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Techno-Economic Analysis of Extractive Butanol Fermentation by Immobilized Cells with Large Extractant Volume

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ABSTRAK

Terdapat beberapa tantangan fermentasi *Acetone-Butanol-Ethanol* (ABE) untuk digunakan dalam skala industri antara lain rendahnya rendemen butanol, tingginya kebutuhan energi untuk pemisahan dan pemurnian, dan persaingan gula dengan kebutuhan pangan sebagai substrat. Penelitian ini mempelajari aspek teknik dan ekonomi dari fermentasi ABE menggunakan sel amobil dengan volum ekstraktan besar. Keseluruhan proses produksi dirancang menggunakan bahan baku jerami padi yang dihidrolisis tak sempurna untuk menghasilkan campuran selobiosa, glukosa, xilosa, dan arabinosa. Gula konsentrat kemudian diumpankan ke fermentasi *fed-batch* ekstraktif menggunakan sel amobil. Akhirnya, ekstraktan diperoleh kembali dan produk dimurnikan dengan kolom distilasi. Dengan mengevaluasi desain proses ini untuk kapasitas skala kecil 238 kg-butanol dan aseton/hari, kebutuhan energi adalah 41,3 MJ/kg-butanol dan aseton dan biayanya adalah 1,91 $/kg-butanol dan aseton. Meskipun biayanya lebih tinggi daripada butanol yang dihasilkan oleh proses petrokimia sebesar 1,08 $/kg-butanol, biayanya dapat berkurang jika skalanya ditingkatkan.

***Kata kunci*:** ABE, Fermentasi, Gula, Sel amobil, Teknik-ekonomi.

ABSTRACT

There are several challenges for ABE fermentation to be used in an industrial scale including the low of butanol yield, the high energy requirement for separation and purification, and the competeness of sugar with food demand as substrat. In this study, techno-economical aspects of ABE fermentation by using immobilized cells with large extractant volume were studied. Overall production process was designed using rice straw as raw material which is semi-hydrolyzed to produce cellobiose, glucose, xylose, and arabinose mixture. Concentrated sugar was then fed to extractive fed-batch fermentation using immobilized cells. Finally, extractant was recovered and products were purified by distillation column. By evaluating this process design for the small scale capacity of 238 kg-butanol and acetone/day, the energy requirement was 41.3 MJ/kg-butanol and acetone and the cost was 1.91 $/kg-butanol and acetone. Although the cost was higher than butanol produced by petrochemical process of 1.08 $/kg-butanol, it may reduce if the scale is increased.

***Keywords*:** ABE, Fermentation, Immobilized cell, Sugar, Techno-economic.

1. INTRODUCTION

Butanol production process has been developed from the period of World War. ABE fermentation was firstly used to produce acetone to be used as a solvent in the production of cordite [1]. Currently, the most widely used process to produce butanol was Oxo process that is petrochemical based using propene as raw material. Along with the higher concern of depleting petroleum

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sources and its increasing prices, biobased process using sustainable or renewable materials has become attractive for both research development or commercial production [2]. ABE fermentation provides several advantages over the petrochemical production process. Various carbon source can be utilized by ABE producer strain, including various sugars of monosaccharide, oligosaccharide [3], and polysaccharide [4], various organic acids such as butyric acid [5], lactic acid [6], and acetic acid [7], carbon dioxide [8]. Besides, biobased production process does not require severe operating condition of high temperature or pressure such used in Oxo process.

There are several challenges for ABE fermentation to be used in an industrial scale. The low titer of butanol resulted from fermentation required high energy for separation and purification of the products. Therefore, simultaneous separation and fermentation was development to reduce the energy requirement after fermentation [9]. Another challenge is that if sugar is used as raw material, the feedstock demand of ABE fermentation will compete with food demand. In order to overcome this challenge, sugar can be produced from the waste of agriculture [10] or other industrial process [11].

Several industry using ABE fermentation has been recently operated. The butanol price was still higher compared with gasoline price in the market. However, the price is reasonable to be sold as butanol for chemical industry [12]. The production process of butanol using fermentation requires lower energy because of the operation near ambient temperature. The energy consuming process in biobased butanol production process is separation. Therefore the concern to optimize this process increased with various separation method considered.

Extraction was evaluated to be one of the process with low energy needs of 7.7 MJ/kg butanol. Compared with other effective butanol separation techniques of gas stripping, pervaporation, and adsorption which consumed energy of 31 MJ/kg, 145 MJ/kg, and 33 MJ/kg butanol respectively [13]. The further optimation has been developed to reduce the energy requirement using extraction method. Extraction and distillation series were modified to several methods and have been evaluated for the energy requirement [14]: high temperature extraction at 80 ℃ using mesytilene which required energy of 3.6 MJ/kg butanol [15], dual extraction method which required energy of 2.6 MJ/kg of butanol [16], and extraction to direct steam distillation which required energy of 2.6 MJ/kg butanol. Direct steam distillation method has an advantage over the dual extraction method with less unit operation needed and operation at ambient pressure. However the direct steam streamed into distillation column could increase the water content in the product which is not desirable. Therefore, the case in this study will be studied based on its properties to decide the suitable recovery method.

