Process design, simulation, and techno-economic analysis of integrated production of furfural and glucose derived from palm oil empty fruit bunches

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Abstract

This study aims to propose a new process design, simulation, and techno-economic analysis of an integrated process plant that produces glucose and furfural from palm oil empty fruit bunches (EFB). In this work, an Aspen Plus-based simulation has been established to develop a process flow diagram of co-production of glucose and furfural along with the mass and energy balances. The plant's economics are analyzed by calculating the fixed capital income (FCI), operating costs, and working capital. In contrast, profitability is determined using cumulative cash flow (CCF), net present value (NPV), and internal rate of return (IRR). The findings show that the production capacity of 10 kilotons per year (ktpy) of glucose and 4.96 ktpy of furfural with a purity of 98.21 and 99.54% - weight, respectively, was achieved in this study. The FCI is calculated as United States Dollar (USD) 20.80 million, while the working and operating expenses are calculated as USD 3.74 million and USD 16.93 million, respectively. This project achieves USD 7.65 million NPV with a positive IRR of 14.25% and a return on investment (ROI) of 22.06%. The present work successfully develops a profitable integrated process plant that is established with future upscaling parameters and key cost drivers. The findings provided in this work offer a platform and motivation for future research on integrated plants in the food, environment, and energy nexus with the co-location principle.

Keywords: process design, palm oil empty fruit bunches, simulation, techno-economic, glucose, furfural

Nomenclature and units

No.	Symbols	Abbreviation
1	MSW	Municipal Solid Waste
2	MT	Metric Tons
3	EFB	Empty Fruit Bunch
4	MEBT	Mass and Energy Balance Table
5	GWP	Global Warming Potential
6	DE	Dextrose Equivalent
7	CARG	Compound Annual Growth Rate
8	ktpy	Kilotons per year
9	tpy	Tons per year
10	CIP	Cleaning in Place
11	HACCP	Hazard Analysis Critical Control Points
12	H_2O	Water
13	PFD	Process Flow Diagram
14	wt.%	Weight Percentage
15	ppm	Parts Per Million
16	°C	Degree Celsius
17	FCI	Fixed Capital Investment
18	NPV	Net Present Value
19	IRR	Internal Rate of Return
20	ROI	Return on Investment
21	CCF	Cumulative Cash Flow
22	VLE	Vapor Liquid Equilibrium
23	LLE	Liquid-Liquid Extraction
24	CPSPO1-1	Solid Molar Volume
25	VSPOLY-1	Solid Heat Capacity
26	GDP	Gross Domestic Product

1. Introduction

The usage of natural resources has spiked up rapidly due to the tremendous increase in the global population. The vast number of resources consumed generally results in more waste. It is anticipated that by 2050, worldwide municipal solid waste (MSW) will be over 3.40 billion metric tonnes (MT) **(Ellis 2018)**. Food waste dominated the highest percentage, followed by paper and paperboard, plastic, yard trimmings, and metals **(Aydin 2015; Aydin et al. 2017; Celep et al. 2013; The World Bank 2021)**. These wastes are usually disposed of in landfills, recycled, composted, or utilized in energy recovery.

Malaysia, Indonesia, and Thailand are the leading producers of palm oil, accounting for more than 90.00% of the global market in total, with Malaysia accounting for 25.90% of global production and 33.70% of global palm oil exports in 2019 **(MPCO 2021; Zafar 2021)**. In a palm oil mill, fresh fruit bunch is employed as a raw material; however, only 23% of the raw material is utilized to make palm oil, with the remainder being trash **(Aziz et al. 2015)**. There is an abundance of waste being produced from palm oil plantations or mills, such as palm oil empty fruit bunch (EFB), fibers, and nutshells, which require processing into value-added products or undergo waste minimization processes **(Aziz et al. 2015)**. In certain scenarios, open-ended burning is used to dispose of EFB trash, causing huge environmental difficulties, including global warming potential (GWP) and air pollution **(Ninduangdee & Kuprianov 2016)**. These EFB can also be reused as biomass, reducing the processing plant's economic and environmental burdens. Since EFB biomass is low in sulphur, it reduces GWP and air pollution **(Ninduangdee & Kuprianov 2016).** Furthermore, EFB can be processed into high-value products such as glucose, xylitol, levulinic acid, and vanillin **(Hafyan et al. 2020; Umana et al. 2020)**. In addition, it can be chemically processed into ethanol, furfural, and lactic acid and converted into energy such as bio-oil, biogas, bioethanol, and methane **(Geng 2013)**.

Malaysia, Indonesia, and Thailand generated 27 million MT of EFB in 2019 **(Zafar 2021)**. Since EFB is readily available in Malaysia, it is an inexpensive renewable energy source **(Alaw & Sulaiman 2019)**. These EFB can be further processed to extract valuable food and chemicals; hence, promoting resource conservation and sustainable processes **(Alaw & Sulaiman 2019)**. However, there is still much waste to date, even if many plants use the EFB to manufacture foods, chemicals, or energy **(Chiew & Shimada 2013)**. This is because the existing process plant only valorizes a particular component in EFB such as cellulose or hemicellulose only. This procedure also requires many separations, increasing the capital cost. To overcome this, lignocellulosic biomass can be processed into numerous products in a single plant in an integrated way **(Jin et al. 2018; Orugba et al. 2021)**. For an integrated process plant producing both food and chemicals, glucose and furfural are selected as end products because they can be produced in a single facility with minimal equipment **(Buaisha et**

al. 2020; Loginova et al. 2021). Moreover, the residual lignin can be used as a fuel for boilers to supply heat energy. Hence, the plant's economic and environmental impacts are reduced as a result of maximizing EFB's potential.

Currently, only a few studies have worked on an integrated process plant employing hemicellulose in biomass. **Loginova et al. (2021)** developed an integrated system to convert different lignocellulosic biomass (i.e., wheat straw, corn cobs, sugar beet pulp, birch sawdust, and oat husk) to furfural and glucose. In addition, an integrated biorefinery process for co-production of glucose and xylose (i.e., raw material for producing furfural) from wheat straw has been developed and analyzed by **Liu et al. (2021)**. Moreover, **Choi et al. (2019)** proposed a simultaneous synthesis process of glucose, ethanol organosolv lignin, and furfural from lignocellulosic biomass. Although their system can produce glucose and xylose or furfural, their study only focused on experimental work and mass balance calculation without sufficient techno-economic analysis.

In addition, those previous studies are not solely dedicated to converting EFB to glucose and furfural. Also, the process conditions were not optimized to maximize the production rate and purity. Existing research also lacks modelling data and production rates for integrated processes. The process design and optimization can also assist in achieving the requisite product purity. Moreover, existing work lacks economic analysis and profitability to assess an integrated plant's viability. Therefore, this work aims to use simulation to obtain the mass and energy balance for optimizing the integrated process and perform an economic analysis to identify the overall profitability and plant viability.

In an integrated production process, maximizing feedstock utilization by producing multiple outputs is crucial in order to increase the efficiency of feedstock consumption. Recent research has revealed that if two industries were co-located, there would be opportunities for major resource sharing, notably in the area of energy **(Ahmad et al. 2019; Solomou et al. 2022)**. In this regard, **Sheppard et al. (2019)** also emphasize the mutual benefits that could result from the co-location of biorefineries and food and beverage manufacturing facilities, such as enhancing the material and energy flows between the two facilities, reducing manufacturing costs, as well as integrating heat and water consumption more effectively than a single facility. Hence, integrated production of furfural and glucose has the potential to demonstrate substantial economic advantages and returns.

The significance of this study is first to address the techno-economic feasibility of an integrated plant that can co-produce food and chemicals using EFB biomass. In this regard, the food safety aspect is indeed important. In the latter stage, cleaning in place (CIP) and hazard analysis critical control points (HACCP) practices that can sterilize the process facilities will be implemented for glucose production to ensure the process is in compliance with food production, safety and hygiene standards **(Banach et al. 2020; Chaib & Barone 2020; Malliaroudaki et al. 2022; Motarjemi &**

Lelieveld 2013). By practicing the CIP and HACCP principles, the proposed model of glucose and furfural co-production can concurrently optimize the planning, design, and operation of value chains for EFB biomass with integrated food and chemical production.

