UNIVERSITY OF INDONESIA

TECHNO-ECONOMIC ANALYSIS OF LARGE-SCALE PRODUCTION OF BIOETHANOL FROM MICROALGAE

THESIS

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ENGINEERING FACULTY UNIVERSITY OF INDONESIA
BACHELOR PROGRAMME
DEPOK
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TECHNO-ECONOMIC ANALYSIS OF LARGE-SCALE PRODUCTION OF BIOETHANOL FROM MICROALGAE

TITLE PAGE

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Prepared as one of the requirement to obtain the title Sarjana Teknik Kimia

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STATEMENT OF ORIGINALITY

This thesis is a result of my own work,

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PREFACE

I would like to express my gratitude to Jesus Christ Lord almighty for His great help to this thesis as it could be done on time. The thesis with title **Techno-economic Analysis of Large-scale Production of Bioethanol from Microalgae** was accomplished as part of academic requirement to achieve the Sarjana Teknik degree in Chemical Engineering Department FTUI.

During the preparation of this thesis the writer would like to acknowledge all the help that has been given the writer during the preparation of this thesis. Therefore the writer would like to send the deepest gratitude to:

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The writer realize that this thesis is far from perfect and therefore would kindly accept any critics or suggestion to improve the writing for the future.

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STATEMENT PAGE OF AGREEMENT OF THESIS PUBLICATION FOR ACADEMIC INTEREST

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ABSTRACT

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Presently, substantial progress has been made in advancing biofuel production to meets global energy demands and the adverse effects high fuel prices. However, food-derived bioethanol feedstocks have aroused social and environmental concerns. *Chlorococcum sp.*, a microalgae strain with high carbohydrate content for fermentation feedstock, is a potential biomass for bioethanol production. This study examines technical and economical feasibility of the production, which capitalise annual biomass of 50,000 tonnes over 10 years operating time. This study explores different technologies configuration at various production stages, where chosen technologies are mainly cost-effective, energy saving, and reliable for large-scale operation. With biomass cultivation in raceway pond, dual-stage flocculation preceding centrifugation dewatering, dilute acid pre-treatment, separate hydrolysis and fermentation, and purification, the overall production cost incurs at AU$ 33 per litre bioethanol produced. The overall finding indicates that the project is technologically feasible, but not economically. Improving cultivation and dewatering can further reduce production cost hence the economic of microalgal bioethanol becomes more competitive and attractive.

Keywords : Techno-economic, Bioethanol Production, Microalgae

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CHAPTER 1
INTRODUCTION

1.1 Background

As the scarcity of fossil fuel becomes more apparent, researchers feel compelled to reduce the dependency on these limited resources with alternative fuels which are sustainable, environmental friendly and cost effective. Bioethanol, a viable option to the aforementioned problem, has become a centre of attraction and a highly demanded by countries in the world. Itwas reportedto burn cleanly especially as the amount of gasoline with which it is blended decreases [1]. It also provides higher octane rating without the expense of engine modification [2].

Bioethanol is a combustible fuel produced from renewable biomass. Currently, there are three common groups, which serve as feedstock for the production of bioethanol: sucrose based (sugar cane, sugar beets, fruits and sweet sorghum), starchy based (corn, wheat, rice and barley) and lignocellulosic biomass (wood, straw, grasses). However, issues concerning social, economic and environmental soon surface when using these feedstock.

The issues regarding the use of sucrose based and starchy based biomasses are not uncalled for. These edible crops are direct sources of food, the fact that plantation crops can be divided into bioethanol crops and multipurpose crops do not ease the situation as the amount of land used raises concern [2]. In addition, the increase in the price of food in global market and the cost associated in growing these feedstocks have rendered them less attractive. Similarly, lignocellulosic biomass albeit cheaper and in abundance, requires higher production costs due to the need to remove lignin during pre-treatment [2].

These concerns have highlighted the need for an alternative biomass and thus, led to a growing interest in microalgae for bioethanol production. Recently, microalgal biomass was found to have the potential to develop into an important fermentation feedstock due to its many advantages. Microalgae has quick growth rates and short harvesting times. The structural compositions of most microalgae

species display high carbohydrate content, which is beneficial for the production of bioethanol. In relation to the environment, it has the ability to capture carbon dioxide and other greenhouse gases during photosynthesis. The plus point extends further for some microalgae species which grow in seawater and wastewater stream, it not only reduces the reliance on fresh water sources, but at the same time treats these contaminated water systems [3]. Also, microalgae as bioethanol biomass is independent and does not compete with agricultural production for food feedstock. Due to these reasons, microalgae is very appealing asanalternative solution in reducing consumption of fossil fuel in the future.

During past years, the amount of works observingmethods to reveal the potential of microalgae in releasing the reducing sugar and converting it to ethanol have increased. However, studies that have been done so far are limited laboratory scale. A large scale of an industrial production of bioethanol from mircoalgal is what this study is called for.

The key point to investigate the feasibility of proposed project of large-scale microalgal bioethanol production can be accomplished based on the economic implications from adopted technologies scheme. Meanwhile, theconversion of microalgal culture to bioethanol consists of few stages;cultivation and dewatering for biomass production, pre-treatment, hydrolysis, fermentation, and purification. Following from the technology configuration, an economic model will be developed to show the efficiency and effectiveness of its economic performance.

1.2 Problem statement

It is indisputable that bioethanol production from biomass is a revolutionary discovery and a key solution to fossil fuel depletion. However, many remain sceptical on the utilization of current agricultural feedstock in the process and thus evoke a “fuel versus food” debate. Microalgae, a renewable resource with rapid growth, is not an agricultural demand. Microalgae biomass has the ability to overcome related issues to be an effective alternative bioethanol feedstock.

Although only limited studies have been done in investigating the potential of microalgal biomass in laboratory scale, the opportunity in broadening it to a large

scale of industrial bioethanol production is very likely. Nevertheless, there is numerous information that can be extracted from the existing well-developed technologies from others biomass for bioethanol production. Thus, results and knowledge from former studies are sufficient in supporting the data required for the technology development.

Furthermore, the prospect of production is not only determined by the technology chosen, but also the economic performance. Economic analysis on net production cost and profit analysis hold the key role on viability of the production. Therefore, this study will examine the feasibility of bioethanol production from microalgae biomass in its technology and economic.

1.3 Project Aim

- To develop the most applicable and cost effective process technology for microalgal bioethanol manufacture on an industrial scale basis, and estimate the net production cost.
- To analyse the economic viability of microalgal bioethanol production based on profitability analysis and sensitivity analysis.

1.4 Scope

The study is mainly concentrated on technology and economic of microalgal bioethanol production in large scale. To master the technologies involved on the bioethanol production, and intensive learningon the literature regarding the methods that has been readily used is essential. The production process is broken down into 6 stages, namely cultivation and dewatering (biomass production), pre-treatment, enzyme hydrolysis and fermentation, and product purification.