A limited number of studies have reported the energy requirement for overall production of butanol. Qureshi and Blaschek [11] reported that the butanol production using ABE fermentation and corn steep liquor as raw material required 116 MJ/kg butanol. It is still higher than the energy required for the for the butanol production using Oxo process of 69 MJ/kg butanol [17]. Previous studies reported that extraction-distillation process consumed the least total cost of 0.27 $/kg-butanol, compared with conventional distillation-distillation process required 0.34 $/kg-butanol, and pervaporation-distillation process which required 0.35 $/kg-butanol [18]. Overall production of petrochemical butanol using Oxo process required the cost of 0.915 $/kg-butanol [17]. While overall biobased production of butanol using ABE fermentation and corn steep liquor as raw material resulted in 1.041 $/kg-butanol [11].

The objective of this study is to evaluate the designed process based on the study performed in previous study. The production

process consisted of sugar production from rice straw using semi-hydrolysis, extractive fed-batch fermentation using immobilized cells at a large extractant volume, and distillation recovery.

2. RESEARCH METHODS

2.1. Designed Process Determination

The production process of butanol designed in this work is based on the recent studies [19–22]. The selected raw material is rice straw which was pretreated by low concentration of sulphuric acid to reduce inhibitor components. The pretreated rice straw was then processed by semi-hydrolysis using cellulase [10, 23, 24].

The sugars obtained from semi-hydrolysis was then fed in ABE fermentation. The mode used in this study was extractive fed-batch fermentation with immobilized cells at a large Ve/Vb. The extractant used in this fermentor was oleyl alcohol – tributyrin mixture 1:1 (v/v) and the immobilization used calcium alginate beads entrapment. The three phases mixture was filtered and decanted. Product-rich extractant was continued to the recovery and purification process, while cell beads and media are recycled to the fermentor.

The extractant was recovered with distillation technique. It was selected because the mixture of oleyl alcohol, tributyrin, butanol, and acetone did not contain any azeotropic mixture. Based on the boiling point difference, it was decided that two distillation column were required: first distillation column to separate oleyl alcohol – tributyrin mixture from butanol – acetone mixture and the second column to separate butanol and acetone.

2.2. Mass Balance of The Designed Process

Mass balance of the overall process was calculated using two different methods. Mass balance for sugar production and fermentation was calculated from the result of the study as mentioned in section 2.1. Mass balance of extractant recovery and product purification was calculated by Aspen Plus v10® simulation software with NRTL method.

The amount of rice straw fed as raw material is based on the rice straw production in Japan 13.5 Mt/year. Assumed that the production plant took the supply from 3 closest prefectures to the plant, 100 kg/h flow rate of rice straw was obtained. The production of rice straw was supplied every 2 months [25]. Rice straw powder was pretreated with solid to liquid ratio of 1:8 (w/w) and 1% (v/v). All pentose was produced during the pretreatment, and small amount of hexose was produced. The result of pretreatment contained hexose of 2.85 g/L and pentose of 17.2 g/L. The pretreated rice straw was semi-hydrolyzed with 2.61 FPU cellulase/g rice straw and the product from this process contained 7.28 g/L cellobiose, 6.24 g/L glucose, 14.5 g/L xylose, and 2.72 g/L arabinan. The total conversion was 39.8% for glucan, 100% for xylan, and 96.6% for arabinan [10].

The concentrated sugar obtained from previous process was used for preculture and main culture fermentation. Preculture was performed using 20 g/L sugar and main culture fermentation used total sugar concentration of 267 g/L. Assumed that no CCR occurred, all sugar was consumed after 96 h and solvent was produced with total concentration 64.6 g/L-broth for butanol, 30.6 g/L-broth for acetone, and 2.10 g/L-broth for ethanol. The distribution of the products in extractant and aqueous phase was calculated using Kd of 3.14, 0.66, and 0.00 for butanol, acetone, and ethanol, respectively.

Product-rich extractant was fed to product purification and extractant recovery. The operating condition of extractant recovery distillation column was 1 atm pressure, 100 ℃ entering temperature. Butanol is used as the light key at the distillate with distillate to feed recovery of 1.00. The product

purification column was set at 1 atm pressure and entering temperature following the temperature of distillate from extractant recovery distillation. Butanol was used as the heavy key with bottom product to feed recovery of 0.99. Reflux ratio, number of stages, and entering stages was optimized to obtain 100% extractant recovery and 99.5% (w/w) butanol product concentration.

