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Conceptual design of an industrial-scale plant for ethanol production from by-products of sweet potato processing

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Abstract

Inclusion of bioethanol in the transportation fuel by displacing part of the petroleum-derived energy has become an emerging trend due to limited availability of fossil fuel and its potential association with global warming. With a suitable strain of microorganism, sugar from the agricultural waste could be fermented to produce bioethanol, which is a highly effi cient energy source generating minimal environmental impact. In this study, a conceptual design of ethanol production by Saccharomyces cerevisiae converting the by-products from sweet potato processing was proposed based on experimental data and previous publications. An industrialscale plant was designed to produce 5 million liters ethanol annually, accounting for approximately 5% of the total sweet potatoderived ethanol production in Taiwan. The fermentation section of the plant consists of two feed tanks, in which sweet potato peel is pre-treated, diluted and standardized, one batch (aerobic) reactor for microbial culture enrichment, and two parallel systems each consisting of two continuous (an aerobic and an anaerobic respectively) reactors for further culture enrichment and ethanol production.

Keywords: Ethanol, Sweet potato waste, Saccharomyces cerevisiae, Bioenergy, Plant design

Introduction

As the world economic growth and technological advancement are both highly rely on the sufficient energy supply, which currently are mainly based on the non-renewable energy sources. Thus, the greatest challenge for the modern society is to find a sustainable way to meet the growing demand of energy used in transportation, electricity, and industrial processes. When the most world utilized energy source [1], fossil fuel, imposes serious environmental stress through generating excessive polluting gases, the renewable biomass energy is now widely regard as the most promising alternative for humankind to meet their increasing energy need with lower environmental cost [2].

Biomass energy currently shares approximately 10% of the world energy supply, which is more than 2 times higher than nuclear power [1]. Biomass being broadly defined as all matters derived from animals and plants is being used as energy source with or without processing. Unprocessed biomass, such as wood or animal waste, is traditionally used to produce energy for heating or power generating. In this sense, processed biofuel made from conversion of biomass into the more usable secondary form includes liquid biofuel: bioethanol and biodiesel. In specific, more recent attention has drawn onto the potential of bioethanol to displace the fossil-based transportation fuels that majorly contribute to the warming global climate [3]. Through fermentation, bioethanol can be produce from any raw material high in sugar or starch content [4]. In 2014, United States (58%) and Brazil (25%) are being the first and second fuel ethanol producer [5] using corn and sugar cane as the major feedstock [6], respectively. However, the generation of bioenergy using energy crops has also arouse opposition in its

growing at the expense of crops used for human food [7,8]. To avoid competition between food and fuel production, agricultural wastes from food processing contain considerable amount of residual starch and lignocellulose are now being regard as the better alternative feedstock for sustainable bioethanol production [6, 9-10].

Sweet potato (Ipomoea batatas) is a root vegetable that belongs to the Convolvulaceae family. It is native to South America and is now widely cultivated throughout the tropical and warm regions that could provide adequate sunlight exposure and water for its growth [11]. In Taiwan, the subtropical climate allows the production of high quality sweet potato, which is an important source for local human food and animal fodder. In 2013, the cultivation acreage of 9662 hectares gave approximately 215,093 metric tons of annual sweet potato production [12]. The starchy, sweat tasting tuberous root of sweet potato makes it a popular crops to be further processed into variety of dessert, snacks, and confectionary products, such as sweet potato cakes, chips, and pies [13]. The production of these products then yield substantial amount of manufacturing by-products in forms of sweet potato peel available for further utilization [14]. Moreover, the high starch content of sweet potato peel qualifies it an ideal feedstock for bioenergy production, providing raw material for ethanol generation and supporting the growth for microorganism [15-17].

In present study, a conceptual industrial-scale plant was designed for ethanol production from sweet potato residues by S. cerevisiae based on the previous published data with special consideration on the Taiwanese local agricultural and industrial environment. The conclusions of this work will be great reference for the researchers in this area and industrial ethanol producers when considering building a large-scale bioethanol production plant using sweet potato by-products as feedstock.

