation section. The yield from the laboratory was used in the model in Aspen Plus. According to Pangsang et al. (2019) and Tareen et al. (2020), ethanol yields obtained from pure OPT and EFB in a fermenter (SSF) for 60 hr were 3.31 % (weight per volume, w/v) and 3.40 % (w/v), respectively. The chemical reaction in this fermenter is shown in Equations 1 and 2. However, because the fermenter was a simultaneous type and used *Saccharomyces cerevisiae*, the hydrolysis and fermentation occurred simultaneously.

$$\underset{\text{Cellulose}}{(C_6H_{10}O_5)_n} + nH_2O \rightarrow \underset{\text{Glucose}}{nC_6H_{12}O_6} \quad (1)$$

$$\underset{\text{Glucose}}{C_6H_{12}O_6} \rightarrow \underset{\text{Ethanol}}{2CH_3CH_2OH} + 2CO_2 \quad (2)$$

Purification

This section involved the dehydration of ethanol. The effluent stream from SSF containing ethanol, water, the remaining solids and other components was fed into a centrifuge (CFUGE) to separate the solid and liquid phases. Then, the liquid stream was transferred to the first fractional distillation column to distill ethanol closer to the azeotrope fraction. Concentrated ethanol was obtained in a distillate stream that was sent to dehydrate to fuel grade using extractive distillation or pervaporation. In principle, these two techniques applied different concepts. Extractive distillation requires a third material to dissolve the target and provide an evaporative gap while pervaporation applies pressure to drive the target component through the membrane. The configurations of the two dehydration technologies are provided in Table 2 and the process flowsheets are illustrated in Fig. 1.

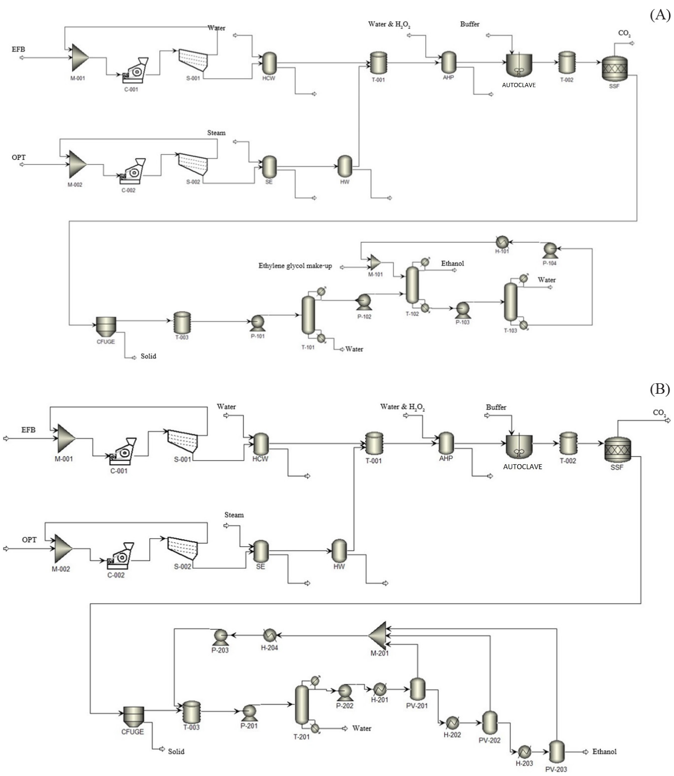

Fig. 1 Process flowsheet of bioethanol production from oil palm trunk (OPT) and empty fruit bunch (EFB) raw materials: (A) extractive distillation; (B) pervaporation, where components are defined in Table 2

