Process Plant Design and Feasibility of an Acetone-Butanol-Ethanol (ABE) Fermentation Biorefinery using Sugarcane Bagasse and *Clostridium acetobutylicum*

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Abstract—The acetone-butanol-ethanol (ABE) fermentation process is a promising technology which utilizes renewable resources such as sugarcane bagasse into a number of products with high market value. In the proposed plant design, sugarcane bagasse will be used as the raw material in the ABE fermentation with *Clostridium acetobutylicum*. For the analysis, it is assumed that sugarcane bagasse will be purchased and transported from established sugar mills. Using dilute acid hydrolysis, celluloses and hemicelluloses present in bagasse will be converted to glucose. The extracted glucose will proceed to the fermentation step in order to be converted to ABE products. Corresponding purification steps follow, based on the required purity grades of the products. The feasibility study aims to produce 157,853 tons of products per annum, catering a variety of industries such as the chemical, paint, polymers, plastic, food, and pulp and paper industries, among many others. The proposed design achieved a market value of USD 528 million from 350,460 tons of bagasse per annum and 28% internal rate of return.

Index Terms—ABE fermentation, bagasse, biorefinery, process design.

I. INTRODUCTION

There is an ever-increasing demand for energy, and currently, the world is highly dependent on exploiting non-renewable resources. The limited supply of fossil fuels and volatile prices of crude oil call for more sustainable energy sources, one of which is biomass. Generation of fuels from biomass-based plants has garnered much attention, including ethanol produced from corn and sugarcane. However, competition between food supply and raw material supply for fuel could pose several problems. Sugarcane bagasse, one of the wastes in the sugar industry, contains enough sugars to provide for a fully functioning biomass-based plant.

Acetone-Butanol-Ethanol process or ABE process utilizes biomass as a raw material for the production of alcohols and solvents. The fermentation of glucose is carried out in an anaerobic bioreactor using bacteria. Among a number of microbial strains producing butanol, *C. acetobutylicum* is the most extensively used species for industrial fermentation. Other strains include *C. beijerinckii*, *C. aurantibutyricum*, and *C. tetanomorphum*. Studies on the aforementioned

microorganisms show higher yields of the products because of their higher resistances to substrate and product inhibitions as compared to *C. acetobutylicum*. Unfortunately, there were only limited taxonomic studies undertaken, all of which are of laboratory-scale, and there are no available approved standards for the classification of these microorganisms yet. Hence, the selection of strains for industrial-scale fermentation is based on the nature of raw material utilized, the ratio of end-products required, the need for supplementary nutrients, and data availability on product yield for an industrial-scale process [1]. Since 1912, *C. acetobutylicum* has already been studied and data are well-documented for fermentation. Kinetic models and mechanisms have been established for the production of butanol.

II. METHODS

A. Process Description

The proposed design for the batch ABE fermentation process can be divided into three main parts: purification of feed, fermentation process and recovery of products. Acetone, butanol, ethanol, carbon dioxide and hydrogen are produced by the fermentation process of *C. acetobutylicum* using glucose as the primary substrate.

B. Economic Analyses

Equipment sizing was performed by manual calculations, engineering heuristics [2] for process equipment and simulation in Hysys™. Equipment costs are calculated based on heuristics given [3]. For uncommon equipment, costs were estimated from papers with market considerations such as that of pressure swing adsorption for hydrogen purification [4] normalized in the year 2016 by accounting the chemical engineering plant cost index (CEPCI 2004: 444.2 and CEPCI 2016: 624.52 by extrapolation) [5]. Furthermore, fermenter costs are estimated using capacity-dependent costing in SuperPro™.

The total capital investment (TCI) was calculated based on established methods [2]. The percentage values of TCI were adopted from an ABE batch plant in 2000 while the rest of the direct and indirect field costs are estimated to be of the following percentages shown in Table I, calculated based from previous economic analysis of ABE fermentation in 1980 [6].

Finally, the sensitivity of the profitability of the process plant with respect to fixed capital investment (FCI), sales and cost of goods sold (COGS) are investigated.

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TABLE I: BREAKDOWN OF INSIDE BATTERY LIMITS INVESTMENT

	% of FCI
<i>Direct Cost</i>	
Piping	7
Instruments	5
Electrical	4
Civil	4
Buildings	3
Lagging	2
Services, yard	16
<i>Indirect Cost</i>	
Engineering and supervision	15
Construction expenses	15

III. RESULTS

A. Process Synthesis

Process flow diagrams (PFD) for the proposed process were developed and are shown in Fig. 1 to 8. The pretreatment process was divided into sections 100, 200 and 300. The fermentation process was divided into sections 400 and 500. The purification process was divided into sections 600, 700 and 800.