Materials and Methods

Raw material supply. The ethanol production plant needs to be constructed near to major sweet potato plantation fields and processing factories in order to minimize the transportation cost of sweet potato waste. The ideal area for building an ethanol production factory from sweet potato waste is suggested to be located in Tainan, which is in close proximity to many large scale sweet potato processing factories. The processing process generates huge amounts of starch rich sweet potato wastes. For example, O-Nong Co., Ltd claims that only 70% of the sweet potato corm is extracted (middle core extracted) and used as raw material for sweet potato products, leaving 30% of sweet potato wastes. The sweet potato wastes are hollowed sweet potato corms, composed mainly of sweet potato skin and inedible attaching sweet potato scrapes. Since the ethanol production technique is mature in Taiwan, a bioethanol production plant using sweet potato waste is proposed.

Our target is to produce 5 % of the annual ethanol demand of Taiwan, which is equal to about 5 million liters of ethanol [18]. This number corresponds to16,667 liters per day, which is equal to about 13.15 metric tons ethanol. To estimate the amount of sweet potato waste input needed, the reaction stoichiometry on a weight basis is implemented [19]:

Starch + 0.1111 Water → 1.1111 Glucose

Glucose →0.4589 Ethanol + 0.4641 CO2 + 0.05 other

Apparently, the yield of ethanol is related to starch composition in sweet potato waste, conversion efficiency of starch to glucose, and conversion efficiency of glucose to ethanol. Based on the result of our previous study, the moisture content of sweet potato waste is approximately 42.3% while the starch content is estimated to be approximately 35.5% (as is) or 61.5% (dry weight basis). Assuming 100% efficiency for both conversion processes, a total of 73 metric ton per day of sweet potato waste is needed. The annual yield of fresh sweet potato in Taiwan is approximately 215,093 metric tons. The strategic location of the proposed plant is estimated to secure most of the sweet potato waste in Taiwan, which equals to approximately 64,528 metric ton of sweet potato waste per day, producing about 14,432 liters ethanol per day. The average density of sweet potato waste mash is accepted to be 1,400 kg/m^3 [20].

Microorganism and culture preparation. S. cerevisiae ATCC 36858 [21] is used throughout this investigation. The medium for the culture growth is described in Table 1, and was sterilized after dissolving the contents (total 150 mL) in Erlenmeyer flasks. The selective medium will be inoculated with the culture and incubated at 30oC for 48 hours, and then stored under 4 °C

Composition of the selective medium for *S. cerevisiae*		Sweet potato Waste Medium for *S. cerevisiae*		Kinetic parameters for *S. cerevisiae*
Glucose	(50 g/L)	Sweet potato waste	(45 g/L)	μ_{max} = 0.3 h^{-1}
Yeast extract	(6 g/L)	Yeast extract	(6 g/L)	$Y_{x/s}$ = 0.3 g/g
$(NH_4)_2SO_4$	(4 g/L)	$(NH_4)_2SO_4$	(4 g/L)	K_s = 2 g/L
$MgSO_4.7H_2O$	(1 g/L)	$MgSO_4.7H_2O$	(1 g/L)	$Y_{x/s}$*= 0.05 g/g
KH_2PO_4	(1.5 g/L)	KH_2PO_4	(1.5 g/L)	$Y_{p/s}$ = 0.46 g/g
$CaCl_2.2H_2O$	(0.3 g/L)	$CaCl_2.2H_2O$	(0.3 g/L)	X_m = 45 g/L
				X_0 = 10 g/L