Batch scheduling

Batch units were scheduled to supply the purification process continuously. The batch units (SE, HW, HCW, AHP and AUTOCLAVE) were located in the pretreatment and fermentation sections. The schedule of the bioethanol process with minimum cycle time was optimized using ABPD, involving full-scale manufacturing with one proven recipe-based modeling technology. Scheduling of bioethanol production step by step was designed and prepared as a recipe. The process was selected as a generic plant to design at the commercial scale and the recipe of scheduling was separated into two sections for pretreatment and fermentation. The residence times for the pretreatment from the laboratory for SE, HW, HCW, AHP and AUTOCLAVE were 4 min, 30 min, 15 min, 30 min and 20 min, respectively. Other command times, for example, charge time, transfer time, heating time and pH adjusting time were assumed to be 15 min, 15 min, 15–30 min and 30 min, respectively. For the fermentation recipe, the optimal fermenting time in SSF was 3,600 min. for the preliminary design of this process, two intermediate storage tanks were considered. The first tank (T-01) was to collect the pretreated OPT and EFB from HW and HCW, while the other one (T-02) was to collect after autoclaving.

Economic assumption

To carry out a techno-economic analysis (TEA) of the bioethanol production, total capital investment (TCI) and total product cost (TPC) were evaluated as the main components of economic evaluation. Internal rate of return (IRR), payback period and net present value (NPV) were selected to determine the feasibility of this process. The key assumptions of parameters for a solid-fluid process are provided in Tables 3A, 3B, 3C. Importantly, this process was designed to support various ratios of OPT and EFB, so the equipment size was calculated for the designed ratio. However, this economic evaluation was performed based on selling gasohol E20 (a mix of 80% gasoline and 20% ethanol) as the main product. The results were determined using the same approach as Do et al. (2014), although the tax basis was slightly different.

Table 3A Key assumptions for total capital investment estimation

Component	Percentage
Fixed capital investment (FCI)	
1. Direct costs	
Purchased equipment	100
Purchased equipment installation	39
Instrumentation and control (installed)	13
Piping (installed)	31
Electrical systems (installed)	10
Building including services	29
Yard improvement	10
Service facilities (installed)	55
Land	6
2. Indirect costs	
Engineering and supervision	32
Construction expenses	34
Contractor's fee	18
Contingency	36
Working capital (WC)	15% of FCI
Total capital investment	FCI+WC

Table 3B Key assumptions for total product cost estimation

Component	Percentage
Direct manufacturing costs (C_{DM})	
Raw material	The unit cost of each material × Total used unit
Utilities	Electricity cost + Heat cost + Cooling cost
Operating labor	Wage rate × No. of operator
Maintenance and repairs	2% of FCI
Operating supplies	0.5% of FCI
Fixed manufacturing costs (C_{FM})	
Local taxes	1% of FCI
Insurance	0.4% of FCI
Plant-overhead costs	5% of TPC
General expenses (GE)	
Administrative costs	5% of TPC
Distributions and selling costs	1% of TPC
Total product cost	$C_{DM} + C_{FM}$ +GE

Table 3C Other key assumptions for techno-economic analysis

Component	Value
Raw material and product price	
OPT (USD/t)	1.00
EFB (USD/t)	1.56
Hydrogen peroxide (USD/t)	810
Ethylene glycol (USD/t)	830
Gasoline (USD/L)	0.490
Gasohol E20 (USD/L)	0.840
Utility price	
Water (USD/t)	0.010
Electricity (USD/MJ)	0.015
Steam (USD/t)	12
Annual working hours	7,200
TAX	30%
Working capital	5%
Depreciation method	Straight line
Salvage of equipment	10% Purchased equipment cost
Plant lifetime	20 yr
Equipment lifetime	20 yr
Weighted average cost of capital	7%

Simulation of dynamics and control

In the purification process of bioethanol production, the dynamic and control systems were performed using APD exported from Aspen Plus. A disturbance rejection was applied in the purification process. The dynamic and control system was implemented to study the ability of the process to control the product specification to a setpoint. Disturbances could affect the process when changing the ratio of OPT and EFB, such as the ethanol concentration and feed flow rate to the purification process. The proportional, integral and derivatives actions of the installed controllers were modified using tuning methods to make the process achieve a new steady state as quickly as possible.