2.3. Heat Integration and Utility Requirement

Heat integration was arranged by listing the energy transferring streams and specified into two types: cold stream, which needed to increase the temperature and hot stream, which needed to decrease the temperature. Assumed that minimum temperature difference was 10 ℃. Temperatures of supply and target of hot stream were determined by Equation 1 and 2.

$$T_S^* = T_i - \frac{1}{2}\Delta T_{min} \quad (1)$$

$$T_t^* = T_o - \frac{1}{2}\Delta T_{min} \quad (2)$$

Temperatures of supply and target of cold stream were by Equation 3 and 4.

$$T_S^* = T_i + \frac{1}{2}\Delta T_{min} \quad (3)$$

$$T_t^* = T_o - \frac{1}{2}\Delta T_{min} \quad (4)$$

Where T_S^* is the temperature of supply, T_t^* is the temperature of target, T_i is the inlet temperature, T_o is the outlet temperature, and ΔT_{min} is minimum temperature difference. All temperatures were calculated in degree Celcius (℃).

Then stream population was lined in and enthalpy changes in each temperature interval was calculated by Equation 5.

$$\Delta H_{interval} = \Delta T \times \left(\sum Cp_C - \sum Cp_H\right) \quad (5)$$

Where $\Delta H_{interval}$ is the enthalpy changes in each interval (MJ/h), ΔT is the temperature difference between interval (K), $\sum Cp_C$ is the total specific heat of cold streams (kJ/kg/K) and $\sum Cp_H$ is the total specific heat of hot streams (kJ/kg/K). The value of specific heat of the liquid and vapour mixtures was estimated using Aspen Plus v10® simulation software with NRTL method. An exception for the specific heat of rice straw which was not available in the simulation, it was obtained from the reference with the value of 1.67 kJ/kg/K [26].

Minimum energy requirement and the pinch point for overall plant were calculated using cascade diagram. The energy transferring streams was then exchanged with other available streams to obtain or release the heat to reach the required temperature. The remaining streams that couldn't be exchanged with others due to insufficient heat or temperature difference were facilitated with energy transfer from utility.

Hot utility was provided by furnace with rice straw biomass as fuel. The heating value of rice straw was obtained from reference [27] with the value of 23.2 MJ/kg. Cold utility was provided by cooling water with the inlet temperature of 25 ℃ and outlet temperature of 37 ℃.

2.4. Cost Estimation

Total cost of the designed plant was estimated by the following equation:

$$C_{TC} = C_{TCI} + C_{TOC}$$

Where C_{TC} is total cost, C_{TCI} is total capital investment, and C_{TOC} is total operating cost.

The capital investment cost estimation was calculated according to the size and specification of the equipment used in overall process including utility. The size and specification of the main equipment were estimated according to the method and approach from Ludwig [28] and Walas [29]. The specification of the pumps were calculated based on the data from Branan [30] and the specification of heat exchanger was calculated based on the method from Ludwig [28].

The cost of all equipments were then calculated with the following Equation 6.

$$cost(a) = cost(b)\left(\frac{cap(a)}{cap(b)}\right)^{X} \qquad (6)$$

Where *cost(a)* is the cost of equipment a, *cost(b)* is the reference cost of the equipment b, *cap(a)* is the capacity of equipment a, *cap(b)* is the reference capacity of equipment b, and *X* is the exponential factor. The data of *cost(b), cap(b),* and *X* were obtained from Peters and Timmerhaus [31].

Total operating cost was calculated based on the recent price of the materials and the annual consumption. Total cost was then used to determine the specific cost of butanol and acetone product.

3. RESULTS AND DISCUSSION

3.1. Optimum Operating Conditions of Immobilized Extractive Fermentation

The series of production process selected in the extractive fed-batch fermentation using immobilized cells is shown in Figure 1. Dry rice straw was used as a lignocellulosic raw material in the bulk form which is grinded in a milling machine to rice straw powder. Rice straw powder is mixed with water and low concentration of sulphuric acid for pretreatment. The purpose of this process is eliminating inhibitor component from the rice straw which is originated from lignin content. Pretreated rice straw mixture was then proceed in a semi-hydrolysis unit using cellulose enzyme to convert cellulose and hemicellulose to sugars. This process resulted in monosaccharide and oligosaccharide mixture. The output from semi-hydrolysis unit contained unconverted rice straw, thus the mixture is filtered in a filtration unit to separate solid and liquid phase. Unconverted rice straw as the filter residue was recycled and returned to the semy-hydrolysis reactor to be proceed again. Sugar mixture in water solution as the filtrate was flown to an evaporation unit in order to increase the sugar concentration by evaporating certain amount of water.

Concentrated sugar mixture obtained from evaporator was fed to the extractive fed-batch fermentation unit. This process consisted of two culture: pre-culture to grow inoculates after and main culture units. Tryptone – Yeast extract – Acetate (TYA) was added as a nutrient medium to grow *C. saccharoperbutylacetonicum* N1-4. This inoculate was then immobilized in calcium alginate beads to be used repeatedly along the production process. Main culture consists of three phases of extractant, medium broth, and immobilized cell beads. During main culture process, various sugars contained in concentrated sugar solution were fermented to butanol, acetone, and ethanol as main solvent products. Butanol and acetone were extracted simultaneously during fermentation. Therefore, butanol and acetone concentration in the medium broth will be maintained in a lower concentration than toxicity limit. While extractant contain higher concentration of solvent products to be processed further.