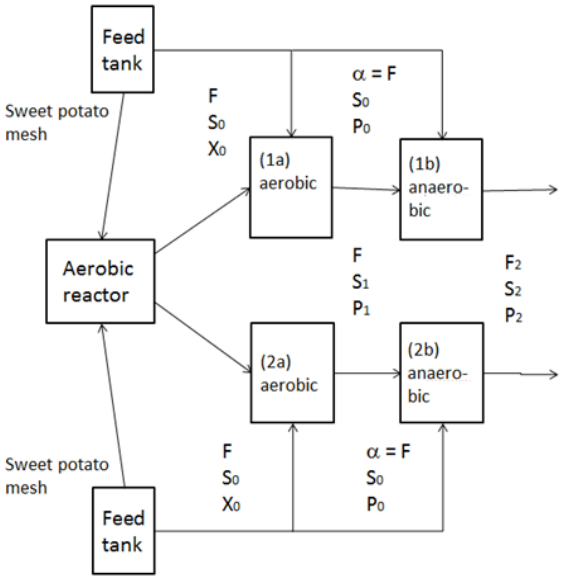
Table 1 Composition of selective medium, Sweet potato Waste medium and kinetic parameters for S. cerevisiae [25-27]

Figure: 1 Appearance of sweet potato waste. The outer part (brown) was sweet potato skin while the inner part (yellowish) was edible sweet potato scrapes.

Sweet potato waste will undergo blending, sieving, liquefaction, saccharification, dilution, standardization and sterilization before entering the fermentation chain. When receiving sweet potato waste, inspection will be conducted on the presence of potentially harmful compounds that affect the growth of microorganisms [22]. After blending the sweet potato waste into waste mash and passing through a 5 mm screen, the sweet potato mash will be diluted to 1.2-16% starch with water. Minute amount of ammonium phosphate will be added to prevent the growth of spoilage microorganisms. Sweet potato mash will be adjusted to pH 4.5 with 1 N H2SO4 [23]. Ammonium phosphate and sulfuric acid were purchased from chemical company. Liquefaction of sweet potato waste will be done by adding 10 mL/L of α-amylase (944 Units/mg protein) and incubate at 95 oC. For saccharification, 0.8 mL/L (300 Unit/ml) of amyloglucosidase will be added to liquefied sweet potato waste and incubated at 60 °C for 72 h (10). In the standardization tank, liquefied and saccharified sweet potato waste will first be diluted with processed water, following by addition of ammonium and phosphate salts to balance the minerals in the feed solution [24]. Later, the feed solution will be sterilized at 121.1 oC for 21 min. In the fifth step, ethanol will be separated from the fermentation broth in a continuous distillation column system and the ethanol-free medium will be recycled to the fermenter.

Bioreactor design. The bioreactors consist of one batch pre-fermenter (enrichment tank), where the biomass is produced by aerobic fermentation, one continuous aerobic fermenter, where biomass concentration is further increased, and continuous anaerobic fermenters, where ethanol is produced. The number of bioreactors was determined using the graphical method, and explained in the following sections. The bench-top scale experiment was performed in 5-L bioreactors (Major Science, New Taipei City, Taiwan).

Figure 2: Schematic diagram of bioreactor systems



The initial assumptions of the system S0, X0, X1, and P0 are 110 g/L, 0 g/L 25 g/L and 0 g/L, respectfully.

Table 2: Annual cost evaluation

Item	Price (NTD)	Note
Sweet potato waste	0	Provided by sweet potato processing factories
Transportation	21,000	3,500/ 20 ft container by shipping (6 rounds).
Electricity	120,000	Estimated by average kwh of apparatus and NTD 1.6/ kwh
Water	24,000	Estimated by 30 units of water usage/season Inlet nozel : 100
Enzyme	7,000	Alpha-amylase and amyloglucosidase
Medium supplements	6,000	(NH4)2SO4 , MgSO4.7H2O , KH2PO4
Total	178,000	

In order to calculate the number of reactors, the kinetic parameters were selected from the range of aerobic and anaerobic fermentation studies.

Continuous system design. The production rate was determined from our target annual production capacity: 0.5% of the total ethanol demand of the Taiwan. Total ethanol demand is 1,000 million liters / year. Taking 0.5% of the total demand resulted in 5 million liters ethanol per year. Considering 300 working days in one year, 695 liters ethanol must be produced in one hour. However, since it is hard to produce this huge amount of ethanol in one series of reactors, two parallel systems (consisting of one aerobic and one anaerobic reactor) identical to each other will be built by just dividing the production rate to two equal parts.