Results and Discussion

Effect of ratio of oil palm trunk-to-empty fruit bunch as raw materials

The ratio of OPT and EFB affected the process since they had different compositions, including the pretreatment procedures. The effects of the feedstock ratios of OPT and EFB are demonstrated in Table 4. To obtain 10,000 L ethanol/d, required 2,000 kg/hr or 48,000 kg/day of OPT and EFB. When the EFB ratio increased, the production rate of ethanol also increased. The ratio at 75:25 provided the highest amount of cellulose in the fermenter feed stream at 721.86 kg/hr. The cellulose content in the fresh feedstocks increased from 38.67–38.85 wt% up to 70.37 wt%. Nevertheless, the highest ethanol yield was obtained at feeding 100% EFB. After the ethanol dehydration process, the highest production rate was 11,593 L/d from feeding 100% EFB using pervaporation and the production rate was 1.95% higher compared to extractive distillation. The production rate increased as the EFB ratio increased.

Table 4 Effect of feedstock ratio of oil palm trunk (OPT) and empty fruit bunch (EFB) on production rate

OPT:EFB	Feedstock (kg/d)		Ethanol product (L/d)	
	OPT	EFB	Extractive distillation	Pervaporation
75:25	36,000	12,000	11,361	11,585
50:50	24,000	24,000	11,365	11,589
25:75	12,000	36,000	11,367	11,591
0:100	-	48,000	11,369	11,593

The energy requirement for the purification processes is shown in Fig. 2. Energy consumption for bioethanol production using extractive distillation or pervaporation remained constant as the ratio of EFB was increased. Pervaporation consumed less energy than extractive distillation. The energy consumption per production rate of pervaporation was 16.49 MJ/L, while extractive distillation consumed 20.27 MJ/L. Most of the energy was consumed in the reboiler of the beer column used to recover ethanol from the broth.

However, the product must be sold as gasohol E20, so the amounts of required gasoline and gasohol E20 production are shown in Table 5.

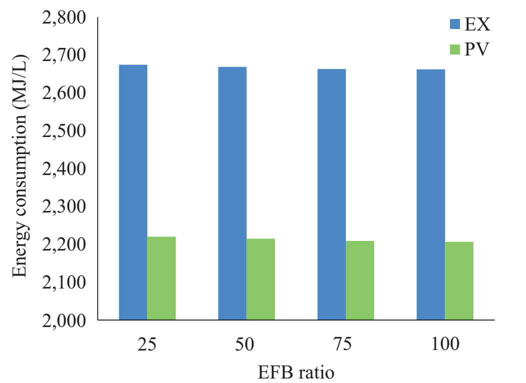

Fig. 2 Comparison of energy consumption of extractive distillation (EX) and pervaporation (PV) at different empty fruit bunch (EFB) ratios

Table 5 Gasoline requirement and Gasohol E20 production feedstock for different feedstock ratios of oil palm trunk (OPT) and empty fruit bunch (EFB)

OPT:EFB	Gasoline (L/day)		Gasohol E20 (L/day)	
	Extractive distillation	Pervaporation	Extractive distillation	Pervaporation
75:25	45,445	46,342	56,806	57,927
50:50	45,459	46,357	56,824	57,946
25:75	45,469	46,366	56,836	57,957
0:100	45,475	46,371	56,844	57,964

Scheduling and production plan

Scheduling of pretreatment and fermentation was generated in ABPD, where the Gantt chart indicated that a cycle took 4,089 min (cycle time) to obtain ethanol. The number of pretreatment batches was designed for two cycles with a batch overlap time. In terms of fermentation batch, the required number of parallel SSF tanks was 15 (6 tanks/d×2.5 d or 3,600 min) based on 50,000 L/tank with 6 batches started and completed every day. Scheduling removed bottlenecks between the fermentation and purification sections. The completed Gantt chart for OPT and EFB and the layout from ABPD are shown in Figs. 3 and 4, respectively. The current study extended Sirikanchittavon et al. (2018) by covering different raw material and overall scenarios.