B. Pretreatment

Pretreatment begins with the hemicellulose hydrolysis section in Fig. 1. Sugarcane bagasse enters the ball mill at a rate of 44,250 kg/h to be homogenized at a size of 0.212 to 1.18 mm. The bagasse is then washed and sent to a heating

vessel together with 200,340 kg/h of 2 wt% sulfuric acid solution. The mixture is then pumped to 300 kPa and sent to the reactor vessel which acts like a flash. The hemicellulose is converted to sugars, while the cellulose and lignin remain as solids. The liquid mixture is sent to a filter, where the lignocellulose is sent to the next section and the pentose liquor is sent to a third-party wastewater treatment.

The delignification section in Fig. 2 utilizes the Organosolv technology. Organosolv is a solvent of 60 wt% ethanol and 40 wt% water. The solvent dissolves the lignin but keeps the cellulose as solids. Lignocellulose is pumped to 1900 kPa before it is sent to a reactor with organosolv heated to 180°C. The mixture proceeds to a flash drum at 850 kPa. Steam is vented out and the liquid mixture is filtered. The solids and a few amount of water proceed to cellulose hydrolysis, and the spent organosolv with the dissolved lignin is sent to the solvent recovery system.

From the delignification section, the cellulose mixture at 35,553 kg/h enters a mixer with 209,497 kg/h of water and 8,888 kg/h of 2 wt% sulfuric acid shown in in Fig. 3. The mixture is pumped to 3600 kPa, heated to 205°C and flashed. The steam is vented out and the hexose liquor is cooled from 205°C to 90°C by the organosolv and dissolved lignin stream from the previous section. The hexose liquor is first filtered to remove any unreacted solids and remaining ashes. The solids are sent to a third-party waste treatment. The liquid proceeds to a three-effect evaporator. The glucose-rich liquor then proceeds to a storage tank before it is used in the fermentation section.

Fig. 1. The first pretreatment process is Unit 100, Cleaning and Hemi-cellulose Hydrolysis, where sugarcane bagasse is converted to cellulignin.

Fig. 2. The second pretreatment process happens in Unit 200, delignification. Cellulignin is pressurised with Organosolv, sent to a reactor where lignin is separated from cellulose. This is followed by filtration to separate cellulose from dissolved lignin in Organosolv.

Fig. 3. In Unit 300 cellulose is hydrolyzed into glucose and concentrated in a triple effect evaporator to form the glucose rich liquor. To save energy the Organosolv Recovery utilizes heat from hydrolysis to preheat the Organosolv-Lignin stream.

Fig. 4. In Unit 400, Pre-Fermentation, glucose is combined with the constituents of the fermentation broth which is then pasteurized.

The spent organosolv stream is preheated from 90°C to 180°C and pumped to 480 kPa before entering the distillation column to recover all the ethanol and the dissolved lignin goes to the bottoms with water. The lignin-water mixture is sent to a flash drum to precipitate the lignin. The mixture is then cooled to 90°C before it is filtered. The solid lignin is then dried and sent to a storage tank at a rate of 9,464 kg/h. The recovered ethanol is further cooled from 125°C to 90°C. The water from the filtered lignin, and the recovered ethanol are mixed together in a vessel before it is recycled back to the delignification section, creating a closed-loop system of an Organosolv recovery system.

C. Fermentation

The fermentation part is divided into the pre-fermentation and the fermentation sections shown in Fig. 4 and 5, respectively. Glucose solution from cellulose hydrolysis section is stored in a holding tank with a capacity of 85,600 L. Glucose solution is then divided into six streams which are

further divided into two streams to enter a total of 12 mixing vessels at a mass flowrate of 5,917.81 kg/h. These mixing vessels will not operate simultaneously since the whole fermentation part of the plant operates in batch. The 12 mixing vessels will be separated into three units, one for each fermentation unit.

Another component entering the mixer is the broth in which the bacteria culture will thrive. The broth with composition described by Yang and Tsao [7] as listed in Table II, will enter the mixing vessel at 18,955.07 kg/h.