Given that Yx/s is 0.3 g of biomass per g of substrate, the substrate concentration in the stream leaving the continuous aerobic fermenter (S1) is calculated by Equation 1, and is determined as 26.67 g/L.

Y_(X⁄S)= ((X_1-X_0))/(S_0-S_1)

Knowing the value of S1, D1 is found to be 0.279 h-1 according to Equation 2.

D_1= μ_1= (μ_max· S_1)/(K+S_1)

Since there are two parallel systems, the flow rate of the final streams (F2) leaving the anaerobic reactors is determined to be 3,821 L/h, as shown in the Equation 3.

2*F_2=prod^' rate *ρ_ethanol· 1/P_2

For each system; assuming α is equal to 1.2, the flow rate coming from the feed tank (F) is calculated to be 1,737 L/h. According to Equation 4, dilution rate for the continuous aerobic fermenter was found to be 0.279 h-1. Therefore, the working volume of this reactor (Vcw) is approximately6.2m3.

D_1= F/V

Assuming the same volume of reactors, D2 is found to be 0.614 h-1 based on Equation 5.

D_2= ((1+1.2)*F)/V

Assuming the head-space as 15%, total volume of the continuous fermenters is found to be 8 m3. Accordingly, the diameter (dc) and height (hc) of the fermenter are calculated based on 1/2 (diameter/ height) ratio, and found to be 1.72 and 3.44 m, respectively. In order to determine the optimum number of reactors to be used in the continuous system, graphical method was used. Production rate is calculated according to Equation 6. Figure 3 displays the plot of $(\frac{dX}{dt})_G$ versus X (arbitrary values between 0 and 33 g/L).

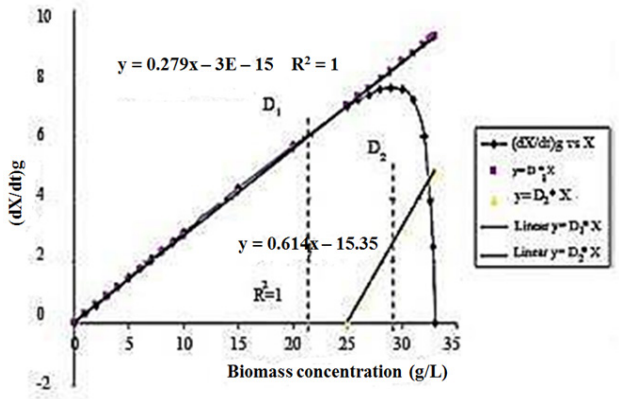


Figure 3: Graphical method to determine number of reactors

$$\left(\frac{dX}{dt}\right)_G = \frac{\mu_{max}*\left(S_o-\left(\frac{X}{Y_{X/S}}\right)\right)}{K_s+\left(S_o-\frac{X}{Y_{X/S}}\right)} * X$$

As it can be seen $D*X$ curve intersects the main curve at X_1 = 25 *g/L*. For the second line, X_2 at the intersection with the main curve is X_2 = 32.5 *g/L*. According to the graph, two reactors will be sufficient to produce almost maximum number of cells, therefore, two reactors; one aerobic and one anaerobic will be included in a system.

For the anaerobic fermentation tank, yield coefficient ($Y_{p/x}$) is determined to be $9.2\ g.\frac{ethanol}{g}.cell$ (Eq.7).

$$Y_{P/X} = \frac{Y_{P/S}}{Y_{X/S}}$$

The final product concentration is determined as 69 *g/L* by using Equation 8.

$$P_2 = Y_{P/X} \cdot (X_2 - X_1)$$

From the mass balance around second tank (anaerobic), μ_2, is calculated by Equation 9 and is determined as 0.142 h^{-1}.

$$\mu_2 = D_2 \cdot \left(1 - \frac{X_1}{X_2}\right)$$

Knowing μ_2, S_2 is found to be 8.34 *g/L* by Equation 10.

$$S_2 = S_1 + \alpha \cdot S_0 - \frac{\mu_2 \cdot X_2}{D_2 \cdot Y_{X/S}}$$

The time of exponential growth phase in the batch fermenter is determined to be 15 *h* (Eq. 11).