Converting EFB components into value-added products reduces waste generation and promotes a circular economy. Using process simulation of Aspen Plus (Aspen Technology Inc.), the production rate, purity, and flowrate of the product can be calculated and adjusted to reduce raw material consumption and eventually reduce the capital cost. When evaluating the profitability of the integrated plant, it helps to compare the main cost drivers of the integrated plant. Furthermore, sensitivity analyses optimize production parameters such as capacity and fixed capital. Overall, the plant viability can be determined by the process sustainability and profitability.

2. Process description

Fig. 1 shows the research flow chart of this study. First of all, the conceptual process design based on the characteristics of the raw material and the requirements of the product (i.e., purity) has been proposed. The output of the process design leads to the generation of a block flow diagram and process flow diagram (PFD). To determine the viability of a proposed plant, unit operations were arranged in series to form a flowsheet that can be simulated in Aspen Plus. Some parameters are manipulated in techno-sensitivity analyses to obtain the optimum operating conditions. The mass and energy balances generated from the Aspen Plus simulation become the basis for analyzing the plant's capital and operating costs before conducting economic sensitivity analyses. Following this, a comprehensive techno-economic analysis determines the plant's viability. In addition, the findings are analyzed by comparing them with other literature reviews that produce furfural and glucose and with other existing commercial plants.

2.1 Raw material and products

2.1.1 Palm oil empty fruit bunch

EFB is lignocellulosic biomass composed of cellulose, hemicellulose, and lignin where it can be broken down into simple sugars to produce various food additives **(Geng 2013)**. On the other hand, this biomass can also be converted to renewable energy such as syngas, ethanol, furfural, and bio-oil through thermochemical or biological conversion **(Geng 2013)**.

Table 1. *Composition of EFB (wt.%) (Reproduced from the work of* ***(Chiesa & Gnansounou 2014)****).*

2.1.2 Furfural

Furfural is an oily, colorless liquid that turns dark brown when exposed to air **(Ebert 2008).** It is used to refine lubricating oil, as a fungicide in tetrahydrofuran manufacturing, and as an industrial solvent **(Ebert 2008)**. Moreover, it is an important raw material for the production of furfuryl alcohol by catalytic hydrogenation and is also used for other transportation fuels like ethyl levulinate and dimethylfuran **(Ebert 2008; Mordor Intelligence 2021)**. Furfural is produced at high purity with low moisture content by dehydration of 5-carbon sugar xylose and arabinose **(Ebert 2008).** Table 2 provides its specifications.

Table 2. *Properties and composition of furfural (Reproduced from the work of* ***(Hongye Holding Group Corporation Ltd. 2020)****).*

In 2018, worldwide furfural exports and imports increased to 146.70 and 155.70 kilotons per year (ktpy), respectively **(UN Data 2018)**. In Malaysia, the import demand for furfural reached 267.80 tpy in 2018 **(UN Data 2018)**, and this value is expected to with a compound annual growth rate (CARG) of 5% by 2026 **(Mordor Intelligence 2021)**. Among the global consumers, China is the largest consumer globally, followed by western Europe and other Asian countries **(IHS Markit 2020)**.

2.1.3 Glucose

Glucose is a viscous, colorless, and sweet fluid that improves the food texture, taste, and gloss **(Wilson & Lilly 1969)**. Therefore, it is often used as a sweetener in the production of food and chemicals industries **(Basso & Serban 2020)**. Glucose can also be converted into bioethanol through fermentation with yeast, producing bioenergy, or into levulinic acid, succinic acid, lactic acid, etc. **(Hafyan et al. 2020)**. Industrial glucose production comes in various compositions and purity, which can be benchmarked by their dextrose equivalent (DE) value, which measures the percentage of inverted sugar, including glucose, oligomer, dextrose, and maltose **(Sarungalo 2005)**. The sweetness of glucose syrup is determined by the feedstock's DE content, including low DE (26-29%), high DE (40-45%), and the sweetest DE (56-64%) **(Sarungalo 2005)**. For EFB, the DE falls between 42 and 45%, and the composition is shown in Table 3 **(21Food 2021)**.

Table 3. *Properties and composition of glucose (Reproduced from the work of* ***(21Food 2021)****).*

In 2018, worldwide glucose exports and imports increased to 2995.50 and 3773.30 ktpy, respectively **(UN Data 2018)**. In Malaysia, the import demand for glucose reached 44.46 tpy in 2018 **(UN Data 2018),** and this value is expected to double by 2030, with a compound annual growth rate (CARG) of 5.20% **(Persistence Market Research 2020)**. Belgium is the worlds' largest consumer, whereas Malaysia stands in 13th place **(UN Data 2018)**.

2.2 Process flowsheet of the integrated process

The glucose and furfural are produced *via* enzymatic hydrolysis and dehydration, respectively. The process includes breaking down the EFB to separate cellulose and hemicellulose in feed preparation. Furthermore, it is followed by enzymatic hydrolysis for glucose production and dehydration for furfural synthesis along with purification and recovery of the products to achieve the desired specification. Fig. 2 shows the associated process description of each unit.

Fig. 2. *Integrated production of glucose and furfural from palm oil empty fruit bunch.*

2.2.1 Feed Preparation

The EFB is purchased from the palm oil processing plant and stored in a warehouse before being conveyed to the feed preparation. Initially, the EFB contains 19.8 wt.% moisture, which is reduced to 10 wt.% using 105 °C hot air in the rotary dryer (S-1001) **(Loh 2018)**. Then, the dried EFB is transported to a grinder (S-1002) to be cut into small pieces with a size between 0.10 to 1.00 mm **(Kenthorai Raman & Gnansounou 2015)**. Furthermore, it is transported to a vibrating screen (S-1003), which contains an 80–20 wired mesh to separate the smaller particles sized below 1 mm. In contrast, the larger particles are recycled back to the grinder (S-1002) to obtain uniformly sized EFB **(Kenthorai Raman & Gnansounou 2015)**.

2.2.2 Pretreatment

In the pretreatment section, process water and 70 wt.% dilute sulfuric acid are pumped from its storage tank to an agitated mixing tank (M-1101) operating at 30 °C and 1 atm. Simultaneously, the recycled water from the furfural purification unit (C-1302) is pumped to the mixing tank (M-1101). Here, the sulfuric acid is diluted using the process and recycled water and then mixed with the ground EFB from the feed preparation unit (CONV-1101) in an agitated mixing tank (M-1102). The flowrate of sulfuric acid and process water is adjusted to 30% of solid loading with 18 mg of sulfuric acid per g of dry EFB **(Humbird et al. 2011)**. The mixture is then pumped (P-1104) to 6 atm and heated (E-1101) to 158 °C before being fed into an insulated continuous stirred tank reactor (CSTR) (R-1101) and stirred for 5 min **(Humbird et al. 2011)**. The reactor converts most hemicellulose to xylose, with secondary reactions as shown in Table 4.

Table 4. *Reaction and its conversion in CSTR (R-1101).*

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The product from CSTR (R-1101) is cooled (E-1102) to 50 °C for solid separation. The cooled product is sent to a decanter centrifuge (S-1101) to separate solid and liquid. The decanter centrifuge can separate particles sized from 0.10 to 1.00 mm using its centrifugal force rotating horizontally, separating the solids on the separator wall, and removing them using a screw conveyor **(Dolphin Centrifuge 2021)**. The solid is completely removed with 5 wt.% moisture and sent to the neutralization unit, while the liquid stream is sent to the dehydration unit.

2.2.3 Furfural synthesis

The liquid stream from the pretreatment is dehydrated to convert xylose into furfural, where the reaction is autocatalyzed by heat. Firstly, the liquid is heated to 170 °C and pumped (P-1201) to 8.80 atm into an insulated CSTR (R-1201) **(Mittal et al. 2017)**. The residence time of the reaction is 20 min. As our processing plant is continuous, the CSTR (R-1201) has a 3.62 m^3 capacity, accounting for the residence time. The reactors were assumed to have a maximum operating capacity of 80% of the designed volume **(Mittal et al. 2017)**. In CSTR (R-1201), the conversion of xylose into furfural can be achieved up to 96.50% **(Mittal et al. 2017)**.