- **Biomass production**

Cultivation and dewatering.Davidson et al. [32] have done study regarding the biomass production previously. Results will be referenced in this study where is appropriate. For cultivation, there are 3 methods will be study out for comparison; raceway pond (open pond), horizontal tubular photobioreator, and external loop photobioreactor. While dewatering, from the technology that has been assessed,

one applicable technology will be chosen. Although drying and lipid extraction are part of the biomass production, these process will not be assessed in this work. However, a scenario with lipid extraction included in the production process will later bediscussed briefly. The estimation the lipid extraction design is also referred from works done Davidson et al.

- **Bioethanol production**

Pretreatment. Among all the pre-treatment method available, only one technology will be examined.

Hydrolysis and fermentation. These two different stages are often combined or conducted separately. In this study, both simultaneous and separate conduction will be assessed in order to analyse the difference.

- **Bioethanol purification**

Bioethanol purification will not be implemented in this study. However, the assessment regarding to process will still be included. Applying work done by Aden et al. [33], the data produced will be adjusted according to the final outcomes from the upstream.

- **Economic Analysis**

Based on different scheme of process technologies, its economic will be evaluated. Parameters that will be involved on the economic analysis will be the total of capital investment, annual fixed capital, variable and fixed operating cost. Based on these, the total production cost, cash flow analysis, and the sensitivity analysis will be assessed to show the credibility of bioethanol production on the selected technology. An additional scenario will also be assessed with considering lipid extraction as part of the process technology in order to compare the economic performance with and without the regarded process.

CHAPTER 2
LITERATURE REVIEW

The conversion of biomass to bioethanol involves a number of methods depending on the composition of the feedstock. In general, the process is in the following sequence: pre-treatment to breakdown the cell walls, hydrolysis of biomass to form simple sugars, fermentation by yeast or bacteria to ethanol and finally product recovery. A cellulosic material to bioethanol featuring enzymatic hydrolysis is shown in the schematic diagram below.

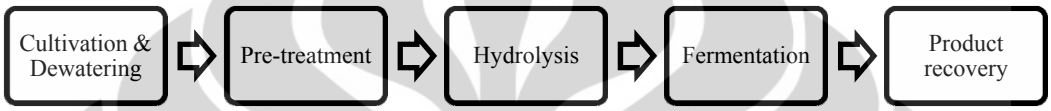


Figure 2.1. Schematic representation of biomass to bioethanol conversion process [1].

2.1 Conditions for microalgal biomass production

Microalgae can be categorized into photoautotrophic, heterotrophic, mixotrophic, and photoheterotrophic organism, depend on their metabolism. A photoautotrophic algae requires sunlight as energy source to produce organic compound from inorganic matter during photosynthesis, whereas heterotrophic algae assimilate organic compound as carbon and energy source [4]. Mixotrophic organism, a hybrid of autotroph and heterotroph, alternates both energy and carbon source; depending on its availability. On the other hand, photoheterotrophic algae demands light as energy, but can only synthesis organic compound from environment as carbon source [5].

Currently, the technologies developed for microalgal biomass production mainly focuses on photoautotroph microalgae. The photosynthesis reaction is generated based on the following reaction [6],

$$CO_2 + \text{light} + H_2O + \text{nutrients} \rightarrow \text{biomass} + O_2 \qquad \text{(Rxn. 2.1)}$$

Light. Light is the limiting factor for photoautotroph organism as it is the sole energy source to utilise carbon dioxide into organic compound. The intensity of light greatly affect culture growth as it aids to attain optimum growth rate [7]. However, light supply beyond its limit can damages light receptor of organism; hence inhibit culture growth.

Temperature. Temperature is another essential element in biomass production. When temperature increases, the growth rate rises exponentially to its optimum point [7]. Some microalgae can endure temperature at 15°C lower than its optimal, but slight increase beyond optimum temperature can drastically decrease growth rate [5]. Moreover, roughly 25% of biomass produced in daylight is lost during night. This is not only caused by respiration, but also temperature variety [8].

Gas exchange. Roughly 50% of microalgae structure is made of carbon, which arrives from carbon dioxide supply [7, 8]. Microalgae capture CO_2 from atmosphere, industrial gases (flue gases), and soluble form of carbonates (Na_2CO_3 and $NaHCO_3$)[4, 6] Most microalgae can tolerate extremely high level of CO_2 from power plant flue gas (approximately 150000 ppmv), while low CO_2 from air can inhibit the grow [4]. In open culture, CO_2 is delivered through sump integrated with pH sensors to indicate extreme CO_2 losses at low pH [7, 8], whereas CO_2 is easily fed through air injection in close pond.

Nutrients. Essential inorganic element such as nitrogen, phosphorous, iron and silicon should be supplied sufficiently, and sometime excessively. Most of microalgae are better in absorbing soluble form of nitrogen rather than take it from air [4]. Singh [7] stated that the ratio of nitrogen and phosphorus for nutrient requirement is roughly 16:1. This shows that although only small amount of phosphorus is needed, the presence of it is important to support the growth cycle [4].

Contamination. Contaminant from unwanted organism can endanger culture life in the pond. Open pond is highly susceptible to contamination as it leads to competition of substrate between the strain and organism, cell destruction if contaminants are toxic, and biomass yield reduction. In some algal cells, severe cell growth due to contamination can be avoided by applying extreme condition [4], such as high saline environment for *D. Saline*, high alkalinity level for *Spirulina*, and rich nutrient supply for *Chlorella* [7, 9].

***Mixing*.** Proper mixing avoids sedimentation and ensures that algae strains are suspended. This helps to equalise gas and light distribution throughout the strains, especially in large-scale production.

2.2 Cultivation

Cultivation is essential at the early stage for biomass production yield. A desirable cultivation system should constitute a balanced environment for microalgae cell growth. Lacking on controls of any of the parametersaffects the quality, quantity, and cost efficiency of microalgae bioethanol production. Two competing cultivation alternatives that are well known for biomass production are open pond and closed pond.

2.2.1 Open Pond

Microalgae cultivation in an open pond, as its name implies, is basically carried out in an open area either in nature ambiance (lakes or ponds) or in an artificial environment with satisfied water resource [4]. Choice of open pond available in research and industrial scale are raceway pond, unmixed open pond, circular pond and thin layer-inclined pond [7, 10]. Raceway pond (figure 2.2) is an artificial system commonly used to cultivate microalgae. It has a shallow, closed loop recirculation channel that is commonly made from concrete and allows lighting throughout the entire culture [4, 6]. The feeding point is equipped with paddlewheel that enhances circulation and mixing of the feed and nutrient within the pond. Excellent mixing and circulation are important not only to establish stable growth and production of microalgae, but also to prevent sedimentation [4].