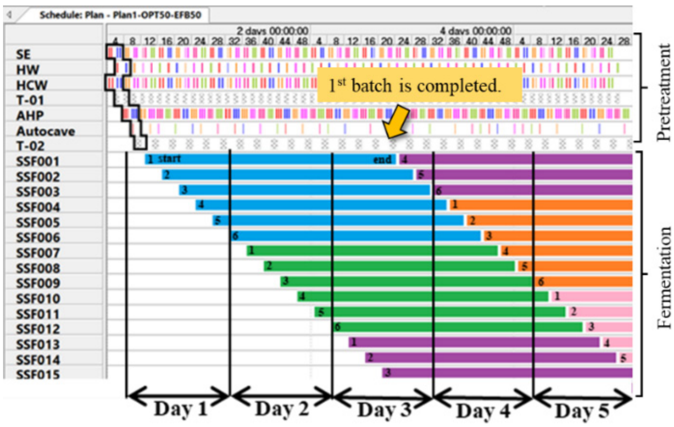

Fig. 3 Completed Gantt chart from full-scale manufacturing with one proven recipe-based modeling technology (ABPD)

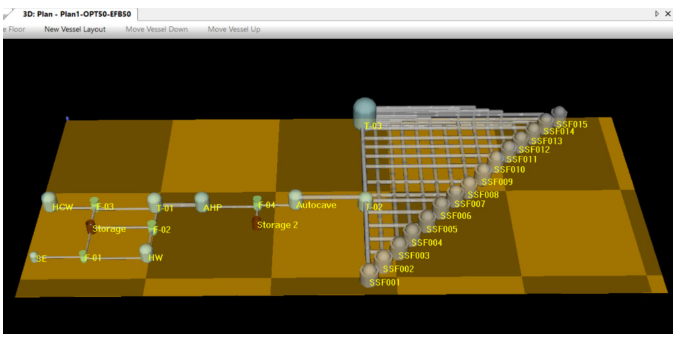

Fig. 4 Plant layout for full-scale manufacturing with one proven recipe-based modeling technology (ABPD)

Economic analysis

The equipment size was determined and the purchase costs were estimated. The major equipment, such as the crusher, pump, heat exchanger and distillation column, were sized and their purchase costs estimated using Aspen Process Economic Analyzer version 11. The summarized cost of bioethanol production from OPT and EFB using extractive distillation and pervaporation is provided in Table 6. Bioethanol production using pervaporation required a higher equipment purchase cost than extractive distillation.

The total capital investment (TCI) was estimated, associated with the construction of the plant. The fixed-capital investment (FCI) and working capital (WC) are summarized in Fig. 5, based on calculations using key assumptions for a solid-fluid process, as previously mentioned. The Chemical Engineering Plant Cost Index (CEPCI) was used to convert the total equipment cost from the reference year to 2020. The TCI amounts for bioethanol production based on extractive distillation and pervaporation were USD 21,292,092 and USD 21,823,680, respectively. Bioethanol production with pervaporation was more expensive than extractive distillation because of the 3% higher purchased equipment cost for pervaporation.

The total product cost (TPC) was estimated based on the direct manufacturing cost, fixed manufacturing cost and general expenses. The feedstock ratio was important because of the number of raw materials and utility requirements. As shown in Figs. 6A and 6B, the TPC increased with increasing EFB, since the price of EFB was higher than for OPT in both cases. Bioethanol production using pervaporation was more profitable than from extractive distillation and as the amount of EFB decreased, the profit increased.

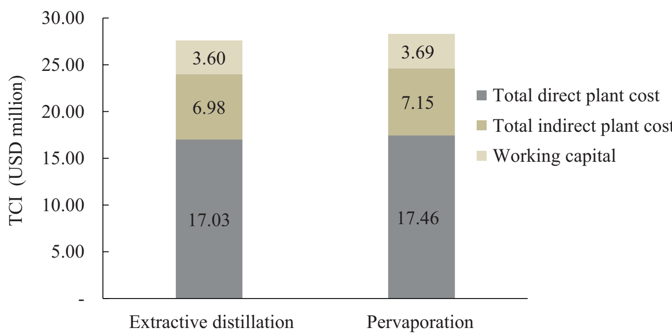

Fig. 5 Total capital investment (TCI) for bioethanol production

Table 6 Summary of purchased equipment costs

Section	Purchased equipment cost (USD)	Total purchased equipment cost (USD)
Feed preparation and pretreatment	4,824,492	-
Extractive distillation	988,254	5,812,747
Pervaporation	1,133,378	5,957,870