2.2.4 Furfural recovery

The furfural mixture from the dehydration unit requires further purification because it contains water, sulfuric acid, and acetic acid, which must be removed to obtain high purity of furfural. Initially, the condensed liquid from the dehydration unit is fed into a distillation column (C-1301), where furfural and water are recovered as distillate (top) and the bottom stream, which contains diluted acid from the distillation column. The bottom stream is directed to the wastewater treatment unit. Meanwhile, the distillate is cooled (E-1303) to 40 °C before it is diverted to a liquid-liquid extraction column (C-1302), where butyl chlorine is introduced as a solvent to break up the azeotrope of furfural and water **(Nhien et al. 2021)**. Fresh butyl chloride from its storage tank is mixed with the recycled butyl chloride from the distillation column (C-1303) in a mixer (M-1301) and heated (E-1304) to 40 °C **(Nhien et al. 2021).** The heated solvent is fed to the extraction column (C-1302), where the water is separated in the bottom stream, recycled to the pretreatment unit, and 5 wt.% of water is purged to avoid accumulation in the production system **(Nhien et al. 2021)**. The furfural-butyl chloride stream from the top is sent to the second distillation column (C-1303) for solvent recovery **(Nhien et al. 2021)**. The furfural-butyl chloride mixture is separated using a distillation column (C-1303), where the purity of furfural is achieved to a value of 99.54 wt.%. The purified furfural (bottom) is cooled to 30°C, which is then transported to the storage tank; the remaining solvent is left as a distillate from the distillation column for recycling.

2.2.5 Glucose synthesis

Glucose synthesis is separated into two units, i.e., neutralization and saccharification units. The wet solid stream from pretreatment is fed into an agitated mixing tank (M-1402). The diluted base (neutralizing liquid), which contains 50 wt.% of sodium hydroxide, process water, and the recycled water from the glucose purification unit (C-1601), are pre-mixed in an agitated mixing tank (M-1401) before being further mixed in M-1402 **(Humbird et al. 2011)**. The flowrate of sodium hydroxide is fed such that it is equimolar to the moles of sulfuric acid and acetic acid in a wet solid stream while the process water flowrate is adjusted to account for 20% solid loading in M-1402 **(Humbird et al. 2011)**. To account for complete neutralization of acetic acid, sodium hydroxide is fed 10 mol.% in excess **(Humbird et al. 2011)**. In this reaction, sulfuric acid and acetic acid are converted into sodium-based salt (e.g., sodium sulfate and sodium acetate). After neutralization of acids, the pH of the slurry is increased to pH 5, and it is further heated (E-1401) to 48 °C **(Humbird et al. 2011)**.

In the saccharification unit, the heated slurry from the neutralization unit is fed into an insulated CSTR (R-1501), which operates at 48 °C and 1 atm **(Humbird et al. 2011)**. Cellulase enzyme from its storage tank is fed into the CSTR (R-1501) **(Humbird et al. 2011)**. With enzymatic hydrolysis, 95.20 wt.% of the cellulose is converted into glucose **(Humbird et al. 2011)**. The residence time of the reaction is two successive days (48 h). The slurry from CSTR (R-1501) is pumped into a decanter centrifuge where all the solids that contain 5 wt.% moisture are removed and sent to the boiler feed unit as fuel stock while the liquid stream is sent to the final stage for glucose purification.

2.2.6 Glucose recovery

The glucose mixture from the saccharification unit contains glucose, water, and other impurities. For glucose purification, the glucose mixture from the saccharification unit is fed into a distillation column (C-1601), where most glucose can be separated in the bottom stream. The distillation column contains four stages with a reflux ratio of 1, and the feed is at the second stage from the top. This separation method increases glucose purity to 96.67 wt.%. The remaining water is removed as a distillate from the distillation column and chilled (E-1603) to 30 °C before being recycled to the neutralization unit with a 5% purge rate. Finally, the glucose stream is cooled (E-1604) to 30 °C and pumped into the glucose storage tank.

3. Process simulation

Aspen Plus features a variety of modules and thermodynamic databases for simulating chemical processes. Therefore, the integrated process of glucose and furfural production from EFB (Fig. 3) is simulated using Aspen Plus V10, with the process conditions obtained from the literature review **(Dolphin Centrifuge 2021; Humbird et al. 2011; Kenthorai Raman & Gnansounou 2015; Loh 2018; Mittal et al. 2017; Nhien et al. 2021)**. Here, the feed preparation unit in Fig. 3 is not included in the process simulation since Aspen Plus limits the usage of solid handling and batch processes. Therefore, it is assumed that the EFB is dried and ground. In some major unit-operation, there is a lack of information on equipment design and operating parameters. Hence, the following assumptions are made in the present work:

- The process is continuous and in a steady state.
- There is no loss of pressure and temperature across equipment and pipelines.
- All transfer pumps increase by 0.1 atm to compensate for pressure loss in the pipeline.
- Positive displacement pumps are used for streams with a capacity lower than 0.55 m^3/h and streams with solid-liquid slurry **(Liquiflo 2016),** while others rely on centrifugal pumps **(Liquiflo 2016)**.

Fig. 3. *Simulation flow diagram of the integrated process.*

3.1 Set up of Aspen simulation

3.1.1 Property Method

This study employed NRTL as the thermodynamic property package, which is identical to the existing studies in literature **(Humbird et al. 2011)**. However, the binary interaction between certain components such as butyl chloride and furfural was not included in the NRTL; therefore, UNIQUAC property package can be used for vapor-liquid equilibrium (VLE) as well as liquid-liquid extraction (LLE) **(Manual 2001)**. Thus, NRTL is chosen as the base property method while UNIQUAC property method is added to the referenced method in Aspen Plus.

3.1.2 Defining Components

All liquid components are available in the Aspen database. However, the solid component, i.e., EFB; however, is not available in the Aspen database. Therefore, it has to be added manually by defining the corresponding properties such as molecular weight, solid enthalpy of formation, solid molar volume (VSPOLY-1), and solid head capacity (CPSPO1-1). This information of which can be .obtained from the literature **(Wooley & Putsche 1996)**.

3.2 Sensitivity analyses

Sensitivity analyses are conducted to maximize the yield and purity of furfural and glucose in the product purification units. The influence of 4 manipulated variables is examined in order to determine the optimum operating conditions. Two-point values are applied to examine the influence of the higher and lower bounds on the parameters to which the yield and purity of products are most sensitive. As Aspen Plus has calculated the binary interaction between the molecules, the reflux ratio, and a number of stages in the distillation column are varied to compare the purity and flowrate of the product. In addition, a solvent used in the liquid-liquid extraction column and the purge percentage of both recycled streams are varied to determine the most suitable recycle flowrate and save cost.

3.2.1 Number of stages

Fig. 4 shows the product composition in the distillation columns 1 to 3 and the liquid-liquid extraction column. The number of stages of distillation columns 1 and 2 is varied from 5 to 15. Based on Fig. 4, distillation column 1 gives the highest purity and flowrate of furfural at Stage 10^{th}; distillation column 2 achieves the maximum purity and flowrate of furfural at Stage 13^{th}. On the other hand, the number of stages of the final distillation column ranges from 2 to 10. The maximum flowrate and purity of glucose were found to be 34.86 kg/h and 98.18 wt.%, respectively, at the 3^{rd} stage before the flowrate and purity declined slightly to 34.86 kg/h and 98.15 wt.%, respectively. For the LLE column, the optimum stage was found to be the 7^{th} stage. From Fig. 4, it is apparent that the number of stages only provides a marginal effect on the furfural flowrate and purity. Therefore, it is not recommended to increase the number of stages as the marginal increase in the furfural flowrate and purity were achieved at the expense of an increase in the column capital cost.

Fig. 4. *Variation of number of stages of (a) distillation column 1, (b) distillation column 2, (c) distillation column 3, and (d) liquid-liquid extraction column.*

3.2.2 Reflux Ratio

Fig. 5 shows the effect of the reflux ratio on the composition and yield of products. The reflux ratio in all three distillation columns was varied from 0.7 to 1.5. Based on Fig. 5, distillation column 1 gives the highest purity and flowrate of furfural at a 1.5 reflux ratio. For distillation column 2, the optimum reflux ratio is observed at the intersection point between both lines, which is 0.9. For distillation column 3, the increase in reflux ratio has no significant effect on the flowrate and purity of products; thus, original reflux ratio of 1 is used. Overall, the optimal reflux ratio is recorded at 1.5 for distillation column 1 and 0.9 for distillation column 2. Meanwhile, the reflux ratio of distillation column 3 remains at 1, based on the original literature review data **(Nhien et al. 2021)**.