TABLE II: COMPOSITION OF THE BROTH MEDIUM

Component	Flowrate (kg/h)	Component	Flowrate (kg/h)
(NH ₄) ₂ SO ₄	49.32	MnSO ₄ ·H ₂ O	0.25
KH ₂ PO ₄	18.49	L-(+)-cysteine	12.33
K ₂ HPO ₄ ·3H ₂ O	18.49	Yeast Extract	123.29
FeSO ₄ ·7H ₂ O	0.25	Asparagine	49.32
MgSO ₄ ·7H ₂ O	4.93	Water	18,653.76
NaCl	24.66

Fig. 5. Unit 500 is the Fermentation process where pasteurized fermentation broth is converted into Acetone, Butanol, Ethanol, Hydrogen and Carbon dioxide by a *Clostridium acetobutylicum* culture.

Fig. 6. ABE are extracted from the broth using liquid-liquid-extraction in Unit 600.

The feed solutions of glucose, water, and broth from four mixing vessels are combined and pasteurized before entering its corresponding fermentation unit. The fermentation section consists of three fermentation units operating in batch. Each fermentation unit is 8 hours out of phase with one another to incorporate a pseudo-continuous process. Each unit has two fermenters with a capacity of $1,350 \text{ m}^3$. The entering bacteria at the first unit is at a rate of 105.89 kg/h . Fermentation will run for 24 hours, then it is drained, washed, stripped of oxygen, and re-filled with the feed.

The bottom stream consists of liquids – water, acetone, butanol, ethanol, and a few unreacted glucose, and broth – and the solid cell mass. This mixture proceeds to a centrifuge, separating the cells from the liquid. The bacteria cells proceed to a holding tank before it is reused in the next fermentation unit. This process is repeated until the bacteria are used three times. The spent bacteria then proceed to a third-party wastewater treatment facility. The liquid products are then sent to the ABE holding tank at a rate of $48,089 \text{ kg/h}$ per fermenter. Gas products are vented out of each fermenter at $1,363.43 \text{ kg/h}$. The non-condensable product gases – hydrogen and carbon dioxide – proceeds to the gas purification section.

D. Purification

The purification part is divided into the ABE extraction

section, the distillation section, and the gas purification section shown in Fig. 6, 7 and 8, respectively. The ABE-water stream at $96,177 \text{ kg/h}$ is pumped to 150 kPa to the top of the liquid-liquid extraction unit and the solvent, oleyl alcohol, enters at the bottom at the same pressure. The raffinate at the bottom contains mostly water and minute concentrations of ABE. The raffinate is then sent to a third-party wastewater treatment. All of the solvent proceed to the extract phase with ABE. The extract phase has a flowrate of $130,375 \text{ kg/h}$. It is then pumped to 291.325 kPa and preheated to 320°C before entering the distillation column. All of the solvent is recovered at the bottom and is recycled back to the extraction unit.

The ABE-rich stream at 2325.75 kg/h proceeds to the distillation section. Acetone still produces 857.80 kg/h of 99.75% pure acetone in the distillate. Butanol is recovered in the second distillation, which produces 988.02 kg/h of 99% pure butanol at the bottoms. Finally, ethanol is recovered in the last column at 444.51 kg/h as 98% ethanol in the distillate stream. The bottoms stream of the last column proceeds to wastewater treatment.

The gas products are separated using two PSA units: one for hydrogen and one for carbon dioxide. The hydrogen PSA unit produces 99.99% pure hydrogen at a rate of 326.30 kg/h , and the carbon dioxide PSA unit produces 99.99% pure carbon dioxide at a rate of 7854.3128 kg/h . The inert gas

nitrogen is recycled for blanketing.

Fig. 7. ABE are separated in Unit 700 to product specifications by a sequence of distillation units.

Fig. 8. In Unit 800, Gas Purification is achieved by a sequence of pressure swing adsorption systems.

E. Economics

The calculation of total capital investment is summarized in Table III. In the calculation of the cash flow, startup expenses are assumed to be lumped with the working capital. The total capital investment of the preliminary design of ABE fermentation is USD 389M.

TABLE III: CALCULATION OF TOTAL CAPITAL INVESTMENT

	Total Cost (USD)
Inside Battery Limits (ISBL) Investment	174,266,000
Outside Battery Limits (OSBL) Investment	69,707,000
Contingency Charges	24,397,300
Startup	80,511,090
Working Capital	40,255,545
Fixed Capital Investment	268,370,300
Total Capital Investment	389,136,935

Operating costs include variable costs of production – sales revenue, cost of raw materials, utilities, consumables and effluent disposal – and fixed costs of production –

operating labor, supervision, etc. Table IV lists all calculations for the operating costs of 330 days or 7,920 hours of operation.