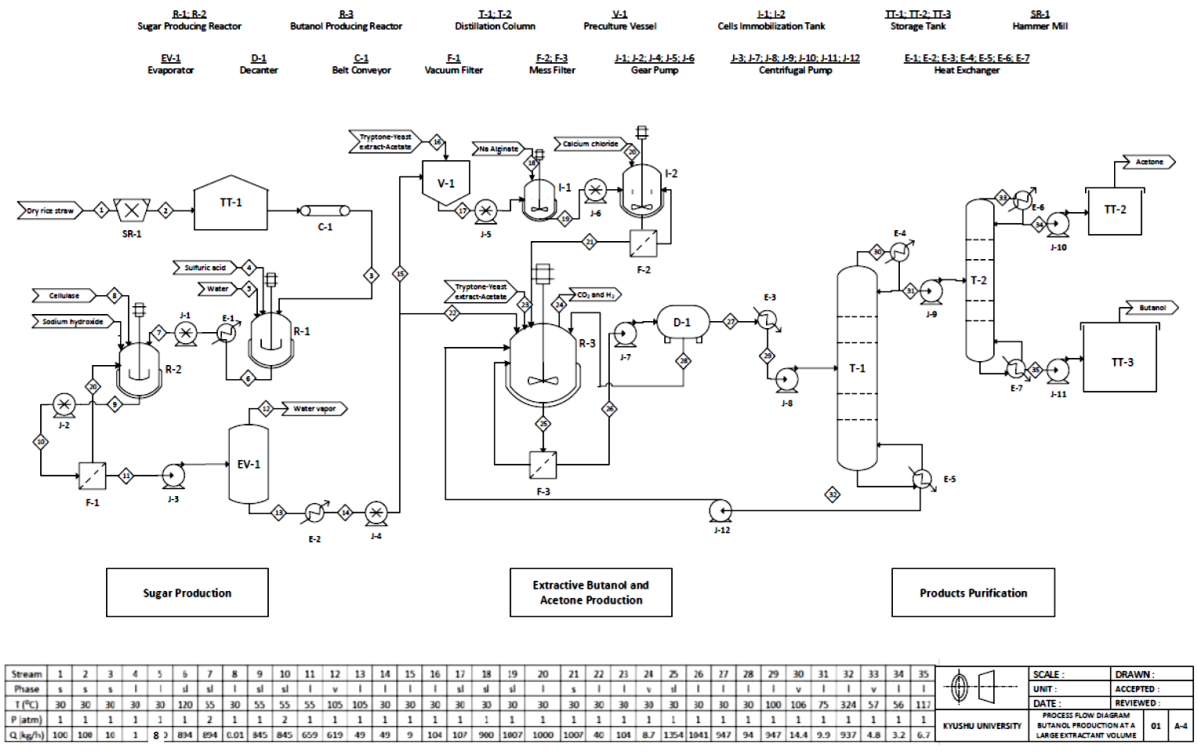
The output mixture from extractive fermentation was then filtered to separate the cell beads from liquid to be used for next cycle of fermentation. The liquid mixture was separated in a decantation unit to product rich extractant and medium broth. The medium broth was recycled to fermentation unit to be used and added with new nutrient for the next cycle. Product rich extractant was recovered in the distillation unit separating it from the extracted butanol and acetone. The butanol and acetone was purified in the last distillation column to reach the desired grade of 99.5% butanol and 99.0 % acetone.

Figure 1 is the process flow diagram showing the operating condition of the processes. Bulk rice straw was grinded using hammer mill (SR-1) and then stored in a warehouse (TT-1). The size of rice straw powder is approximately 0 – 3 mm. Using belt conveyor, the rice straw powder was fed into a pretreatment reactor (R-1). All processes in this production were operated in the atmospheric pressure accept for this pretreatment reactor because it is operated in a high temperature containing water. The temperature of pretreatment reactor was 120

ºC, in the pressure of 2 atm, using 1% *v/v* of sulphuric acid (H_2SO_4) for 30 min. The pretreated rice straw mixture was then cooled down to 55 ºC, it is processed in semi-hydrolysis reactor (R-2) operating at 55 ºC, pressure 1 atm, for 12 h using cellulase enzyme of 2610 FPU/kg rice straw. The process is continued with filtration (F-1) then evaporation (EV-1). Evaporation unit is operated at the temperature of 105 ºC, pressure 1 atm.