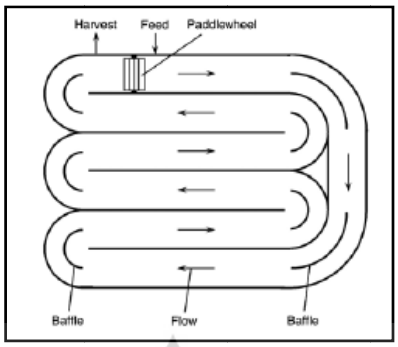


Figure 2.2. Raceway pond system[1]

The raceway pond is relatively easy for operation since it has a large open access area for cleaning and maintenance. The simplicity in raceway pond configuration not only eliminates huge capital cost for equipment installation and operation cost, but also reduces energy consumption.

Raceway pond, on the other hand, is largely dependent on its surrounding environment. This indicates that a lack of control in light intensity and temperature, and limitation of CO_2 supply in air would inhibit mass transfer for cell growth [5]. Furthermore, large open-air pond makes it highly susceptible to undesired contaminants and water losses caused by evaporation [9].

2.2.2 Photobioreactor

Photobioreactor (PBR) is an effective closed cultivation system for both industrial and experimental scale. It optimizes cell culture and nutrients circulation within a narrow closed vehicle to improve photosynthesis mechanism. The closed environment brings great control in maintaining growing parameters and improving biomass volumetric productivity by approximately 15 times as compared to raceway pond [10]. The two most common photobioreactor designs are tubular and flat photobioreactors as discussed below.

Tubular photobioreactor. Tubular photobioreactor shown in figure 2.3 can be implemented for outdoor broth culture. It consists of an array of transparent straight tubes made from plastic or glass with diameter up to 0.1meter to ensure penetration of sunlight through the culture, especially in dense culture broth [6, 8].

The configuration of solar tubing can be arranged in parallel, straight parallel horizontal (figure 2.4) or external loop (figure 2.5)[4]. For the external loop array maximise the collection of solar radiation and minimise the occupied area of land. Vertical air lift tubular permit large mass transfer with lower energy utilization whereas horizontal tubular is scalable but is a waste of area [7]. Photobioreactor configuration allows the broth to be circulated continuously from the degassing chamber to the solar tubes through mechanical or airlift pump [8]. This not only enhances O_2 and CO_2 exchange but also promotes homogenation of broth mixture through high turbulence.

Flat-plate- Flat plate is another favourable PBR, which has large surface area for illumination made from transparent material such as PVC or glass [6, 10]. It is low in energy requirement and oxygen accumulation and high in mass transfer efficiency and photosynthesis activity.

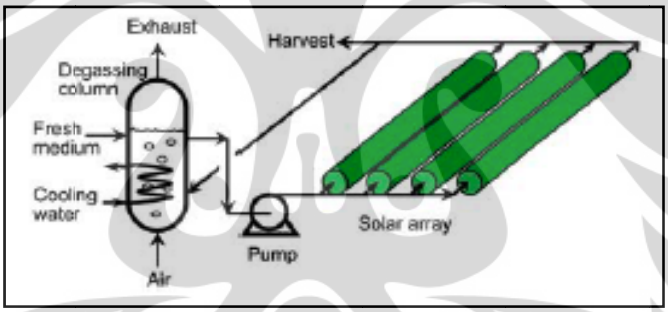


Figure 2.3.Tubular photobioreactor with parallel horizontal tubes [1].

In general, PBR overcome most of the limitation of open pond. It minimises the CO_2 losses, evaporation, and able to achieve high culture density. Homogenised broth is well-achieved by aiding airlift pump. Moreover, PBR is highly favourable for a sensitive strain as it limits contamination due to gas exchange [6]. However, culture may encounter shear damage in result of high turbulence by mechanical and bubble mixing. Dissolve oxygen concentration generated in the tube can also inhibit photosynthesis if the level is greater than the saturated air, and develop photooxydative damage to cells if it is exposed to intense sunlight [8]. Overheating, bio-fouling, and impracticable maintenance and cleaning in closed system are the main drawbacks of PBR [5]. In relation to cost, PBR requires high

installation, material, and operating costs as it utilises complex mechanical machinery, along with the higher energy consumption, contributes to the increase in production cost. Therefore, the high capital and operating cost has compensated for the greater volumetric biomass productivity [5].

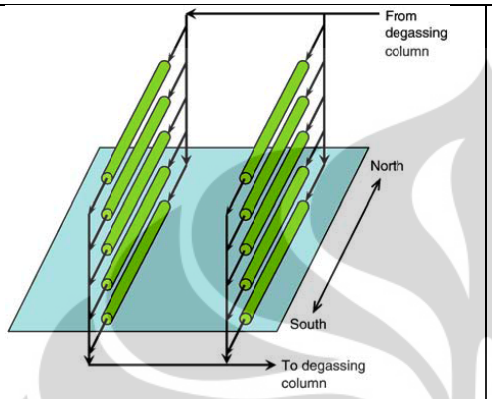


Figure 2.4. Horizontal straight parallel solar array [4].

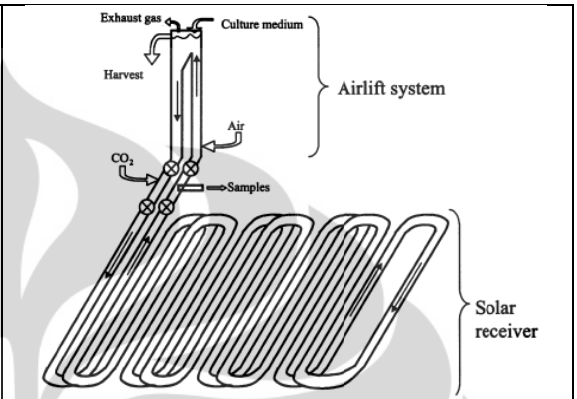


Figure 2.5. External loop photobioreactor [34]

2.3 Dewatering

Dewatering, otherwise known as harvesting is followed after cultivation to separate microalgae from the culture media. This can be achieved mainly by filtration, chemical flocculation, and centrifugation, practiced either individually or in combination [11]. However, these unit operations are often tedious and expensive as algal cells are not only small in size but have low specific gravity, which hinders separation from the bulk liquid. According to Gong and Jiang [12], this stage incurs approximately 20-30% of the total cost which explains the importance of selecting a suitable approach which is both cost effective and high recovery efficient.

2.3.1 Filtration

Filtration is used to harvest microalgal biomass from photobioreactors, where microalgal cells attain higher density. Conventional filtrations include rotary drum pre-coat filtration, press filters and tangential flow filtration (TFF) [13]. Filter

presses operating under pressure or vacuum can be used to recover large quantities of biomass, but may be relatively slow for some applications. In such cases, alternative modes of filtration can be applied; namely membrane microfiltration and ultra-filtration depending on the size of the algal cells. The broth containing algae enters the filter in which the algae accumulates and allows the medium to flow through. The process continues until the micro filters are saturated with a thick algae paste [10].