Fig. 5. *Variation of reflux ratio of (a) distillation column 1, (b) distillation column 2 and (c) distillation column 3.*

3.2.3 Solvent type

The type of solvent is analyzed in a liquid-liquid extraction column (C-1302) between butyl chloride and toluene. According to Nhien et al. **(2021)**, the feed-to-solvent ratio for furfural is lower for toluene at 2.68, whereas butyl chloride has a feed-to-solvent ratio of 7.68. From process simulation using Aspen Plus, the purity and flowrate of furfural were studied for both solvents, and the result is recorded in Table 5. Based on the table, butyl chloride and toluene solvents produce similar furfural purity (99.98 and 99.50 wt.%, respectively) and flowrate (17.20 and 17.27 kg/h, respectively). Thus, both are promising solvents, and a literature review is performed to select the most suitable solvent. Although a higher amount of butyl chloride is needed, as mentioned previously, the amount of energy required in the butyl chloride case is less than that in the toluene case **(Nhien et al. 2017)**. With this regard, Nhien et al. showed that butyl chloride is the best solvent for the furfural manufacturing process, saving 44.70% of the total annual cost while lowering CO_2 emissions by 45.50%, in comparison to using toluene and benzene as solvents. Therefore, the plant expenses can be saved. The selling price of furfural can be increased due to its higher grade and purity when butyl chloride is selected as a solvent in the furfural production unit.

Table 5. *Furfural purity and flowrate from Aspen Plus simulation for two solvents.*

3.2.4 Recycle flow rate

To avoid the accumulation of undesirable chemicals in the process as a result of recycling, some parts of the recycling streams are purged before being recycled back into the process. In this sensitivity analysis, the first recycle stream (Recycle 1) consists of 99.5 wt.% water, 0.3 wt.% acetic acid, 0.1 wt.% butyl chloride, and 0.1 wt.% furfural, while the second recycle stream (Recycle 2) mainly consists of water (99.90 wt.%) with 0.10 wt.% of furfural as an impurity. Fig. 6 shows the effect of purge fraction on the flowrate and purity of products. The purge percentage for both Recycle 1 and 2 is varied from 0% to 10%. The first recycle stream with 99.50 wt.% of water recycled back into pretreatment unit achieves the optimum flowrate and purity of furfural (17.53 kg/h and 99.54 wt.%, respectively) at a 5% purge percentage. Based on Fig. 6, the flowrate of the furfural is inversely proportional to the purge percentage, while the purity of the furfural is directly proportional to the purge percentage. In this aspect, decreasing the mass flowrate of recycling streams improve the purity of furfural. Due to minor improvement in flowrate and a slight decrease in purity of furfural, the optimum purge percentage is recorded at the intersection point between both lines. The second recycling stream which recycles 99.90 wt.% of water back to the neutralization unit, has an optimum glucose purity and flow rate (34.86 kg/h and 98.25 wt.%) at a purge percentage of 10%. In this case, the purge fraction has no significant effect on the glucose flow rate while the purity of the glucose increases with increasing purge percentage. Overall, 95% and 90% of water is recovered and recycled back to the pretreatment and neutralization units for reuse purposes, respectively.

Fig. 6. *Variation of purge percentage of (a) recycle 1 and (b) recycle 2.*

3.3 Simulation Results

The optimization results are recorded in Table 6 and 7. Based on the integrated process of glucose and furfural production from EFB, the desired final product purity is obtained using a basis of 100 kg/h of dry EFB. After optimization, the final purity and flowrate of the product are achieved with the absence of cellulase enzymes in the final product. Further, the process is optimized by changing the equipment design based on optimum operating conditions to maximize the purity and flowrate of final products. Therefore, optimized simulation attains better plant efficiency and minimizes waste produced. After optimization, the mass flowrate is scaled by adjusting the feedstock (dry EFB) to attain the annual plant capacity (i.e., 10 ktpy of glucose and 4.96 ktpy of furfural), where the plant can achieve a purity of 98.21% for glucose and 99.54% for furfural. Besides the purity and flowrate of products, the overall heating duty of the plant is reduced to 3385.91 kW, whereas the overall cooling duty is reduced to 2247.39 kW after optimization. According to Table 7, the amount of heating and cooling duties is reduced by 0.27% and 0.75%, respectively, which can minimize energy consumption and production costs in the economic analysis.

The experimental work of **Montané et al. (2002)** revealed that the hydrolysis of lignocellulosic biomass with H_2SO_4 yields a range of 50 to 65% of furfural. Similarly, the experimental work of **Jung et al. (2013)** revealed that 88.5% of glucose yield was determined after 48 hours of the hydrolysis process. Both furfural and glucose yields (67.59% and 94.45%, respectively) in this study are higher than the findings in the experimental work. Meanwhile, the furfural and glucose purities (99.54 wt.% and 98.21 wt.%, respectively) obtained from the optimized Aspen Plus simulation were indeed higher than those of the commercially available furfural (98 wt.%) and glucose (98 wt.%) **(Fisher Scientific 2022; Sigma-Aldrich 2022)**, thus demonstrating the feasibility of the proposed process design.

Table 6. *Final product purity and flowrate from Aspen Plus simulation after optimization.*

Table 7. *Heating and cooling duties from Aspen Plus simulation after optimization.*

4. Economic analysis

For economic analysis, the total capital investment is calculated using all the equipment costs according to its capacities and flowrate of feedstock, product, and utility costs. The summary of feedstock, product, and utilities is shown in Table 8.

Table 8. *Feedstock, product, and utility flowrate.*

4.1 Fixed capital investment

Fixed capital investment (FCI) is calculated to build a new plant, including the building, construction, and equipment costs required. The fixed capital is calculated using equipment prices from 2006 and 2002 adjusted to inflation and location **(Peters et al. 2003; Sinnott & Towler 2020)**. The total fixed capital investment was found to be United States Dollar (USD) 20.80 million. Among them, the inside battery limit (IBL) cost alone covers 48% of the FCI as the plant uses various types and numbers of equipment.

Table 9. *Total fixed capital summary.*

4.1.1 Direct cost and Indirect cost

Direct cost can be divided into IBL and outside battery limit (OBL). IBL is the total cost of purchasing and installing all process equipment inside the process boundary; OBL is the total cost required for building and yard improvements along with the construction and installation of utility systems, storage, and waste treatment systems that are outside the process boundary.

IBL costs are adjusted along with installation, electrical system, control system, phase handle, and piping cost within the process boundary. The cost of equipment is determined from Chemical Engineering Design by Gavin and Sinnott at U.S. Gulf in 2006 and Plant Design and Economics for Chemical Engineers by Peters in the USA in 2002 **(Peters et al. 2003; Sinnott & Towler 2020).** Finally, the cost of each piece of equipment can be calculated by adjusting the location factor to Malaysia, cost indices to 2020, and its lang factor accordingly.

In order to cover the expenses on food production safety procedures, 15% of additional equipment costs were assumed in the glucose production units. Remarkably, the 15% increase in equipment costs has covered various essential food safety factors, such as a) cleaning agents and disinfectants for CIP, b) extra pump work and water needed for washing, flushing, and rinsing, as well as c) water that meets food safety standards, which can be purchased from portable water production **(Moerman et al. 2014)**. Besides that, sterilization is accomplished in the pre-treatment unit using the heat treatment technique, with the stream being heated to 158 °C using E-1101 and cooled to the saccharification temperature of 30 °C using E-1102 before entering the glucose production units **(Pina et al. 2014)**. In this study, hot pressure steam was selected for the sterilization since other sterilization methods such as gamma rays and ethylene oxide gas might prevent polymerization after repeated irradiation and cause potential toxicity **(Lightfoot & Maier 1998)**. Moreover, glucose was recovered at 110.18 °C as the final product from the distillate stream of the distillation column (C-1601), exceeding the minimum sterilizing temperature (110 °C). In this case, microorganisms such as coliforms, lactic acid bacteria, and *Bacillus* cannot withstand temperatures above 110 °C **(Błaszczyk & Iciek 2021)**.