Concentrated sugar was fed to the preculture tank (V-1) which is operated at 30 ºC for 15 h with initial sugar concentration of 20 g/L. The inoculum from V-1 was transferred to gel mixing tank (I-1) to mix the preculture with sodium alginate gel and saline. This mixed gel was then dropped to calcium chloride in beads dropping tank (I-2) to form calcium alginate beads containing immobilized cells. Fed-batch fermentation (R-3) was operated at 30 ºC for 96 h with initial sugar concentration of 100 g/L and concentrated sugar was fed at 24 h, 36 h, 48 h, 60 h, 72 h so that the concentration in the medium was increased to approximately 50 g/L. The ratio of extractant to broth volume (V_e/V_b) was 10 and ratio of cell beads to broth (V_c/V_b) was 3. Filtration (F-2) separated cell beads to be recycled. Liquid phase from filtrate entered the decanter (D-1) which is operated at 30 ºC, 1 atm.

Product rich extractant was fed to the first distillation column (T-1) operated in 1 atm. The temperature of condenser was 75 ºC and reboiler was 324 ºC based on the boiling point of distillate and bottom product, respectively. Extractant could be recovered without loss with this operating condition. The distillate from T-1 containing butanol and acetone mixture was purified in the second distillation column (T-2). This column was operated at 1 atm with condenser at temperature of 56 ºC and reboiler at 117 ºC. This operation resulted in distillate from T-2 contained 99.1% acetone and the bottom product contained 99.5%. Acetone was stored in TT-2 and butanol was stored in TT-3 with atmospheric pressure.



Stream	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35
Phase	s	s	s	l	l	sl	sl	l	sl	sl	l	v	l	l	l	l	l	sl	sl	sl	l	s	l	l	v	sl	l	l	l	l	v	l	l	v	l	l
T (ºC)	30	30	30	30	30	120	55	30	55	55	55	105	105	30	30	30	30	30	30	30	30	30	30	30	30	30	30	30	100	106	75	324	57	56	117
P (atm)	1	1	1	1	1	1	2	1	1	2	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Q (kg/h)	100	100	10	1	8	3	894	894	0.01	845	845	659	619	49	49	9	104	107	900	1807	1000	1007	40	104	8.7	1354	1041	547	94	947	14.4	9.9	937	4.8	3.2	6.7

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	PROCESS FLOW DIAGRAM BUTANOL PRODUCTION AT A LARGE EXTRACTANT VOLUME	01 A-4

Figure 1. Process flow diagram of designed production plant

3.2. Mass Balance of Immobilized Extractive Fermentation

In the block of sugar production, material balances was calculated according to the result from the reference Zhao et al. [10] which was conducted in previous work. Figure 2 shows the material balances of sugar production block. During pretreatment process, inhibitor was removed, most of the hemicellulose was converted pentose, and small amount of cellulose was converted to glucose. The process resulted in product of glucose, xylose, and arabinose. After the mixture was semi-hydrolyzed, most of the cellulose was converted resulted in a sugar mixture containing cellobiose, glucose, xylose, and arabinose [32]. The residue of unhydrolyzed rice straw is recycled to the hydrolysis reactor to be processed further. The filtrate containing clear sugar mixture was concentrated in evaporator. This sugar production block yielded 28.9% w/w sugar per rice straw.

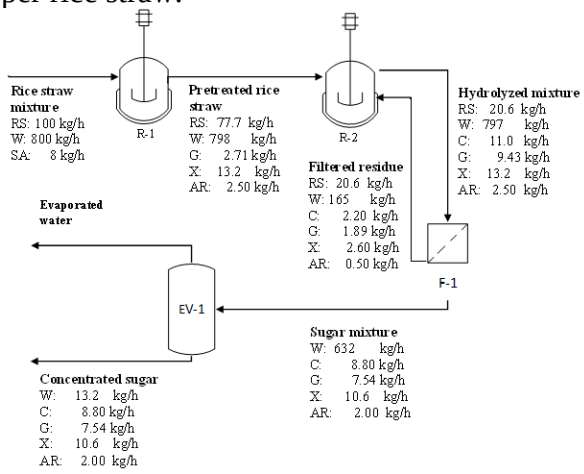


Figure 2. Mass balance of sugar production block in the designed process plant.
*The abbreviation in the figure: R-1, pretreatment reactor; R-2, hydrolysis reactor; F-1, hydrolysate mixture filter; EV-1, sugar solution evaporator; RS, rice straw; W, water; SA, sulphuric acid; G, glucose; X, xylose; AR, arabinose; C, cellobiose.