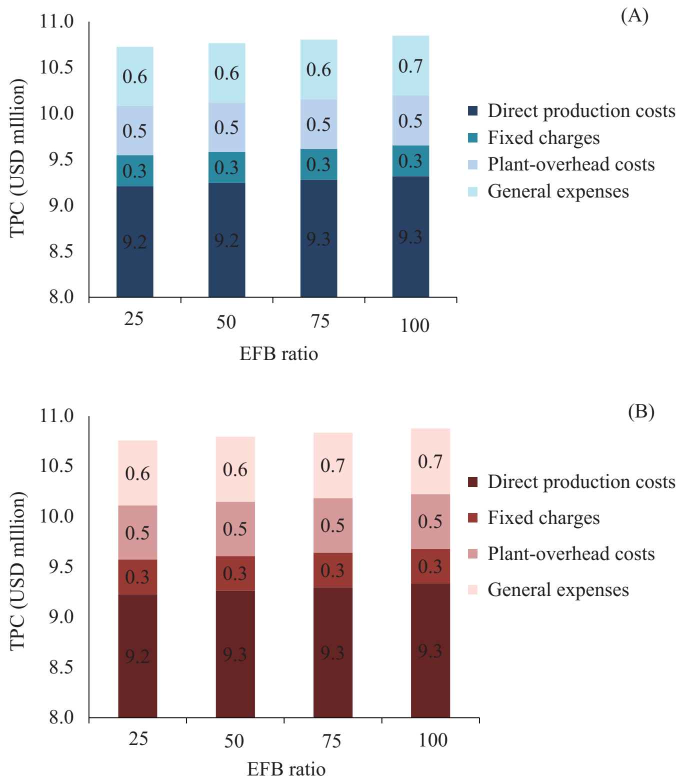

Fig. 6 Total production cost (TPC) of bioethanol production at different empty fruit bunch (EFB) ratios: (A) extractive distillation; (B) pervaporation

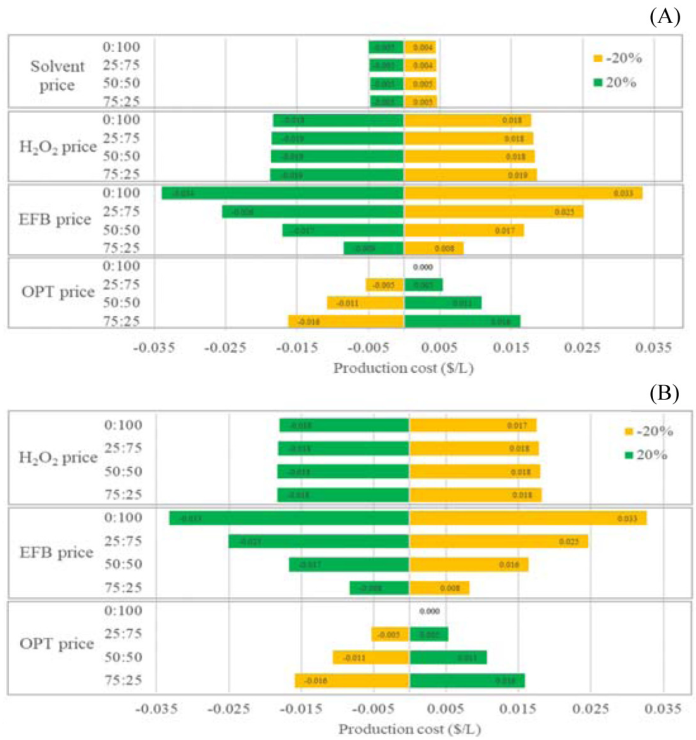

Fig. 7 Bioethanol production cost: (A) extractive distillation; (B) pervaporation for different feedstock ratios of oil palm trunk (OPT) and empty fruit bunch (EFB), where all prices in US dollars

The results clearly showed that the TCI amounts using extractive distillation and pervaporation were constant for the various ratios of OPT and EFB. When the EFB ratio increased, the TPC increased, while the profits from using either extractive distillation or pervaporation decreased. For these reasons, economic parameters were used as decision indicators, namely the internal rate of return, payback period and net present value. The parameters were calculated at various ratios of the feedstocks. As shown in Table 7, the net present value and the internal rate of return of bioethanol production with pervaporation were higher than for extractive distillation, accompanied by a minimum payback period.