There are a lot of critical parameters that need to be controlled in this study, such as a) temperature and b) pressure of the condenser or reboiler on the distillation column as well as c) product composition at the outlet of the distillation column in the product recovery units. In order to achieve commercial standards of furfural and glucose production, several process control philosophies are proposed, including installing temperature controllers, pressure controllers, and composition analyzers on the distillation column or its nearby stream **(Li et al. 2019)**. In this context, a composition analyzer was installed at the product outlet streams of the distillation column to ensure the purity of the products (i.e., furfural and glucose) met the commercial standard. Additionally, temperature and pressure controllers are installed on the distillation column to prevent unforeseen incidents, such as oversupply of hot pressurized steam and runaway reaction. Notably, in order to meet sterilization and cleaning standards of the food industry, the outlet temperature of glucose products at the distillate stream must be set above 110 °C to disinfect the glucose products **(Błaszczyk & Iciek 2021; Mironescu & Mironescu 2006)**.

Overall, the total cost of IBL is calculated as USD 10.50 million. The major cost driver of direct cost is the saccharification unit which takes 43.51% (USD 4.57 million) of IBL cost. This is because eight reactors are used for higher conversion into furfural using cellulase enzyme. Other than a reactor, a distillation column has a higher price in purification units.

Table 10. *Total equipment cost summary.*

OBL cost is identified for each facility which is estimated at a percent of FCI **(Peters et al. 2003)**. These general facilities include buildings, yard improvements, communication, sanitary disposal, and safety installation follow similar factors as well. The total cost of OBL was calculated as USD 4.66 million. On the other hand, indirect costs include legal expenses, construction costs, and contractor fees, and the cost estimation is similar to the OBL cost **(Peters et al. 2003)**. Overall, the total indirect cost is calculated as USD 5.65 million.

4.2 Operating cost

Operating costs comprise both the manufacturing and non-manufacturing costs of the plant. The manufacturing costs comprise variable and fixed costs, where variable cost includes labor wages, maintenance costs, insurance, and interest, and fixed cost includes raw materials and utility costs incurred annually **(Malaysia Indeed 2021; Peters et al. 2003)**. Non-manufacturing cost includes administration, research, and development, as well as distribution and marketing. The total unit operating cost and annual operating capital are calculated as USD 1138.66/tonne of glucose and USD 16.93 million/year, respectively. In the plant, cellulase enzyme has contributed 85.53% (USD 2.79 million/year) of feedstock cost due to the high cost of cellulase and the absence of an enzyme recovery unit. Nonetheless, employing cellulase improves the conversion of furfural and eventually resulting in higher profits.

Table 11. *Operating cost summary.*

4.3 Working capital

Working capital is the money needed to start up and run a plant until it generates enough revenue to pay off debt and purchase inventory **(Sinnott & Towler 2020)**. The raw materials and final products stock have three weeks inventory, and water has one week inventory as it is easily available

(Sinnott & Towler 2020). It is assumed that six weeks inventory value is allocated to both debtors and creditors. The working capital is calculated as USD 3.74 million with the previous inventories.

Table 12. *Working capital summary.*

4.4 Profitability evaluation

4.4.1 Cumulative cash flow

The production plant runs for 20 years in Malaysia, with an additional year for planning and design in 2021, two years for building and installation, and one year for decommissioning in the final year (2044). It is assumed that 70% of the total working capital is used in that year, and 15% of total working capital is shared equally in the following years as the growth rate in production is linear. With a declining balance, the fixed capital begins to depreciate at a rate of 13.91%. In Malaysia, a company that has a fixed capital of above USD 0.61 million pays a 24% corporate tax **(LHDN Malaysia 2020; XE currency 2021)**. The impact of inflation during the lifetime of plant operation is estimated to be negligible. Fig. 7 shows the plant recovering full FCI at the initial stage of the 9th year and profiting until the 22nd operating year. The remaining residual value and working capital are anticipated to be recovered at the end of the project.

Fig. 7. *Cumulative cash flow diagram (CFF).*

4.4.2 Net Present Value, Internal Rate of Return and Return on Investment

The discounted cash flow after tax is used to compute this project's net present value (NPV). The discount rate is assumed to be 10% as the source of funds between equity and loan is uncertain at this stage of the project. Based on the assumptions, the NPV is calculated as USD 7.65 million by summing discount cash flows after tax at a 10% discount rate throughout 23 years. The discount rate that gives a zero NPV is the internal rate of return (IRR). Fig. 8 shows a graph of NPVs against a range of discount rates where the IRR value is calculated to be 14.25%. Moreover, the return on investment (ROI) is determined using the annual operating profit and total investment, which is calculated to be 22.06%.

Fig. 8. *Net present value against discount rate per annum.*

4.4.3 Sensitivity analysis

Using the cumulative cash flow table, the various parameters affecting the profitability of the plant can be varied to calculate the change in NPV and IRR. With a positive IRR, the NPV of the plant increases, thus increasing its profit. On the other hand, negative NPV and IRR indicated a decrease in plant profit. This parametric change must be avoided to maintain the profitability of the plant.

Fig. 9. *(a) change in NPV with various parameter changes; (b) change in IRR with various parameter changes; (c) Parameter change for sensitivity analysis.*

The tornado chart shows that capacity is the major cost driver and cutting capacity by 20% dramatically reduces the NPV and IRR. To make a plant profitable, its capacity should be expanded or maintained. Aside from capacity, the selling price of furfural significantly influences NPV and IRR. From the market analysis, the furfural selling price is very elastic and has fluctuated drastically over the past few years **(Mordor Intelligence 2021)**. Furthermore, the fixed operating costs, FCI, and glucose selling price have a lower impact on the profitability of the plant.

4.5 Plant Viability

Constructing an integrated process plant in Malaysia that uses EFB to manufacture glucose and furfural will result in a profitable process plant. According to market research, the demand for glucose and furfural is increasing rapidly in Malaysia **(Persistence Market Research 2020)**. Therefore, the process plant can be upscaled in the future for a higher profit. Nonetheless, the plant's capacity is not projected to decline, implying that the plant will be profitable in the future. In addition, the selling price of furfural fluctuates greatly, currently showing a lower value in 2021 **(Mordor Intelligence 2021)**. As Malaysia is going through the phase of vaccination for the COVID-19 coronavirus, the economy of all the process plants is affected. After the vaccination phase, the country's economy is expected to recover, leading to a higher product selling price and higher NPV.

In 2020, the palm oil sector was the greatest contributor to the Agricommodity crops of Malaysia, contributing USD 11.65 billion, or 3.6% of the total gross domestic product (GDP), and is forecasted to expand to USD 16.75 billion by 2030 **(Ministry of Plantation Industries and Commodities 2021; XE currency 2021)**. However, the potential of downstream activities has not yet been fully developed, resulting in the abundance of biomass produced by the industry. Hence, this proposed integrated plant that produces value-added products (glucose and furfural) in palm oil downstream segments is critical in achieving the GDP target of USD 16.75 billion by 2030 in the palm oil industry.

In terms of operating costs, the cellulase enzyme (catalyst) is the major cost driver of the process plant using EFB, which is cheap biomass available in Malaysia. As the research is conducted through simulation, the flow of catalyst and catalyst recycling is not considered, where installing an enzyme separation and recycling stream can reduce the variable cost greatly. In terms of FCI, the enzymatic reactor has the highest equipment cost because of its large residence time. Optimizing the reactor process condition can reduce the residence time resulting in a smaller reactor. Finally, various parameters can be further optimized in the process plant in the future, allowing for increased profit and sustainability.