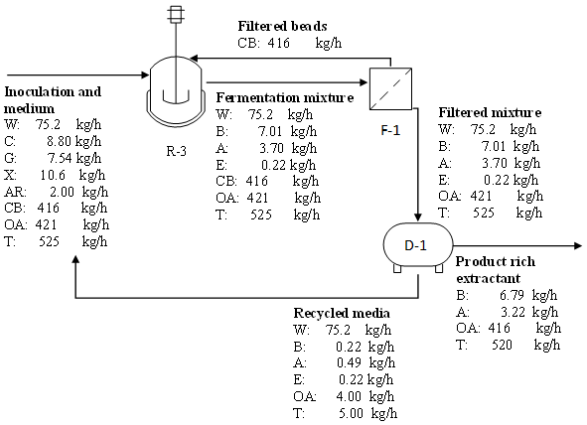


Figure 3. Mass balance of extractive fermentation block in the designed process plant.
*The abbreviation in the figure: R-3, extractive fermentor; F-2, fermentation mixture filter; D-1, decanter; W, water; C, cellobiose; G, glucose; X, xylose; AR, arabinose; CB, cell beads; OA: oleyl alcohol; T: tributyrin; B, butanol; A, acetone; E, ethanol.

Concentrated sugar obtained from sugar production block was fed to the extractive fermentor as shown in Figure 3. The operating time of the fermentor was 96 h. Water containing nutrient was added to the fermentor so that the water flow increased from 13.2 kg/h in concentrated sugar to 75.2 kg/h. The mass of immobilized cell beads was 416 kg/h to fulfil the ratio of Vc/Vb¬ of 3. This fermentor resulted in 7.01 kg/h butanol, 3.70 kg/h acetone, and 0.22 kg/h of ethanol. Both butanol and acetone was extracted by oleyl alcohol and tributyrin mixture because of the high Kd of 3.14 for butanol and 0.66 for acetone. While ethanol was not extracted left in the medium broth. Beads are filtered and to prevent the loss of beads mass, the beads was dipped in calcium chloride before being recycled to the fermentor. The filtered liquid mixture containing product rich extractant and used medium broth was separated in decanter. Upper phase of the decanter was product rich extractant with specific gravity of 0.9 containing 6.79 kg/h butanol and 3.22 kg/h acetone, no water content in the extractant. The bottom phase of decanter was the used media and small amount of extractant to

prevent water mix between phase in the purified extractant. The recycled media contained 75.2 kg/h of water, 0.22 kg/h butanol, 0.49 kg/h acetone, 0.22 kg/h ethanol. The product rich extractant contain only butanol of 6.79 kg/h and acetone of 3.22 kg/h. The process of fermentation block yielded 34.6% w/w butanol and acetone per sugars.

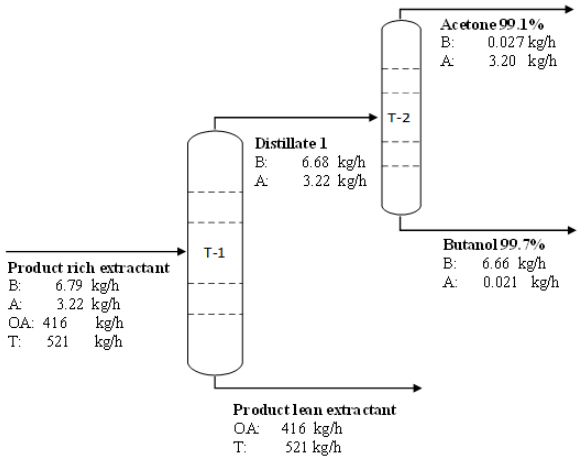


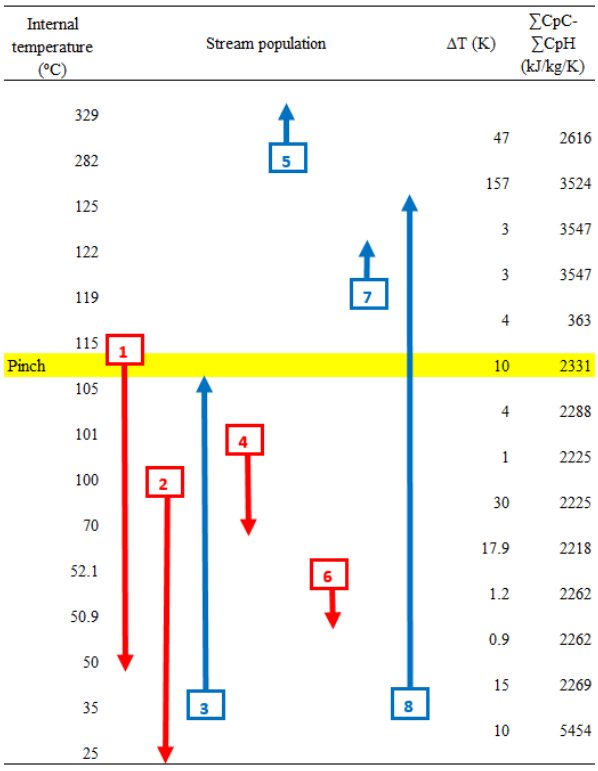
Figure 4. Mass balance of product purification block in the designed process plant.

*The abbreviation in the figure: T-1, extractant recovery distillation; T-2, product purification distillation; OA: oleyl alcohol; T: tributyrin; B, butanol; A, acetone.