If the plant were operated using pervaporation and the supply of OPT varied seasonally, the net present value, internal rate of return and payback period would vary in the ranges USD 4,560,173–5,380,867, 9.00–9.35% and 14.66–15.39 yr, respectively. A ratio of OPT-to-EFB of 75:25 produced the best economic performance, with the highest net present value, internal rate of return and the lowest payback period. These results indicated that bioethanol production from OPT and EFB using pervaporation was feasible for investment.

Table 7 Economic parameters for bioethanol production for different feedstock ratios of oil palm trunk (OPT) and empty fruit bunch (EFB)

OPT:EFB	Extractive distillation			Pervaporation		
	NPV (USD)	IRR (%)	PBP (yr)	NPV (USD)	IRR (%)	PBP (yr)
75:25	4,070,435	8.83	15.81	5,380,867	9.35	14.66
50:50	3,812,858	8.72	16.08	5,123,960	9.24	14.89
25:75	3,548,883	8.61	16.36	4,860,633	9.13	15.12
0:100	3,248,715	8.47	16.74	4,560,174	9.00	15.39

The production costs at the various ratios using pervaporation were in the range USD 1.10-1.12/L ethanol, for extractive distillation they were in the range USD 1.14–1.18/L ethanol, as shown in Table 8. These results confirmed that pervaporation should be selected as the purification process. The uncertainty of raw materials and utility prices was investigated using ±20% of the solvent, hydrogen peroxide, EFB and OPT prices. In bioethanol production using extractive distillation, the uncertainties associated with the solvent, hydrogen peroxide, EFB and OPT prices were in the ranges USD 1.155–1.165/L, USD 1.141–1.179/L, USD 1.126–1.193/L and USD 1.144–1.176/L, respectively. For bioethanol production using pervaporation, the uncertainties associated with the hydrogen peroxide, EFB and OPT prices were in the ranges USD 1.092–1.128/L, USD 1.077–1.143/L and USD 1.094–1.126/L, respectively.

Process dynamics and controllability

The dynamics and control system results of the purification process were analyzed. The economic results indicated that pervaporation was the most suitable technology for ethanol dehydration. Therefore, the controllability of the pervaporation process was studies. The controlled and manipulated variables are described in Table 9, with the dead time for temperature at stage 19 and in composition measurement were assumed to be 1 min and 5 min, respectively. The process flowsheet for the controller system is illustrated in Fig. 8. The tuning parameters, including the Proportional Integral Derivative (PID) controller action, are provided in Table 10. The tuning method was selected based on disturbance rejection, type of loop and minimum time to reach a new steady state. The studied disturbances in this section were feed flow rate and concentration of ethanol in broth. The feed flow rate was varied from 10,227±500 kg/hr while the concentration of ethanol was varied from 3.37±1.00 %w/v, made at time = 0.2 hours. The manipulated variable to control the product specification was selected as the retentate feed temperature. Besides, it indicated that the controlled composition of product specification should be the impurity; the composition of water in the product stream is selected to be the controlled variable in this study. Furthermore, the three retentate temperature controllers were operated on a cascade receiving a setpoint signal from CC-201. The results indicated that the controllers could control the process variable back to setpoint. The dynamics of the impurity in the product stream were drawn in Figs. 9 and 10. When the process faces disturbances in feed flow rate and ethanol concentration in broth, the controllers rejected the disturbances and go back to the setpoint. As a result, the settling time was 9.0 hours. It proved that when the broth flow rate was changed, the controller can work without retuning.