The size of the algal cells is an important parameter in the separation as it directly affects the filterability and the setting rate of the particle [14]. Therefore, the separation of large microalgae such as *Coelastrum proboscideum* and *S. platensis* is more suited using this method as compared to *Scenedesmus*, *Dunaliella* or *Chlorella*, which are of smaller dimensions [5]. Filtration has proved to be feasible at a laboratory scale; however, it is limited to recover small quantities of algal biomass only[12]. This is because it suffers from problems such as membrane clogging and formation of compressible filter cakes when applied to large scale use [9]. In addition, there is still considerable scope to lower the cost of maintenance arising from membrane replacement and pumping which renders this method less ideal in this study.

2.3.2 Flocculation

Most microalgal cells have a size range of 5 to 50μm and form stable suspensions with negatively charged cellular surfaces[15]. Flocculation, an initial dewatering/pre-treatment process is therefore required to allow the cells to form larger clumps, which then promote or enhance subsequent dewatering methods. Flocculants are chemicals added to counter the surface charge on algae without affecting the composition and toxicity of the product [10]. The stability of the suspension is influenced by the interactive forces between the cells and water and cells. The mechanisms of flocculation, namely charge neutralisation and polymer bridging have brought forth the use of two types of flocculants, metal salts and cationic polymers. Most microalgal systems rely on cheap chemical flocculants

and when operate under optimum condition, make this technology more favourable and economically viable as compared to other dewatering options.
Multivalent salts such as ferric chloride ($FeCl_3$), aluminium sulphate ($Al_2(SO_4)_3$) and ferric sulphate ($Fe(SO_4)_3$) are commonly used and its effectiveness vary according to their charge density. Alum is found to be an effective flocculant for *Scenedesmus* and *Chlorella*[12]. However, the algal chemical sludge may not be suitable for certain downstream applications, such as animal feed supply or for anaerobic digestion [9].

Organic cationic polymeris also used extensively for recovering microalgae cultures and is the preferred option when microalgae residueis used in the downstream process. It is less sensitive to pH and requires lower dosage for flocculation process[10]. Bridging of the cells occur when segments of the polymer chain adsorb on different particles and help microalgae aggregate. Molecular weight, charge density, dosage, pH are of the many factors, which affects the performance of the flocculants. Increased molecular weight and charge on the polymers are found to increase their binding capabilities [15]. Overdose of polymer, on the other hand, may cause settling or clarification problems. According to Sukenik et al.[16], the chemical condition of marine water (with salinity up to 36g/L) imposes problems when using polymeric flocculants during flocculation.The use of chemical flocculent such as “Chitosan” (cationic polymer) has also been studied but was found to be too expensive for economic algae dewatering.[10]

2.3.3 Centrifugation

Centrifugation is currently the preferred method to harvest the algal cells from photobioreactors[12]. It involves separating the algal growth medium into regions of different densities by means of centripetal acceleration. The algae can then be removed from the culture by draining the excess medium. The supernatant separated from the medium can also be used to identify the biomass composition from the medium [17]. The settling characteristics of microalgal cells and residence time of slurry in the centrifuge can affect the recovery efficiency and

cost of centrifugation [12]. Therefore, even though centrifugal recovery is rapid, it is considered to be too costly and energy intensive as a primary harvesting technique in a large scale setting. The energy input alone for such scale has been estimated at 3000kWh/ton[9].

Centrifugal recovery is capable of concentrating any type of microorganisms and is a good secondary harvesting method to concentrate an initial slurry (10-20g/L) to an algal paste (100-200g/L)[9].

2.4 Pre-treatment

Early pre-treatment is normally carried out prior to hydrolysis to improve extraction of sugar from microalgae by breaking down the crystalline structure. In addition, it is an important step to improve enzymatic attack and penetration to cellulose and to further improve hydrolysis performance [18]. It would also solubilise and hydrolyse hemicellulose without any extend of degradation to other fermentable materials [19]. Polysaccharides, which are the main composition of carbohydrate in microalgae, are entrapped between cell walls and intercellular matrices [20]. The structure of cell wall itself is majorly composed of crystalline cellulose that is strong and has high resistance to hydrolysis [21]. Without pre-treatment, glucose conversion can be difficult and enzyme accessibility during hydrolysis is limited.

Pre-treatment contributes major economic cost of bioethanol production since its effectiveness of obtaining fermentable sugar extensively affects downstream steps [18].However, with the absence of lignin in microalgae, lignin conversion can be dismissed, hence reduces cost and shorten the treatment duration [13]. Due to this reason, a simplified single stage pre-treatment and hydrolysis can be established.

Below are the factors that need to be considered in minimising the production cost [18, 19, 22]:

- Pre-treatment should maximize fermentable sugar yield for hydrolysis with minimum carbohydrate losses and by-product or toxic formation that hinder hydrolysis.

- The method or chemical applied should be effective, inexpensive, and moderate in cost.
- The reduction of biomass size is not necessary; thereby milling or grinding that is energy-intensive and expensive can be negated.
- Utilizing high dry raw material helps to minimize energy consumption.
- Low heat and power demand and consumption during pre-treatment are important.

The application of pre-treatment can be carried out using physical, chemical, biological, thermal, and enzymatic method.

2.4.1 Physical Method

2.4.1.1 Mechanical comminution

This method involves mechanical machinery to shatter biomass into fine powder by applying a combination of chipping, grinding, and milling, to increase specific surface and reduce the degree of polymerization [23]. The resulting particle size normally alternates between 10-30 mm after chipping and 0.2-2 mm after milling [22].

Although mechanical comminution does not involve toxic chemical, the power requirement can be fairly high depending on the final size reduction [2]. Studies reported that particle size reduction does not necessarily improve hydrolysis and it still requires further processing; hence it is not an economically feasible option [2, 18].

2.4.1.2 Pyrolysis

Pyrolysis has been widely used for lignocellulosic pre-treatment. The material is treated at high temperature of above 300°C in order to decompose the cellulose into gaseous product and residue. Treatment under low temperature is not effective as it slows down the decomposition resulting in fewer volatile product, but this hindrance can be overcome by adding catalyst such as zinc chloride or sodium carbonate or by introducing oxygen during the process [22].

2.4.2 Chemical Method

2.4.2.1 Alkaline Hydrolysis

Alkaline pre-treatment involves the solvation and saponification process by immersing cell in alkaline solution (such as NaOH) and applying continuous heating [3]. Dilute NaOH pre-treatment is found to be effective for feedstock containing low levels of lignin; hence making this a plausible method for microalgae. Without any lignin removal, this treatment leads to an immediate destruction of cellulose crystal structure by causing the pores of the biomass to 'swell', leading to an increase in internal surface area, a decrease in the degree of polymerisation and crystallinity of cellulose, thereupon reducing the crosslinks between the hemicelluloses. Unfortunately, there are insufficient data to support this method to be economical in a commercial scale due to the lack of experimental research of alkaline hydrolysis on microalgae.