$$t_{exp} = \frac{1}{\mu_{max}} \cdot ln\left(\frac{X_m}{X_0}\right)$$

Since the working volume of the batch fermenter is considered to be 1% of the continuous fermenter, the working volume for the batch fermenter is found to be 0.062 m^3 (Eq. 12).

$$V_{BW} = V_{CW} \cdot \frac{1}{100}$$

Assuming the head-space as 15%, the total volume of the batch fermenter is found to be 0.0713 m^3 (Eq 13), with a diameter of 0.356 *m* and a height of 0.712 *m*.

$$V_T = 3.14 \cdot \frac{d_B^3}{2}$$

Sterilization time for batch fermenter. The probability of an unsuccessful sterilization is given in Equation 14.

$$1 - P_{0(t)} = [1 - 1(1 - P(t))^{N_0}] = 1 - (1 - e^{-kt*t})^{N_0}$$

Assuming 1-P_o(t), k_d, n_o, N_o ,kd are 10^{-3}, 1 min^{-1}, 105 *cells/L* and 274,000,000 cells respectively, sterilization time needed is found to be 21 *min* based on the sterilization chart [30].

Aeration and agitation. Aeration is vital for aerobic fermentation and is used to meet the oxygen demand of yeast during fermentation. The oxygen uptake rate (OUR), volumetric oxygen transfer coefficient (kLa), superficial gas exit speed, total orifice area, and number of nozzles required are determined in this section. Agitation, which is needed to provide uniform (homogeneous) mixture in the fermenter, is also discussed here. Agitation increases the oxygen transfer rate (OTR) by breaking the air into finer bubbles, increasing the surface area (higher interface area). In this section, impeller diameter, impeller speed, and power requirement are determined for both batch and continuous fermenters.

Oxygen uptake rate (OUR). Dissolved oxygen (DO) is an important parameter in aerobic fermentation, and may be a limiting substrate toward microbes. Sparging air through the fermentation broth can introduce oxygen into the system. Oxygen uptake rate is defined as the oxygen consumption rate of microorganisms to grow. Oxygen uptake rate is represented below where q_(O_2) is the specific rate of oxygen consumption (mg O_2/g dw cells·h), and X is the cell concentration (g dw cells/L). In this study, the supply of oxygen should meet the maximum requirement when microorganisms reach the maximum biomass concentration (Xm). Then, the critical value for oxygen supply becomes equal to 1.04 g O2/L h.Oxygen uptake rate (OUR) equation is given in Equation 15.

OUR=q_(O_2)*X

Maximum biomass concentration obtained in the continuous aeration reactor is 25 g/L. Assuming DO of 30% for biomass production bioreactor, OUR is calculated using Equations 16 and 17 [24].

q_(O_2) q_ =8 (mMO_2)/g d_w h

OUR=(8 (mMO_2)/(g·d_w·h))*(25 (gd_w)/L)*(32 g)/(1000 mM)=6.4 (gO_2)/(L·h)

Standard saturated oxygen concentration. The total amount of chlorine added to selective medium is 6.8 g/L, which gives an equilibrium DO level (CL*) value around 7.06 mg O2/L. When DO is 30%, OUR is equal to 6.4 g O_2/Lh. KnowingC_L^*, CL is determined as 2.12 mg/L based on Equation 18.

C_L=0.3*C_L^*

At steady state, OUR equals to oxygen transfer rate (OTR) as shown in Equation 19. The volumetric oxygen transfer coefficient (KLa) is found to be 0.344 s-1.

OUR=OTR=6.4 (gO_2)/(L·h)=(C_L^*-C_L)*K_L a

Impeller diameter. The impeller diameter is assumed to be 20% of the reactor diameter. Therefore, diameters of the impellers for batch (dib)and continuous (dic) fermenters are 0.26 and 1.22 m respectfully.