5. Conclusion

In this work, an integrated process plant producing both glucose and furfural from palm oil empty fruit bunches (EFB) is designed and simulated using the most suitable process conditions to maximize the yield and purity of the product. The mass and energy balance results are obtained, and a process flow diagram is drawn to support the equipment used in the plant. For a production capacity of 10 ktpy of glucose and 4.96 ktpy of furfural, purity of 98.21% and 99.54% is achieved, respectively. The plant economy is analyzed by obtaining the fixed capital investment (FCI), operating costs, and working capital, where the profitability is analyzed using cumulative cash flow (CCF), net present value (NPV), and internal return rate (IRR). For the capacity of 10 kilotons per year (ktpy) of glucose and 4.96 ktpy of furfural, the FCI is calculated as USD 20.80 million. The working and operating costs are calculated as USD 3.74 million and USD 16.93 million, respectively. On the evaluation of the profitability of the plant, the NPV, IRR, and return on investment (ROI) of the process plant are calculated as USD 7.65 million, 14.25%, and 22.06%, respectively. From the obtained results, it can be concluded that the designed process plant is profitable, and this profit can be increased with optimization, which is suggested to be included in the future. Therefore, the designed integrated process plant reduces the waste from the palm oil mill, which makes it both a profitable and sustainable plant. Integrated production of furfural and glucose using EFB-hemicellulose biomass is a promising pathway. Considering the residue (i.e., cellulose and lignin) produced from the proposed plant, future work on converting the residue to valuable products is suggested. For instance, hydrogen that acts as an alternative clean energy carrier to non-renewable fossil fuels can be produced from the EFB residue. To the best of our knowledge, no commercial EFB-based hydrogen production plant is currently available. Indeed, it is worthwhile to study hydrogen production from EFB components so that, in the future, not only food and chemicals but also food, chemicals, and energy can be produced from EFB to fully utilized the potential of EFB. Potential synergistic utilization of similar raw materials and utilities in the food, environmental, and energy nexus to produce value-added outputs makes co-location principles economically and commercially attractive.

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Figure and table list

Figure List

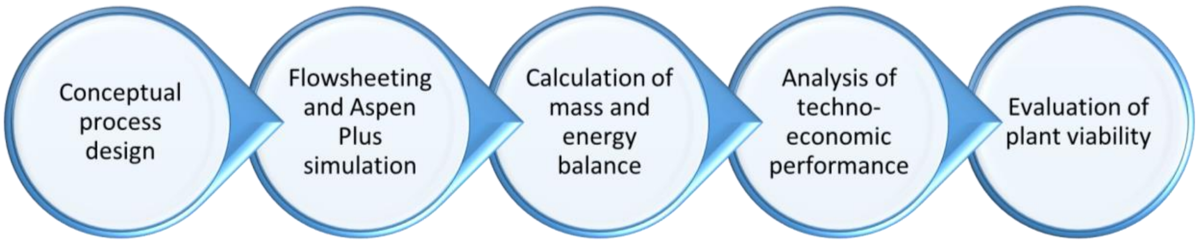


Fig. 1. Research flow chart.