Table 8 Production cost of bioethanol production for different feedstock ratios of oil palm trunk (OPT) and empty fruit bunch (EFB)

OPT:EFB	Production cost (USD/L ethanol)	
	Extractive distillation	Pervaporation
75:25	1.14	1.10
50:50	1.16	1.11
25:75	1.17	1.12
0:100	1.18	1.12
Average	1.16	1.11

Conclusions

Bioethanol production was investigated using the oil palm trunk and empty fruit bunch as feedstocks. In the purification process, 99.0 wt% ethanol was attained using both extractive distillation and pervaporation. The highest ethanol production rate was 11,593 L/day with feeding only EFB and using pervaporation as the purification technology. For the minimum batch cycle time, the pretreatment was operated on two cycles and fermentation was separated into 15 tanks based on 50,000 L/tank, with 6 batches started and completed every day. Scheduling was able to remove the bottleneck between the fermentation and purification sections.

Table 9 Controlled and manipulated variables

Controller	Manipulated variable	Controlled variable
FC-201	%Open valve	Flow control
LC-201	Bottom flow rate	Base level
LC-202	Distillate flow rate	Reflux drum level
PC-201	Condenser duty	Column pressure
PC-202	Speed of the distillate pump	Retentate pressure
TC-201	Reboiler heat input	The temperature at stage 19
TC-202	Heat input of H-201	Inlet retentate temperature
TC-203	Heat input of H-202	Inlet retentate temperature
TC-204	Heat input of H-203	Inlet retentate temperature
TC-205	Heat input of H-204	Outlet temperature
CC-201	Setpoint of TC-202-204	Composition of the retentate product (Cascade control TC-202-204)

Table 10 Tuning parameters

Controller	Open loop gain (%/%)	Time constant (min)	Deadtime (min)	Prerational gain (%/%)	Integral time (min)	Derivative time (min)	Action	Method
				Open-loop control				
PC-202	2.54	0.60	1.20	0.20	1.61	0.00	Reverse	IAE
TC-202	0.29	0.60	1.20	3.20	1.82	0.32	Reverse	CHC
TC-203	0.31	0.51	1.80	2.04	2.31	0.40	Reverse	CHC
TC-204	0.18	0.35	1.20	3.52	1.56	0.27	Reverse	CHC
TC-205	2.82	0.61	1.20	0.27	1.15	0.63	Reverse	IAE
				Closed-loop control				
Controller	Ultimate gain (%/%)	Ultimate period (min)	-	Prerational gain (%/%)	Integral time (min)	Derivative time (min)	Action	Method*
LC-202	206.92	1.20	-	94.05	1.00	0.00	Direct	ZN
PC-201	84.49	3.00	-	38.41	2.50	0.00	Reverse	ZN
TC-201	173.84	4.80	-	79.02	4.00	0.00	Reverse	ZN
CC-201	2.28	67.80	-	1.67	42.38	6.78	Direct	ZN

IAE = integral of absolute error method; CHC = Cohen-Coon method, ZN = Ziegler-*Nichols* method

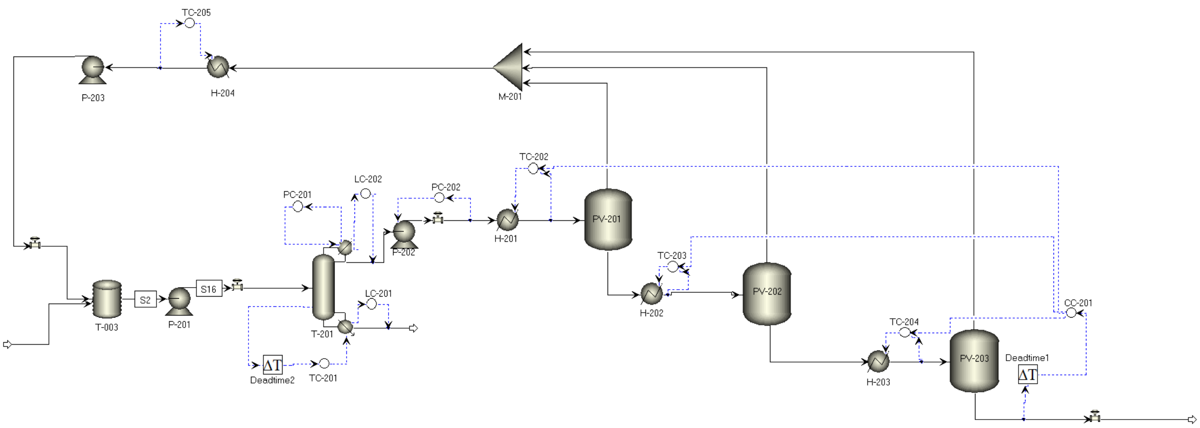

Fig. 8 Pervaporation process with the control system using the Aspen Plus Dynamics system, where components are defined in Table 2