2.4.2.2 Acid Hydrolysis

Acid pre-treatment comes in two alternatives; concentrated and diluted acid, with sulphuric acid as the common acid employed and other acid, such as hydrochloric, nitric, and trifluoracetic. Generally, the acid must be recovered for the process to be economically feasible [22]. Upon which, a neutralization of pH is necessary for the downstream enzymatic hydrolysis. According to Nguyen et al.[24], 58% of glucose release and approximately 29wt% (g ethanol/g microalgae) ethanol yield are achieved after pre-treating *Chlamydomonas reinhardtii* biomass with 3% sulphuric acid at 110°C for 15-20min reaction time. Harun and Danquah also reported a bioethanol concentration of 7.20g/L can be achieved using 1% (v/v) of sulphuric acid at 140°C for 30minutes. Therefore, time, temperature, microalgae loading and acid concentration are the key parameters for process optimisation.

Concentrated acid treatment is less favourable, albeit a powerful agent for cellulose hydrolysis. The utilisation of strong acid not only increase the amount of hazardous, toxic and corrosive chemicals but also increases reactor corrosion and opposes environmental sustainability [18].

Dilute acid process, on the other hand, is an attractive alternative for microalgae pre-treatment due to its high sugar-release in rapid hydrolysis rate [25]. Currently, dilute acid pre-treatment can be carried out in a continuous-flow process under high temperature (greater than 160°C) or as a batch process under low temperature (less than 160°C) [22]. However, a recent study shows that acid pre-treatment temperature of approximately 140°C is most suitable, as higher temperature would distort the equilibrial formation of simple sugar moieties. The duration of the process is most optimal at shorter times, consequently presenting a positive impact on energy consumption [23]. Figure 2.6 shows macroscopic image of algae cell Chlorococum sp. before and after acid treatment with 1% (v/v) of sulfuric acid for 30 min at 140°C [2].

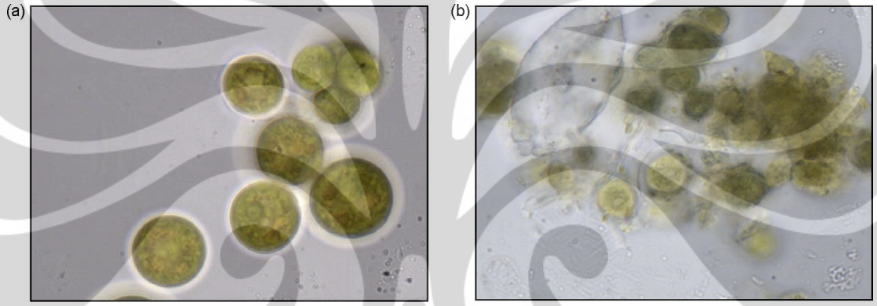


Figure 2.6. Magnified image of Chlorococum sp. before (a) and after (b) acid pre-treatment

It is undeniable that acid pre-treatment has a potential of improving bioethanol production levels from microalgal biomass [23]. It outweighs other pre-treatment methods in terms of cost effectiveness and energy consumption.

2.4.2.3 Biological method

The biological pre-treatment involves microorganism such as brown, white, soft-rot fungi for lignin and hemicellulose degradation to improve enzymatic hydrolysis. The method is fairly environmental friendly, as it requires no chemicals, modest environmental condition, and is low in capital cost and energy [18]. However, limitations occur as the pre-treatment rate is relatively slow for large production and some fermentable material is lost from microorganism activities during process [2, 19]. Furthermore, this method achieves high lignin and hemicellulose break up but low when it comes to cellulose. Since microalgae

is mainly cellulose and lignin free; therefore biological method might not be the best option. Galbe et al.[19]suggested this method could be applied as preliminary step, followed by other pre-treatment methods.

2.5 Hydrolysis

In the case of microalgal biomass, it is known that complex carbohydrates are entrapped in the microalgal cell wall and must be released and converted into simple sugars prior to fermentation. Hydrolysis can be categorised into those that uses chemicals (a form of pre-treatment mentioned in Section 2.3) and those that uses cellulose enzymes [1]. Even though the use of chemical acid such as sulphuric acid in hydrolysis process is more technologically mature, alkaline and enzymatic hydrolysis are comparable in performance.

2.5.1 Enzymatic Hydrolysis

The choice of enzyme used in hydrolysis is based on cellulose crystallinity, substrate surface area, cell wall thickness, porosity, and hemicelluloses or lignin contents [26]. Owing to the absence of lignin in microalgae, enzymatic hydrolysis can be carried out under mild operating conditions (pH usually 4.8 and temperature between 45-50°C) resulting in lower utility costs as compared to the aforementioned hydrolysis methods [22].

Cellulase and amylase are both existing hydrolysing enzymes that reduce cellulosic-based material and starch containing biomass respectively. The former enzyme not only has a higher cost (15-20cents/gallon bioethanol) compared to amylase which costs 2-4cent/gallon, it is also 100times slower and hence requires higher enzyme loading due to the complex nature of cellulose [2]. α-Amylase is used for starch liquefaction in hydrolysis. Cellulases, on the other hand, are highly specific enzymes, which does not affect other substrates other than cellulose. Therefore, the pre-treatment methods discussed in Section 2.4 are required to make yields more economical by breaking the hemicellulose structure [3]. Harun et al. [2] has also noted that combined acid pre-treatment and enzymatic pre-treatment, which is also well known as saccharification, of green algal

Chalmydomonas reinhardii has a slightly higher sugar liberation than single step enzyme hydrolysis[10].

Cellulases are usually a mixture of several enzymes, with at least three major groups acting on the cellulosic materials synergistically in the following order,

- Endoglucanases hydrolyse the non covalent interactions (β-1,4 bond) within the crystalline structure of cellulose molecule [26],
- Exoglacanases hydrolyse the individual cellulose fibers into simple sugars with the cellobiohydrolases attack the chain ends producing cellobiose [26], and
- β-glucosidases are only active on cello-oligosaccharides and cellobiose, release glucose monomers by converting cellubiose units [7, 27].

Glucose and cellubiose are the two major products from enzymatic hydrolysis. However, there are reports indicating the products act as inhibitors in some enzymatic hydrolysis reaction that uses cellulase prepared from *Trichoderma reesei*.

2.6 Fermentation

Upon hydrolysis completion, most monomer sugars have been liberated from microalgal biomass. The released fermentable sugar is subsequently converted into ethanol product by employing microorganism for an extent period of time in batch fermenter [10, 13]. Wide ranges of microorganism can be applied for bioethanol fermentation, such as bacteria, yeast, and fungi. Table 2.1 has summarised the advantages and disadvantages of some of the available microorganisms [13].

Depending on the process combination, fermentation can be approached through several options. When hydrolysis and fermentation are performed separately, the process is identified as separate hydrolysis and fermentation, hereafter termed as SHF. The combination of both processes, which is carried out subsequently in the same batch is referred as simultaneous saccharification and fermentation (SSF).