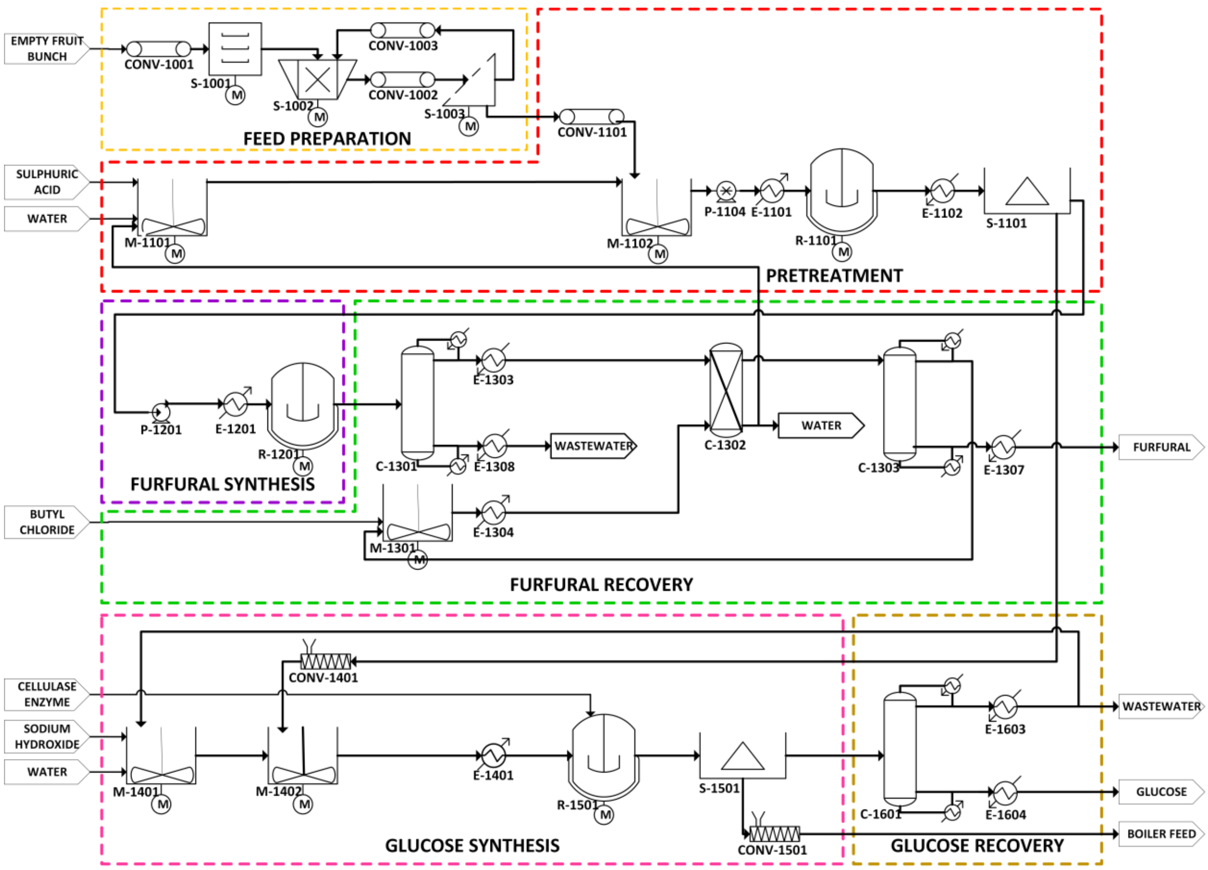


***Fig. 2.** Integrated production of glucose and furfural from palm oil empty fruit bunch.*

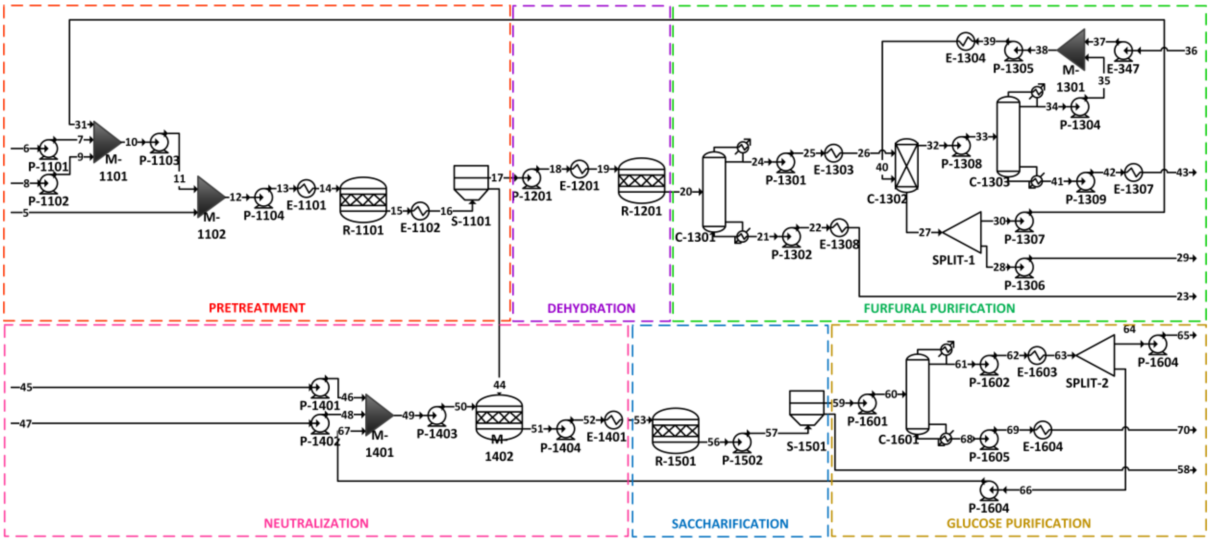


Fig. 3. *Simulation flow diagram of the integrated process.*

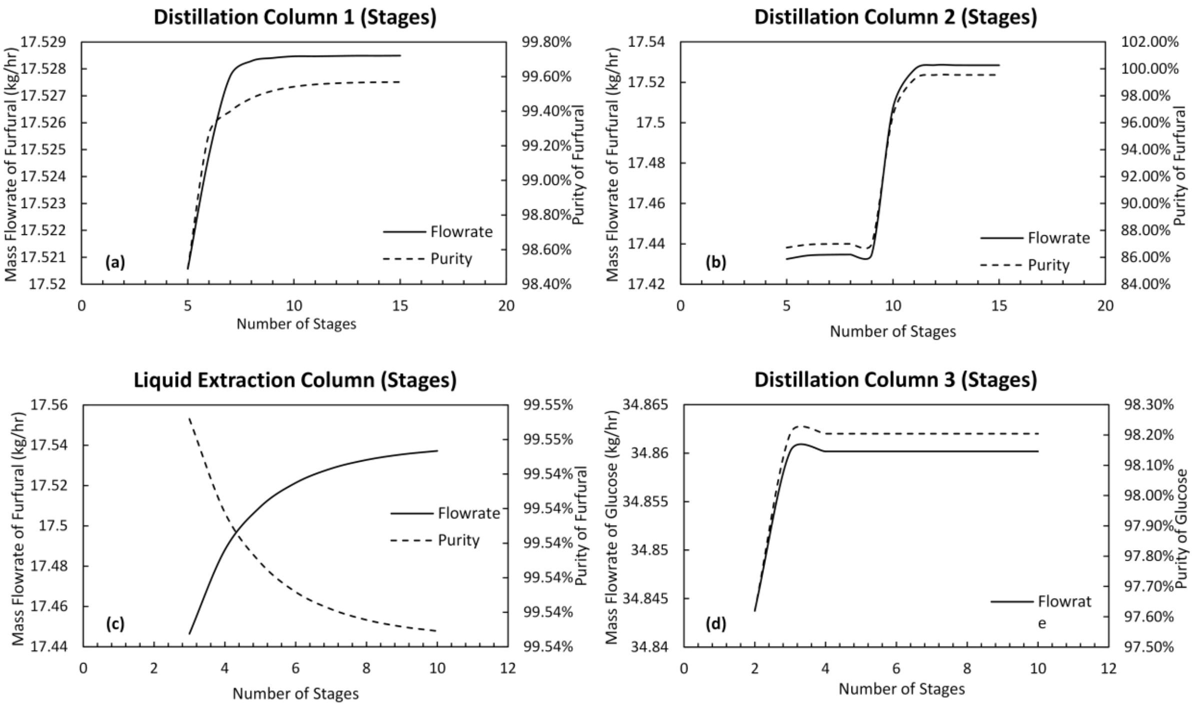


Fig. 4. *Variation of number of stages of (a) distillation column 1, (b) distillation column 2, (c) distillation column 3, and (d) liquid-liquid extraction column.*

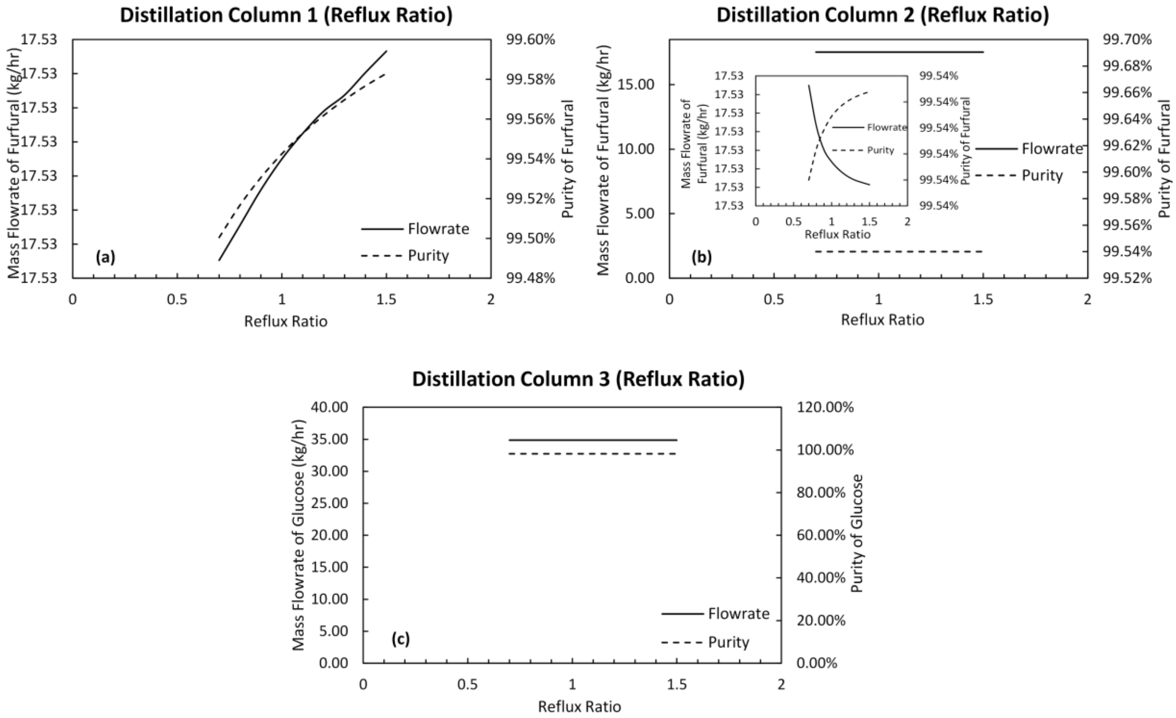


Fig. 5. *Variation of reflux ratio of (a) distillation column 1, (b) distillation column 2 and (c) distillation column 3.*