Reynolds number of the impeller is given in Equation 20.

$$Re_i = \frac{\rho * N * (D_i)^2}{\mu}$$

From Reynolds number, impeller speed can be derived as given in Equation 21.

$$N_i = \frac{Re_i * \mu}{\rho * D_i^2}$$

For the batch fermenter, assuming Reynold's number is 1000, the speed of the impeller in the batch fermenter (Nib) is determined as 148 rpm. For the continuous fermenter, Reynolds number is assumed to be 10,000, thus the speed of the impellers in the continuous fermenters (Nic) is found to be 67.2 rpm.

Power requirement. Since the viscosity is between 100 and 1000 cp (200 cp), and considering cost-efficiency, the type of impeller is chosen to be propeller [28].

The impeller power number equation is given as below (Eq. 22):

p_no=P_g/(ρ*N^3*D_i^5)

According to the chart of 'impeller power number vs. impeller Reynolds number for different impeller types [28]; Pno for marine impeller at Rei=1000 is 0.5, and at Rei=10,000 is 0.4. Then, for batch pre-fermenter; impeller power is found to be 10.74 W (0.014 hp) (Eq. 23).

P_gb=P_no*ρ*N^3*D_i^5

For continuous aerobic fermenters; impeller power is found to be 1823 W (2.44 hp) (Eq. 24).

$$P_{gc} = P_{no} * \rho * N^3 * D_i^5 = 0.41200 * 1.12^3 * 1.22^5$$

Superficial gas exit speed. Assume k (constant) equal to 1; for batch pre-fermenter, superficial gas exit speed (Vsb) is 20 mm/s while superficial gas exit speed for continuous aerobic fermenter (Vsc) is 30 mm/s (Eq. 25).

$$V_{sb} and\ V_{sc} = \left[\frac{K_L a}{k * \left(\frac{P_g}{V_{pf}}\right)^{0.4} * (N)^{0.5}}\right]^2$$

Total orifice area. Total orifice area is given as follows (Eq. 26);

Total Orifice Area= Q/Velocity

Assuming 0.01 vvm (29); total orifice area for the batch pre-fermenter is calculated to be 0.03 m2. For continuous aerobic fermenter; total orifice area is found to be 2 m2.

Number of orifices. Number of nozzles can be calculated using Equation 27.

Number of Nozzles=(Total Area)/(Area of Nozzle)

For the batch pre-fermenter, assuming the diameter of one nozzle is 0.5 cm, the area of one nozzle is calculated to be 0.0000196 m2 and number of nozzles for the batch fermenter is determined to be 1,531.

For continuous aerobic fermenter, assuming the diameter of one nozzle is 5 cm, the area of one nozzle is found to be 0.00196 m2.The number of nozzles for the continuous fermenter is determined to be 1,020 using Equation 27.

Downstream process and recovery. On the completion of ethanol fermentation, the medium broth contains the desired product, biomass, and varying amounts of other impurities, e.g. mineral salts, organic acids, and second metabolites from cells, etc. In the continuous submerged fermentation, the biomass is difficult to be removed from fermentation broth due to the viscous nature of the sweet potato mash. In the first step, fermented broth (also called beer) is processed through a fermented mash column where steam is used to strip off almost all of the ethanol, along with some water, from the slurry. The required area of this separation column is calculated to determine the size of continuous filter. The assumptions for calculating the filtration are φ=0.5, n=100 rpm, r_m=1000 m-1, ∝=10 m/kg, and ΔP=0.0025 N/m2.

The continuous rotatory mash column has to filter 87,000 liters fermentation broth with final biomass concentration, c=32.8 g/l, within one hour in order to meet the whole fermentation system from the Ruth equation (Eq. 28).

$$\left(\frac{V}{n}\right)^2 + 2V_0\left(\frac{V}{n}\right) = K\frac{\phi}{n}$$

The filter contact area (A) is calculated to be 9.18 m2 and the total area of the filter is 18.36 m2 by using Equations 29-30, as follows:

$$V_0 = \frac{r_m}{\alpha c} A$$

$$K = \left(\frac{2A^2}{\propto c\mu}\right)\Delta Pg$$

The slurry part of fermentation broth will lower the filtration efficiency. Therefore, one filter aid needs to be used to accompany with the filtration process.