Comparing between these methods, SSF seems more attractive and favourable. An investigation by Saha et al.[28], regarding lime pre-treated wheat straw fermentation by *E.coli* has reported that SSF is capable to double the ethanol yield (around 0.56 g g^{-1} straw) compared to SHF at total hydrolysis and fermentation time of 144 hr. Similarly, study on steam pre-treated corn stover from Öhgren, K., et al. [29] also showed that application of SSF using *Saccharomyces cerevisae* gives 13 % higher ethanol yield than SHF.

Inhibitor built up caused by glucose decomposition in pre-treatment can be a hindrance to enzymatic hydrolysis and microorganism activity. Some alternatives, that help to overcome this issue, are by adapting fermentation organism in pre-treatment hydrolyzate [29] or detoxification after acid pre-treatment, prior fermentation [28]. Moreover, SSF has shown to withstand and reduce inhibitory effects upon fermentation, which improves the degree of ethanol productivity and concentration [29, 30]. In fact, Öhgren, K., et al. [29] has investigated that SSF from steam pre-treated corn stover gives higher overall ethanol in the presence of inhibitory effect from pre-treatment hydrolyze (72.4 %) than in its absence (64.1%), while SHF results the contrary. Nevertheless, SSF operating in the presence of inhibitor still attain higher yield compared to SHF without inhibitor [29].

Owing to rapid production of ethanol, SSF also eliminates contamination from undesired microorganism, as they are weak towards ethanol. As capital cost for contaminant purification is omitted, more profit is achieved with high quality ethanol product. Although there is no bioethanol production from microalgae that has been commercialized yet [10], studies have shown that microalgal bioethanol has lower energy requirement and simpler procedure compared to biodiesel production. Moreover, carbon dioxide generated from fermentation can be recycled back to cultivation process, which negates the greenhouse effect [3].

Table 2.1. Comparison of different fermentation microorganism

Micro-organism	Advantage	Disadvantage
Saccharomyces cerevisae (yeast)	High tolerance to inhibitors and osmotic pressure increase. Achieved bioethanol yield up to 18 % of fermentation broth.	Restricted to hexose sugar fermentation only, not to others simple sugar. Enzyme and yeast have different optimum temperature; hence optimum condition can be achieved when SSF is adopted.
Zymomonas mobilis (bacteria)	Higher bioethanol yield using alternative pathway for sugar fermentation. A non-biohazardous material. Demand simple nutrients.	Limitation to glucose, fructose, and sucrose fermentation only.
Escherichia coli (bacteria)	No complex growth medium required. Valid for wide ranges of sugar.	Narrow pH ranges during growth. Less hardy than yeast. Lack of supporting evidence regarding its residual. By-product; succinic acid, acetic acid [28]

2.7 Product Recovery

Purification of ethanol from the concentration produced by fermentation to concentrations useful as fuels is normally accomplished via distillation. Initially, the fermentation product is transferred to dehydration column to eliminate water carried away. Subsequently, Bioethanol is recovered from the impurities, such as dissolve CO2 and soluble material, in a stripping column upon which is concentrated in a rectifying column.

2.8 Economic Aspect

The major aspect that determines the successful of a project design is stated on its economic operation. The aimed of a project plant developed is to earn profits and

be able to operate well. Deducting total incomes with the expenses makes up net profits. However, during operation, there might be range of indirect cost that incurs beside direct cost. Thus, an estimation of the capital investment needed for the project establishment and the total production cost are vital as it illustrates the profitability of the project before it can be examined [39]. This section will discuss about the capital, the operating cost of the plant for cost estimation and methods to measure profitability of the project.

2.8.1 Plant Capital Investment

Before a production plant can be commenced, a certain amount of money is necessary to be invested for the project. This portion of money is ought to look after payment of some fix expenses for as land, equipment and instruments purchase and installation during the process initiation. Along process operation, a working capital cost is also required either during plant start up or end of process until the revenue is received. In other word, total capital investment consists of fixed capital investment and working capital.

2.8.1.1 Fixed capital investment

Fixed capital investment or FCI handles all expenses incurred as part of plant initiation. This cost is mainly subjected into direct and indirect fixed capital. Table 2.1 list the items involved in direct and indirect cost. Along with the table, variations of percentage on each item are shown. This indicates the proportion of the items contributes to the fixed capital investment.

Table 2.1 List fixed capital investment items [14]

No.	Item	Component	Range, %
Direct Cost			
1.	Equipment	All equipment in the flow sheet, uninstalled spare	15-40

		parts, surplus equipment, inflation on cost allowance, taxes, insurance, duties, allowance for modification during start up.	
2.	Equipment installation	Installation on equipment purchased, supports, paints, insulation.	6-14
3.	Instrument and controls	Purchase, installation, and calibration.	2-8
4,	Piping	Pipes material, fitting and vales, insulation.	3-20
5.	Electrical equipment	Switch, motors, wires, conduit, panels, atc.	2-10
6.	Buildings	Building construction, building auxiliaries, building services.	3-18
7.	Yard improvement	Site development for roads, walkways, railroads, parking, etc.	2-5
8.	Service facilities	Utilities, non process area equipment and furniture, distribution and packaging (raw material and product storage and handling)	8-20
9.	Land	Surveys and fees, property cost.	1-2
Indirect Cost			
10.	Engineering and supervision	Engineering cost-administrative, process, design, consultant fees, etc	4-21
11.	Construction expenses	Construction operation and maintenance, construction tools and supervision, warehouse, safety, permits, taxes and insurances	4-16
12.	Contractor's fees		2-6
13.	Contingency	Cost allocation in case of emergency or accident	5-15

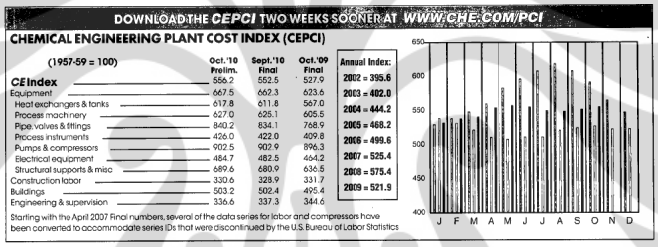
2.8.1.2 Working capital

Portion of investment that has been allocated for working capital is mainly counted for process start up, initial catalyst charges, raw materials and intermediate in the process, finished products inventories, and funds to cover outstanding accounts from customers [39]. The amount of capital investments that goes for working capital are ranges from 5 to 30 % depends on the ranges on products produced [39]. A simple single production may incur low percentage, while production with various has higher working capital.