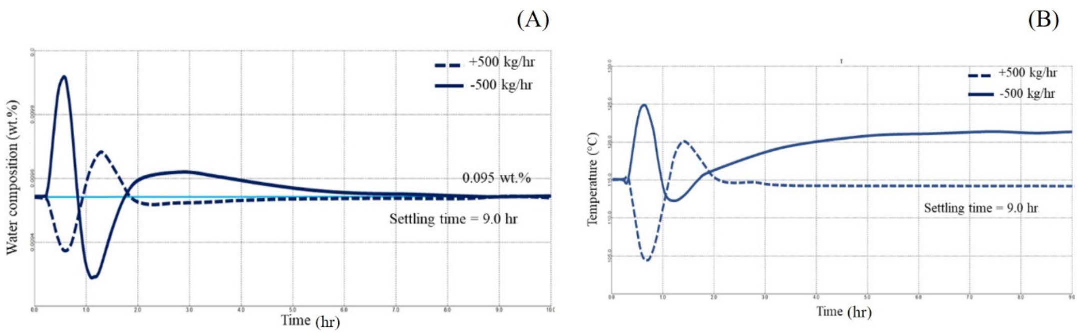

Fig. 9 Process dynamics: (A) CC-201; (B) cascade control of TC-202-204

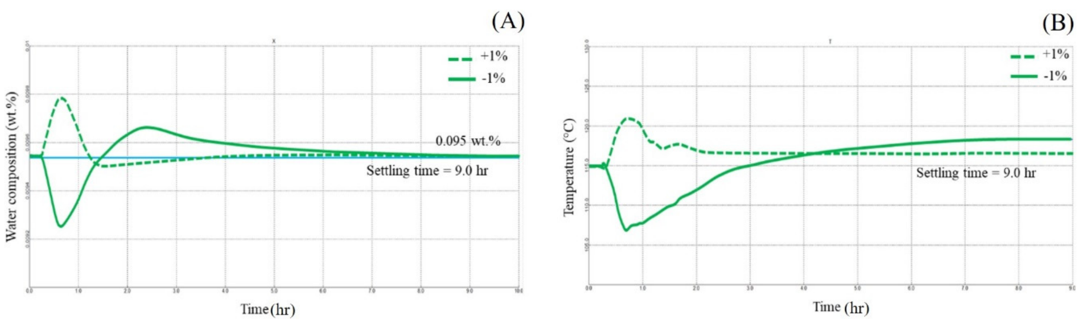

Fig. 10 Process dynamics: (A) CC-201; (B) cascade control of TC-202-204

The techno-economic analysis indicated that the total capital investment of the bioethanol production using pervaporation was higher than for extractive distillation. However, the net present value (NPV), internal rate of return (IRR) and payback period of production using pervaporation were greater based on selling gasohol E20 as the main product. Using pervaporation, the NPV, IRR and payback period were USD 4,560,174–5,380,867, 9.00–9.35% and 14.66–15.39 yr, respectively. Furthermore, feeding 75% OPT and 25% EFB was the best ratio for operation that produced values for the NPV, IRR and payback period of USD 5,380,867, 9.35% and 14.66 yr, respectively. Thus, OPT should be fed into the process as much as possible because the OPT price is lower than the EFB cost, resulting in lower total product costs. Therefore, bioethanol production was a feasible investment when sold as gasohol E20. The control system implemented in the pervaporation process was able to handle disturbances from changing the feedstock ratio, allowing PID action to remove disturbances during fermentation and drive the process variables back to the setpoint.

Conflict of Interest

The authors declare that there are no conflicts of interest.

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