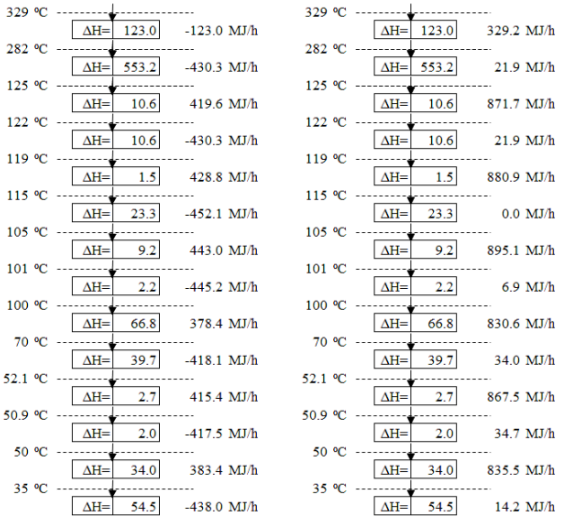
Product rich extractant stream was recovered using distillation (Figure 4). This mixture has no azeotropic properties. Therefore high concentration of the compound can be obtained from each column. Column T-1 is the extractant recovery column, 100% extractant could be recovered in consequence of large difference in boiling temperature of oleyl alcohol and tributyrin mixture of 324 ℃ and butanol-acetone mixture with boiling point of 106 ℃. In order to provide the required composition of distillate and bottom product, both column was optimized using simulation. The optimized reflux ratio is 0.5 for both column. For column T-1, the number of stages was 12 and the feeding stages is number 6 and for column T-2, the number of stages was 22 and the feeding stages was 10. The total produced butanol and acetone was 9.9 kg/h, the calculated material efficiency was 9.9% (w/w) butanol and acetone per rice straw. The previous studies reported that the butanol production using petrochemical process resulted in material efficiency of 92% (w/w) butanol per propene [17]. Another study using corn steep liquor in ABE fermentation with distillation as separation method resulted in 35% material efficiency [11]. The material efficiency of the designed process was lower compared to the reported studies due to the lignin content in rice straw which reduces the convertible rice straw amount to the sugars.

3.3. Energy Balance of The Designed Process

Based on the designed operating conditions and mass balance, energy requirement for the plant is analysed. The highest temperature in the process was 324 ℃ and the lowest was 30 ℃. The cascade diagram in Figure 5 shows that there are deficits in some interval. Energy required to heat the cold stream is much larger than the heat which could be transferred from the available hot stream. According to the heat interval diagram, the minimum heat to be added was 452.1 MJ/h. The heat was added in the cascade diagram to overcome the energy deficit from the heat interval diagram. Based on the cascade diagram, the energy need to be added to the first interval was 329.2 MJ/h and the temperature pinch point was 105 ℃.



(a)



(b)

Figure 5. Stream population (a) and dascade diagram of the designed process (b)

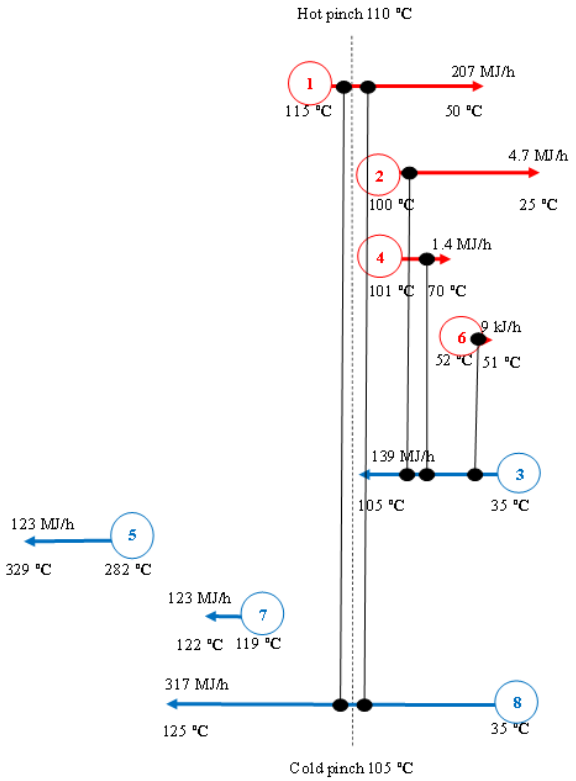


Figure 6. Heat exchanger network
*Symbols: red arrows, hot stream; blue arrows, cold stream.

By considering the temperature difference and required heat, the heat exchanger network was decided as shown in Figure 6. Stream no. 1 exchanged the heat with stream no. 8, remaining heat required for stream 8 will be provided from utility. Stream no. 2 which released energy of 4.7 MJ/h was exchanged with stream no. 3 with the required heat of. The stream no. 4 released the heat of 1.4 MJ/h, also exchanged with stream no. 3. The stream no. 6 was also exchanged with stream no. 3. All the hot streams are exchanged and almost fulfil the requirement, only 1.25 MJ/h needed to cool down the rest of stream no. 2 to 30 °C using utility. The cold stream required 372 MJ/h from the utility to reach the target temperature, especially because of the high temperature required for the reboilers of distillation column.