2.8.1.3 Cost indexes

Some books provide data for cost related design or equipment that useful to indicate a rough estimation on the condition demanded. For instance, Peters and Timerhaus [14] introduce some data and tables regarding to the equipment and instruments price based on the year of the book was published. However, due to

time value of money in each cost data changes from year to year, a particular conversion factor is need to update the price to the recent value. This conversion factor is commonly known as cost index. By applying the cost index, cost at certain time in the past can be roughly equalized to cost at present time. The present value is equal to past value multiplied by the ratio of index in the past and present. There are different index sources that are widely used; Marshal and Swift Equipment Cost Indexes, Engineering News-Record Construction, and Nelson-Farrar Refinery Cost Index [14]. However, Chemical Engineering Plant Cost Index (CEPCI) based on journal 'Chemical Engineering' is recommended as it is very convenient to be applied. It presents the cost index related equipment, construction labour, buildings, engineering and supervision. The example of CEPCI is shown in figure 2.7.


DOWNLOAD THE CEPCI TWO WEEKS SOONER AT WWW.CHE.COM/PCI

CHEMICAL ENGINEERING PLANT COST INDEX (CEPCI)

(1957-59 = 100)	Oct.'10 Prelim.	Sept.'10 Final	Oct.'09 Final
CE Index	556.2	552.5	527.9
Equipment	667.5	662.3	623.6
Heat exchangers & tanks	617.8	611.8	567.0
Process machinery	627.0	625.1	605.5
Pipe, valves & fittings	840.2	834.1	768.9
Process instruments	426.0	422.0	409.8
Pumps & compressors	902.5	902.9	896.3
Electrical equipment	484.7	482.5	464.2
Structural supports & misc	689.6	680.9	636.5
Construction labor	330.6	328.9	331.7
Buildings	503.2	502.4	495.4
Engineering & supervision	336.6	337.3	344.6

Starting with the April 2007 Final numbers, several of the data series for labor and compressors have been converted to accommodate series IDs that were discontinued by the U.S. Bureau of Labor Statistics

Annual Index:
2002 = 395.6
2003 = 402.0
2004 = 444.2
2005 = 468.2
2006 = 499.6
2007 = 525.4
2008 = 575.4
2009 = 521.9

Figure 2.7 Cost Index published by 'Chemical Engineering' journal

2.8.1.4 Capital source

The fund allocated on the capital investments commonly comes from equity or loans. Equity fund is invested based on the shareholders' equity, which have value though opportunity cost. Shareholders itself expect that their investment will be returned through the revenue gained. Meanwhile, loan sources are mostly banks. It is counted as a debt and the loaner is obligated to return the fund according to the principal and interest that have been agreed.

2.8.2 Total production cost estimation

Beside capital investment, another economic component that is also important to estimate the total production cost is operating cost and annual fixed capital. These will determine the viability of process whether an alternative process scheme is necessary. Moreover, operating cost itself is group into variable and fixed

operating cost, which is classified accordingly to indecency towards the production rate.

2.8.2.1 Variable operating cost

Variable operating cost counts for materials cost that is spent in order to achieve the expected product rate. Materials that are covered in this are raw materials, utilities (service), and shipping and packaging [39]. The amount raw material is reliant to its requirement to satisfy the process. The kind raw materials in the project are also diverse. It can include chemical material, enzyme, catalyst, and other non-chemical materials. Similarly, utilities consumption is dependent on duty of machinery in the process scheme. Utilities do not consider electricity only, but also fuels and steam. The quote price of raw material and utilities are varied in different countries and significantly depended on market price.

2.8.2.2 Fixed operating cost

Fixed operating cost is invariant to the production rate. Fixed operating costs are general expenses or bills that have to be paid regardless the amount of products. This involved expenses on,

Labour do not only count for technical personnel, but also non-technical, such as administrator, accountant and security. Salaries to be paid on labour can be daily or annual basis, depending on the shift timing.

Supervision is regarded to manager or supervisor who is directly observed the operation of the plant. The number of management team depends on the size of the plant and can be cost consuming as well (roughly 20 percent on total labour cost)

Maintenance covers the cost of maintenance annually and can take up 5 to 15 per cent on purchased cost of machineries.

Operatingsupplies incurs on miscellaneous supplies to support process running, including charts, lubricants, test chemicals, maintenance supplies that does not covered as raw materials

Plantoverhead cost are assigned for hospital and medical services, general plant maintenance and overhead, general management, canteen, warehouse, salvage service, warehouse and storage facilities, payroll overhead including pensions, vacation allowance, social security, and life insurance,

Tax and contingency

2.8.2.3 Annual fixed capital / fixed charges

Fixed charges or fixed capital is cost that adds up to the production cost due to portion of capital investment, as it is an ongoing project through its lifetime. This cost is not reliant on product rate and it is charged annually regardless the production is operating or not. Annual fixed capital takes part 10 to 20 per cent on the annual production cost [14] due to depreciation, property tax, insurance tax, debt service, and purchased tax.

Depreciation is money spent for the decrease of value of asset during its lifetime due to physical wear. It is not an annual expenses but an allowance that company has put aside. In a simple straight-line depreciation method, items (excluding contractor's fees) contributed to capital investment are spread equally among the project lifetime.

Property tax is decided by the regional laws. In a region with high population, the portion can take up 2 – 4% of FCI, while small population area tax is 1-2% of FCI.[14].
Insurance tax comes from annual tax rates based on the clauses agreed between manufacturing and insurance company.

Debt service is a tax rates incurred if the capital source comes from bank loans. Certain tax rate has to be paid based on agreement.

Purchase tax. Taxes from the purchased asset in the capital investment (excluding contingency), which equally paid every year of, project lifetime.

2.8.3 Profitability

The profitability can be considered successful if the return on the investment is achieved. By this, profits gained from the production has overcome or covered all the capital investment and production cost over the lifetime. The key evaluation on the profitability is determined net present value, pay back period, rate of return and discounted cash flow rate of return.

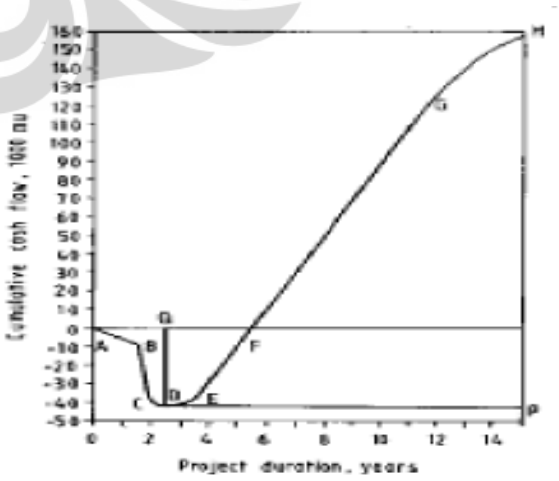
2.8.3.1 Cash Flow

Cash flow is the net flow of money in or out of the project over the lifetime where a one-year basis is mainly used for evaluation. Expenditures are denoted as negative flow, while incomes has positive flow. A typical cash flow can be seen in figure 2.8 [40]. AB indicates the money invested on research and development, design and other preliminary activities. BC is expenditures on land, building, and equipment. Point D indicated the maximum debt in the project. DEF shows that project is commissioning and project revenue is increasing over the years. Point F is the break point (or pay back period) where the debt is paid off. EFG indicates the constant positive rate of return of investment.