After removing the biomass and impurities in the filtration process, continuous-feed distillation is carried out where ethanol is separated from water base on the differences in boiling. The product stream that is ready to leave the distillation columns contains about 95% ethanol by volume (190-proof). The ethanol is further dehydrated by using the molecular sieve system, in where water is trapped and adsorbed in the microporous beads while larger ethanol molecules flow around them via a sieving action [30]. The final product from the molecular sieve system is 99.6% pure ethanol vapor, which will be further condensed and mixed with denaturant (e.g. gasoline) and stored in tanks before being transported for sale as a motor fuel additive [30].

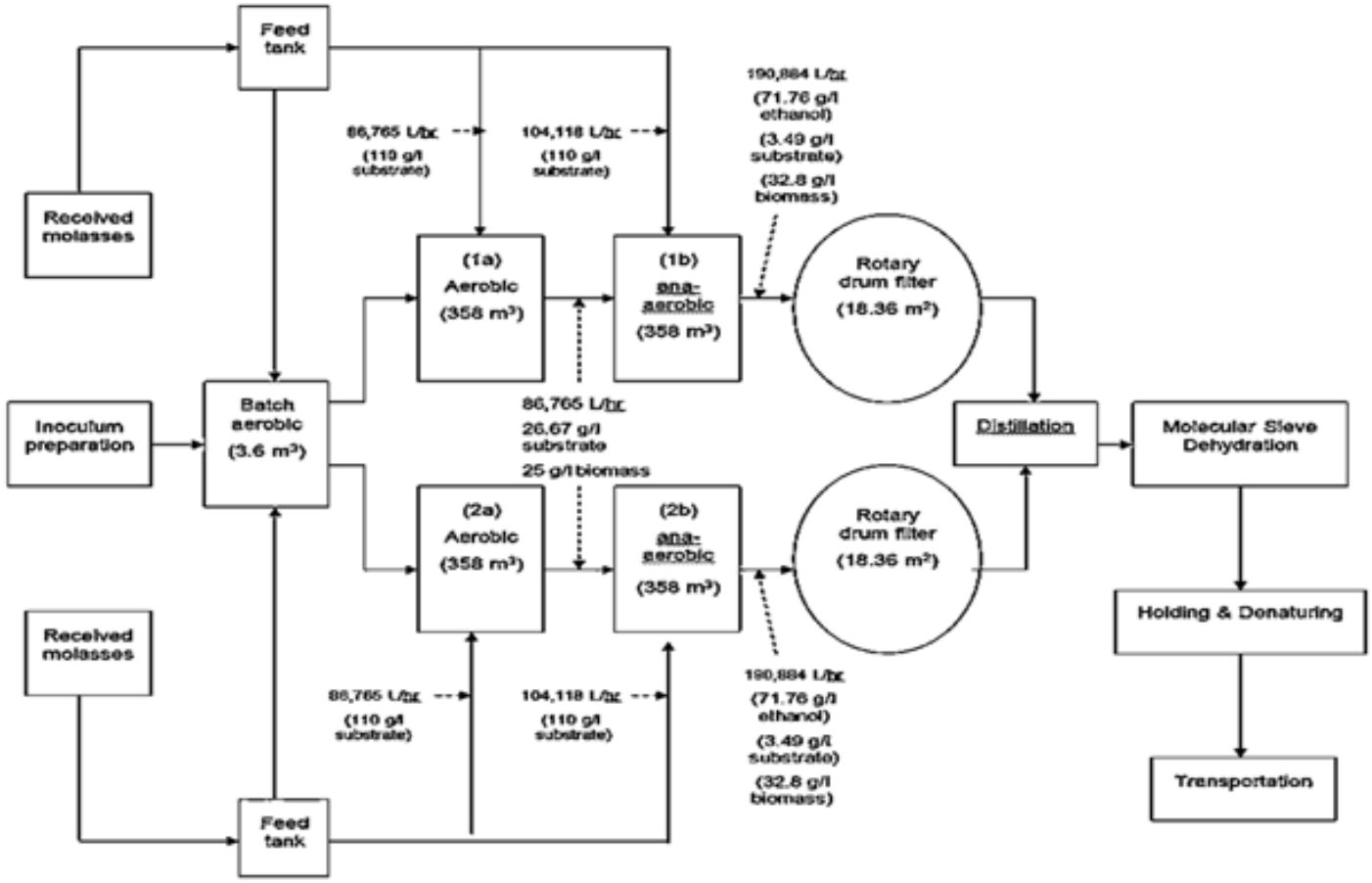


Figure 4: Schematic diagram of the overall system

Results and discussion

The fermentation part of the plant comprises one batch (aerobic) reactor and two parallel systems each composed of two continuous (one aerobic and one anaerobic) reactors to receive sweet potato waste from two feed tanks that dilute and standardize the sweet potato mash. The schematic diagram of the overall system is given in Figure 4

The batch fermenter (0.0713 m3) was designed for the initial production of biomass. The tank was sterilized for 21 min and the impeller with a diameter of 0.26 m rotates at 148 rpm with 10.74 W (0.014 hp) power. In total, 1531 nozzles sparge air from the bottom of the tank. At the end of the fermentation, the ingredients with the biomass product (45 g/L) are pumped into the continuous aerobic fermenter.

The continuous aerobic fermenter (8 m3) was designed for the further accumulation of biomass. The biomass concentration of this tank is increased to 25 g/L, while about 76% of the substrate is consumed. The Oxygen uptake rate (OUR) is 6.4 g O_2/Lh with the volumetric oxygen transfer coefficient (kLa) of 0.344 s-1. This reactor had an impeller with a diameter of 1.22m that rotates at 67.2 rpm, with 1,823 W (2.44 hp) power. A total of 1,020 nozzles sparge air from the bottom of the tank.