TABLE IV: REVENUE AND PRODUCTION COSTS

	Total Cost (USD)
Products (Acetone, Butanol, Ethanol, Lignin, CO ₂ , H ₂)	Annual Revenue: 528,163,758
Raw Materials	136,737,428
Utilities, Consumables, Effluent Disposal	Annual Cost: 102,827,335
Fixed Costs of Production	15,133,964

During the investing phase, four different scenarios were taken into account for the construction period of the plant. The scenario with the least expenditure is one with one-time loan from the bank for the basis of FCI. A two-year construction period was taken into account with the first year using all of the equity and the rest is loaned from the bank (40% and then 60% for years 1 and 2, respectively) resulted to an interest charge of USD 9,680,000. This scenario was

chosen as the basis for cash flow of construction costs for this design. It is noted that for all the cases aforementioned, the interest rate used is 6%.

Cash flow is calculated with a salvage value of 15% of FCI, and 10-year depreciable life. Depreciation was calculated using straight line depreciation method. As a result, an IRR of 28% was achieved by the proposed plant operations with a total accumulated net present value of USD 707M. A straight payback of 5.44 years and discounted payback of 5.91 years is also achieved in this preliminary design.

The net cash flow diagram shown in Fig. 9 indicates an expected positive flow of cash by the start of sales. On the first year of sales, it was based upon that only 30% of the sales are achieved in the operating year while it is 90% in the second year. The rest of the years will then have 100% sales per year. At the end of 10 years, depreciation, working capital and land costs will be accounted back as forms of cash flow input.

Fig. 9. The cash flow diagram shows a positive cash flow at year 3 during the first year of sales.

In the sensitivity analysis, the cost of goods sold (COGS) is more variable and dependent on the changes in the percentage sales. From Fig. 10, the slope of FCI vs discounted payback is 2.2 while that of COGS is 5.5. Thus, COGS tend to change more dramatically with respect to percentage sales. That is because the FCI is primarily dependent on heuristics and cost of equipment which are nearly fixed as compared to COGS that is highly dependent on the amount produced due to higher values of variable costs.

Fig. 10. Sensitivity plot shows that the discounted payback is most sensitive to the cost of goods sold.

The breakeven point obtained is 29.18%, shown in Fig 11.

This explains the positive cash flow at the start of its operation, having 30% sales procurable.

Fig. 11. The breakeven even point occurs at 30% of sales.

A comparison of the results in this study is shown in Table V. Although this study used half the butanol yield and a lower sugar feed rate, the IRR is better than the result of the simulation by Mariano *et al.* Though the IRR of this study is inferior to that of Mansur *et al.*, the study period differs greatly. The Net Present Value of this study at USD 707M is much higher than the USD 118.8M from Mansur *et al.* The better profitability in this study may be attributed to the utilization of low cost sugarcane bagasse as the process feed.

TABLE V: COMPARISON WITH OTHER STUDIES

	Mansur <i>et al.</i> [8]	Mariano <i>et al.</i> [9]	This study
Bacteria	Novel <i>Clostridium</i> strain	<i>Clostridium beijerinckii</i> BA101	<i>Clostridium acetobutylicum</i>
Yield (g Butanol/g sugar)	0.2236	0.34	0.17
Feed	Sugarcane juice	Sugarcane juice	Sugarcane bagasse
Sugar feed rate (tons/yr)	215,040	2,000,000	140,607
IRR	37.03%	14.80%	28%
Net present value	USD 118.8 M	-	USD 707M
Study period	20 years	25 years	10 years

IV. CONCLUSION

This feasibility report showed that a sugarcane bagasse based ABE biorefinery in the Philippines may very well be realized. With an IRR of 28% and a payback of 5.44 years, the detailed design of this biorefinery should be further explored.

For the optimization of process streams inside the plant, there are several opportunities to save on utilities. The various streams containing steam from several reactors and flash drums could be used to heat some of the process fluids, instead of relying on heating coils. The distillate from the multiple-effect evaporator may be used as the washing solvent for the bagasse and the make-up water at the beginning of the cellulose hydrolysis section. Another use for the steam is for the generation of electricity to cut back on energy-intensive processes, including the ball mill, mixers coupled with heaters, and the centrifuges. A fraction of the lignin may also be utilized as fuel for the boilers.

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