Table 1. Energy and utility requirement

Cold utility	25	ton/h
Hot utility	372	MJ/h
Electricity (10%)	37.2	MJ/h
Total energy required	409.2	MJ/h
Butanol energy produced	150.4	MJ/h
Energy efficiency	37	%
Specific energy requirement	41.3	MJ/kg acetone - butanol

Table 1 shows the energy requirement for this plant. Cold utility used water in cooling tower with temperature of 25 ℃ heated to maximum temperature of 37 ℃. Hot utility used furnace with rice straw as fuel required 372 MJ/h or the rice straw feeding rate of 16 kg/h. The electricity used in all this plant is 37.2 MJ/h resulted in the total required energy of 409.2 MJ/h. The butanol energy produced from this plant was 150.4 MJ/h and acetone was not calculated because only butanol could be used as fuel. The energy efficiency of produced butanol respect to the energy required for the plant operation was 37%. The specific energy required for the products was calculated by considering both butanol and acetone production flow because both are valuable in the market. It is calculated by dividing the total required energy of this plant with the total mass flows of butanol and acetone products. The specific energy requirement was 41.3 MJ/kg butanol and acetone.

This process consumed relatively lower energy compared with the previous report. Butanol production process using petrochemical feedstock of propene and Oxo synthesis process required 69 MJ/kg butanol [17]. Biobased production using ABE fermentation with corn steep liquor [11] required high energy input of 116 MJ/kg butanol. The use of simultaneous extraction-fermentation which resulted in no azeotropic mixture and the atmospheric pressure operation in almost all unit has successfully reduced the energy requirement to the lowest compared with the existing reports.

3.4. Cost Estimation of Immobilized Extractive Fermentation

Total cost was estimated by summarizing investment and operational for the plant. In order to determine the investment cost, the cost of the equipments was determined. The result of cost estimation is shown in Table 2. The cost of each equipment was estimated using Table 6-4 and 6-5 [31].

Table 2. Estimation of investment cost

Equipment	Unit	Cost per unit ($)	Total Cost ($)
Pretreatment reactor	1	10,000.00	10,000.00
Hydrolysis	1	49,710.00	49,710.00
Fermentor	2	42,435.00	84,870.00
Distillation 1	1	102,000.00	102,000.00
Distillation 2	1	70,000.00	70,000.00
Hammer mill	1	350.00	350.00
Belt converyor	2	10,700.00	21,400.00
Evaporator	1	15,300.00	15,300.00
Filter 1	2	21,700.00	43,400.00
Filter 2	2	6,000.00	12,000.00
Filter 3	2	6,000.00	12,000.00
Preculture tank	1	30,000.00	30,000.00
Gel mixing tank	1	39,906.00	39,906.00
Beads dropping tank	2	39,906.00	79,812.00
Decanter	2	30,300.00	60,600.00
Rice straw storage	1	7,700.00	7,700.00
Pretreatment holding tank	1	16,300.00	16,300.00
Acetone tank	1	8,160.00	8,160.00
Butanol tank	1	16,300.00	16,300.00
Heat exchanger			129,000.00
Pump			9,485.00
Cooling tower			3,500.00
Sodium alginate	7200 kg	5.00	36,000.00
Oleyl alcohol	40000 kg	5.00	200,000.00
Tributyrin	50000 kg	10.00	500,000.00
	Total		1,557,798.00

Total investment for equipment was $1,424,995.00 The operational cost was listed in Table 3 in a year basis, total operational cost required was $87,717.00. Materials cost was estimated by the price in the credible global e-commerce.

Table 3. Estimation of the operating cost

Materials	Consumption (kg/year)	Unit cost ($)	Total cost ($/year)
Rice straw	969,000	0.02	14,535.00
Tryptone	5,220	11.00	57,420.00
Ammonium acetate	2,610	1.00	2,610.00
Yeast extract	1,740	7.00	12,180.00
Calcium chloride	10,800	0.09	972.00
Total			87,717.00

4. CONCLUSION

From this study, a small scale production with capacity of 238 kg-butanol and acetone/day from 2.4 ton-rice straw/day was designed using semi-hydrolysis and extractive fermentation with immobilized cells at a large extractant volume. Low specific energy requirement of 41.3 MJ/kg was achieved. Butanol-acetone production cost in the designed process was 1.91 $/kg. In order to improve the economical feasibility, further optimation in fermentation process is needed, especially to reduce the requirement of nutrient which consumed the highest cost. It is suggested to increase butanol concentration in extractant and conduct study concerning recyclability of extractant and immobilized beads.

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