The continuous anaerobic fermenter (8 m3), which receives a stream from the continuous aerobic fermenter, was designed for the production of ethanol. Since the substrate concentration coming from the aerobic fermenter is quite low (26.67 g/L), a stream of fresh medium, which has 1.2 times higher flow rate than the stream coming from the aerobic fermenter, is pumped to this reactor from the feed tank. In this reactor, the final product (ethanol) concentration of 71.76 g/L is achieved. Also, the final biomass and substrate concentrations are 32.8 g/L and 3.49 g/L, respectively. This reactor has an impeller of same specifications as that of the continuous aerobic fermenter. In the downstream process, a rotary drum filter with a total filter area of 18.36 m2 was implemented.

The biggest issue with the bioreactor design was to incorporate the anaerobic and aerobic fermentation parameters in to the system. Basically, this problem arises from the fact that S. cerevisiae shows Pasteur Effect [37]. This microorganism has two different metabolisms under aerobic and anaerobic conditions.

Especially, in the section where the optimum number of reactors was determined, a number of growth and production parameters such as the yield factors were selected from the literature to be in the range of both anaerobic and aerobic fermentation to help the construction of the productivity curve.

Another difficulty was to make different assumptions for the parameter values in both aeration and agitation section. Since the diameters of the batch and continuous bioreactors are not

the same, different assumptions had to be made for each reactor. For instance, Reynolds number was selected to be higher for the continuous system to assure a reasonable agitation speed for both reactors. As another example, the diameter of the nozzle had to be assumed bigger for the continuous fermenter comparing to batch fermentation to assure a reasonable number of orifices.

Conclusions

In this conceptual study, bioethanol production from sweet potato waste by Saccharomyces cerevisiae was proposed based on references, data collection. An industrial-scale plant was designed to produce 5 million liters ethanol per year in 300 working days , which accounts for about 0.5% of the total ethanol demand Taiwan , producing clean and sustainable energy from agricultural waste abundantly available locally. For future studies, we recommend that purification steps be focused on for the complete design of the plant. In addition, a cost analysis would be beneficial to determine the economic feasibility of the plant. It is anticipated that this conceptual design provides useful information for bioethanol producers.

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