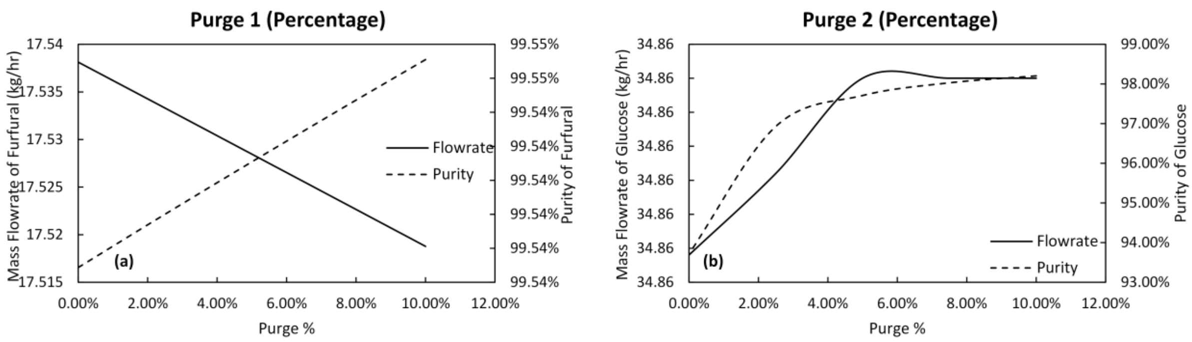


Fig. 6. *Variation of purge percentage of (a) recycle 1 and (b) recycle 2.*

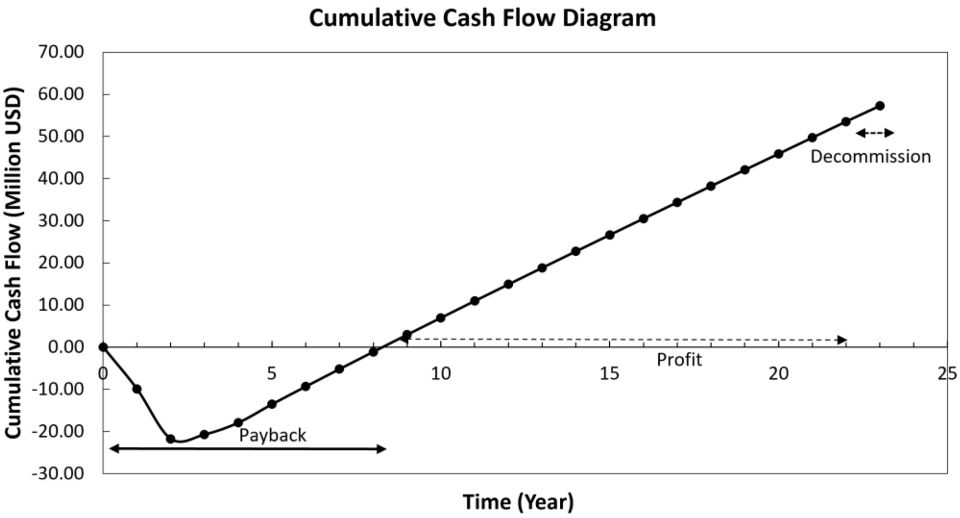


__Fig. 7.__ Cumulative cash flow diagram (CFF).

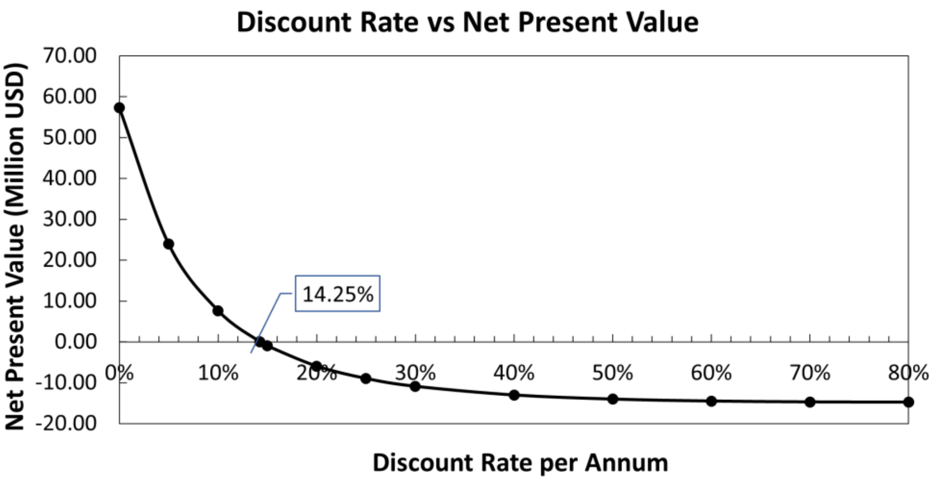
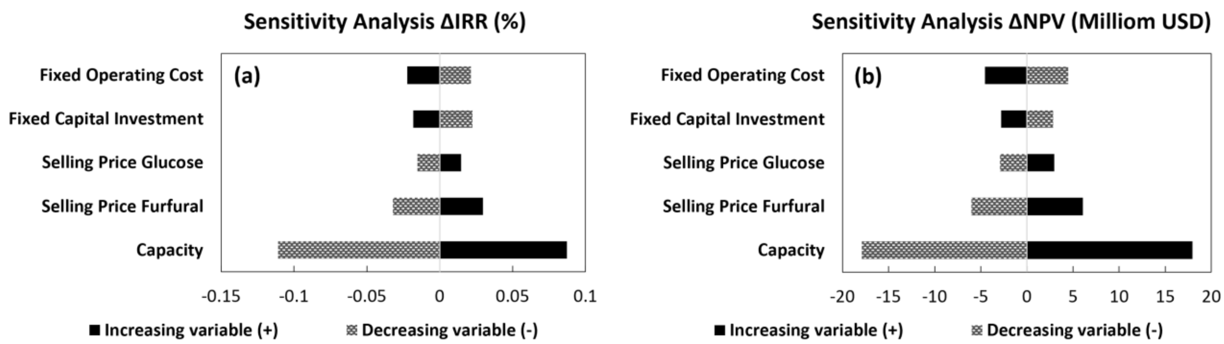


__Fig. 8.__ Net present value against discount rate per annum.



(c)

Parameter	Variation (%)
Total Capacity	± 20
Selling Price Furfural	± 10
Selling Price Glucose	± 10
Fixed Capital Investmer	± 15
Fixed Operating Cost	± 10

Fig. 9. *(a) change in NPV with various parameter changes; (b) change in IRR with various parameter changes; (c) Parameter change for sensitivity analysis.*

Table List

Table 1. *Composition of EFB (wt.%) (Reproduced from the work of **(Chiesa & Gnansounou 2014)**).*

Components	Sub-Components	Percentage (wt.%)
Cellulose	Glucan	29.60
Hemicellulose	Xylan	20.80
	Acetyl group	1.50
Lignin	Acid insoluble Lignin	20.70
	Acid soluble Lignin	2.20
Others	Ash	5.40
	Water	19.80

Table 2*. Properties and composition of furfural (Reproduced from the work of **(Hongye Holding Group Corporation Ltd. 2020)**).*

Components	Value	Units
Purity	98.50	wt.%
Moisture	0.20	wt.%
Acid mol/L	0.02	%

Table 3. *Properties and composition of glucose (Reproduced from the work of **(21Food 2021)**).*

Components	Value	Units
Dry Solids	80.00–84.00	wt.%
Moisture	16.00–20.00	wt.%
Dextrose Equivalent (DE)	42.00–45.00	%
Sulfate Ash	0.40	wt.% max
SO_2	200.00	ppm
pH	4.00–6.00	pH

Table 4. *Reaction and its conversion in CSTR (R-1101).*

Components	Reactant	Reaction	Conversion
Cellulose	Glucan	$(Glucan)_n + n\ H_2O \rightarrow n$ Glucose	10.30%
Hemicellulose	Xylan	$(Xylan)_n + n\ H_2O \rightarrow n$ Xylose	97.40%
	Xylose	Xylose $\rightarrow$ Furfural + 3 H_2O	5.00%
Lignin	Lignin	$(Lignin)_n \rightarrow n$ Soluble Lignin	5.00%

Others	Acetate	Acetate → Acetic Acid	100.00%

Table 5. *Furfural purity and flowrate from ASPEN simulation for two solvents.*

Solvent	Furfural purity (wt.%)	Furfural flowrate (kg/h)
Toluene	99.50	17.29
Butyl Chloride	99.98	17.20

Table 6. *Final product purity and flowrate from ASPEN simulation after optimization.*

Product	Purity (wt.%)	Flowrate (kg/h)	Yield (%)
Glucose	98.21	34.86	94.45
Furfural	99.54	17.53	67.59

Table 7. *Heating and cooling duties from Aspen Plus simulation after optimization.*

Type of duty	**Base value** (kW)	**Optimum value** (kW)	Net savings **(%)**
Heating duty	3395.18	3385.91	0.27%
Cooling duty	2264.31	2247.39	0.75%

Table 8. *Feedstock, product, and utility flowrate.*

Feed stream		**Product stream**	
Feedstock	**Flowrate (kg/h)**	**Product**	**Flowrate (kg/h)**
EFB	4435.10	Glucose	1262.63
Sulfuric acid	91.46	Furfural	626.34
Process water	1560.10	Lignin	1402.42
Butyl chloride	2.44	Wastewater	2318.05
Sodium hydroxide	3.23		
Cellulase enzyme	23.55		
Utility stream			
Utility	**Description**	**Value**	**Unit**
Hot Air	5% relative humidity, 105 °C	19762.24	kg/h
Cooling Water	30 °C	1446933.58	kg/h
Hot Water	80 °C	5840.99	kg/h
Low-Pressure Steam	3 bar saturated vapor	15001.77	kg/h
High-Pressure Steam	20 bar saturated vapor	31230.91	kg/h
Electricity	unit	177.02	kWh/h

Table 9. *Total fixed capital summary.*

Fixed Capital Cost (FCI)	Cost (million USD)
Total IBL cost	10.50
Total OBL cost	4.66
Indirect cost	5.65

Total FCI	**20.80**

Table 10. *Total equipment cost summary.*

Unit	Cost (million USD)
Feed preparation unit	1.77
Pretreatment unit	0.72
Dehydration unit	0.46
Glucose purification unit	2.12
Neutralization unit	0.22
Saccharification unit	4.57
Furfural purification unit	0.63
Total equipment cost	**10.50**

Table 11. *Operating cost summary.*

Operating Cost	Unit Cost (USD/ tonne product)	Cost (million USD/y)
Feedstock	326.96	3.27
Utilities	78.79	0.79
Process labor cost	22.43	0.22
Maintenance cost	153.18	1.53
Operating supplies	22.98	0.23
Plant overhead	153.18	3.27
Insurance and interest	131.30	1.31
Depreciation cost	304.40	3.04
Non-manufacturing cost	47.55	3.27
Total operating cost	**1240.77**	**16.93**

Table 12. *Working capital summary.*

Working Capital	Cost (million USD)
Feedstock inventory	0.21
Product inventory	1.01
Debtor amount	2.02
Creditor amount	0.50
Total working capital	**3.74**