Figure 2.8 Cumulative cash flow diagram

2.8.3.2 Net present value (discounted cash flow)

The value of money alters from time to tome. Money obtained in the year one of a project will worth more the following years. This is done so because of the interest involved and cash flow that is cumulated throughout years. This behaviour is called time value of money and the useful method to predict and estimate the value of project asset or profit based on the time value of money is known as discounted cash flow. In figure the 2.8 for instance, future cash flow that has been cumulated and



discounted can be presented as net present worth (NPV) by applying;

$$NPV = \sum_{n=1}^{n=i} \frac{NFW}{(1+i)^n}$$

Where NFW is the net future worth, i is the project interest, and n is the year of the subsequent NFW.

2.8.3.3 Payback Period

As it has been mention previously in figure 2.8 that payback period is the time when all production revenue has covered the debt of capital investment during project preliminary. It is essential to investigate the payback time as it ensures that the project is profitable before the lifetime end.

2.8.3.4 Rate of return

The rate of return (ROR) is useful as it illustrate how hard is the investment money being uses. The projection of cash flow does not indicate it. Two cases of project will shows similar cumulative cash flow even though capital cost of each project is significantly different. The simplest estimation on ROR can be done as followed,

$$ROR = \frac{Cumulative\ net\ cash\ flow\ at\ end\ of\ project}{Life\ of\ project\ x\ original\ investment}\ x\ 100\ \%$$

2.8.3.5 Discounted Cash Flow Rate of Return (DCFRR)

Discounted cash flow rate of return (DCFRR) or internal rate of return (IRR) is the interest rate measured when the break point of pay back time (point F in figure 2.8) occurred exactly at the end of project lifetime. This rate is not the actual project interest, but the maximum interest that the project can afford.

2.8.4 Sensitivity analysis

The use of sensitivity analysis is to ensure that the project is viable even though major components contribute to cash flow has altered. Components that are considered, for instance, are raw material cost, production cost, capital and operating cost, or interest rate. To observe the fluctuation behaviour, comparison on the actual and the altered cash flow are essential. If the altered condition still

achieves cash flow within the project lifetime, the project is feasible. In addition, alteration can be accomplished by dropping or increasing the component mentioned to ± 20 to 50 per cent.

CHAPTER 3
METHODOLOGY

Methodology gives a comprehensive guideline of how the work will be broken down. It is started with a work plan, a brief list of steps that will be conducted and its expected outcomes. Following section will discuss methods regarding the design basis, strain selection, technology selection, and economic analysis

3.1 Work plan

This project has been divided into three main stages outlined as follows;

Stage 1: Process basis

- Gathering basis information on production of microalgae bioethanol
- Microalgae strain selection.

Expected Outcome- To have determined the basis used for production design and potential microalgae for the studies

Stage 2: Process Technology Selection and Design

- Weighing the advantages and limitations on the process and decide on an optimum preferred method.
- Developing process designs for each stages based on the available performance data found in the literature
- Generating a Process Flow Diagram based on selected technology

Expected Outcome- To able to conceptualize and conduct a technical assessment and design process on microalgae to bioethanol conversion process.

Stage 3: Economic Evaluation

- Estimating the equipment cost for each stages based on the designed process
- Applying the economic model to determine the relevant cost estimates
- Estimating the overall production cost.
- Determining the profitability of the project bioethanol plant using microalgae as feedstock

- Performing a sensitivity analysis on the projected costs

Expected Outcome- To complete economic analysis and determine the economic feasibility of the project

3.2 Design Basis

The bioethanol production plant is designed based on 50000 tonnes of dry biomass per year. The life span of the plant is expected at 10 years with 330 days of operating days in a year. A total of two weeks of plant shut down is scheduled for cleaning and maintenance purposes of the facilities and equipment. As the total completion time of biomass production is estimated around 10 days, the production will be distributed into 33 batches a year. Therefore, cultivation and dewatering stages are assumed to generate 1515 tonnes of dry biomass for downstream processes in a batch production.

3.3 Strain Selection

The culture strain that is adopted in this study is marine algae. This algae strain has carbohydrate and starch content of 32.52 (% w/w) and 11.32 (% w/w), respectively, with the composition of monosaccharides shown in table 3.1. Considering the highest carbohydrate content of up to 50 % w/w are found in *Spirogyra sp.* and *Porphyridium cruentum*, the carbohydrate content of*Chlorococum sp*. is moderately high as compared to other microalgae strains (Table A.1.1)[31].

Table 3.1 Composition of Chlorococum sp. [3]

Component	Composition (% w/w)
Total carbohydrate	**32.52**
Xylose	9.54
Mannose	4.87
Glucose	15.22
Galactose	2.89
Starch	**11.32**
Lipid	**19.30**
Others	**36.86**

3.4 Technology Selection

In selecting the most suitable method at each stage, it is important to keep in mind of the effectiveness and efficiency of the technologies in the designed process

configurations, which later contributes to the total production cost of bioethanol produced. The technology selections are useful as basis information in implementing process design and developing process flow diagram.

The production of biomass that includes cultivation and dewatering has been assed previously by Davidson et al [32]. Drying as part of the biomass production, however, was not assessed. Results from their study will use in this research, as it is taken into account into the total production cost. Thus, this study will briefly present the method regarding biomass production that has been adopted and the result that has been obtained whenever it is necessary.

Cultivation.The reliability of the cultivation is observed and evaluated in both open and close pond. Three predominant methods of cultivation that is considered are the raceway pond, horizontal tubular photobioreactor, and helical coiled (external loop) tubular photobioreator.

Dewatering.Removal of culture broth should be performed using dual stage dewatering option. Standalone centrifugation is rapid and has efficiency in slurry separation regardless the particle size. However, Figure A.1.1 has shown that relying on standalone centrifugation will incur intensive utilities cost, which is not feasible. Study by Davidson et al. [32] has proven that dual stage dewatering with flocculation reduces the cost burden of single dewatering system (figure A.1.2). Table A.1.3 indicates flocculants capabilities in achieving high cell recovery in different algae strain. The type of flocculants used for the following economic study is Chitosan where its optimum working capacity in different strains depicted in Table A.1.3. Following this, solar drying supposed to be done to obtained dry biomass.

Pre-treatment.Considering the large volume of dry biomass to be processed, combination of mechanical forces is not feasible due to high-energy consumption. Similarly, biological activity of microorganism is not preferable due to slow process; hence this method is not advisable for large production. Chemical

methods, on the contrary, have been recognised as a favourite pre-treatment method for a wide range of biomass. Study from Galbe et al. [19] has summarized the sugar yield obtained from corn stover using chemical pre-treatment to be considerably high, at a majority of up to 90 % of glucose yield conversion (Table B.1.5). However, Harun et al.[23]has carried out an in-depth investigation of *Chlorococum sp.* using two different chemical methods, alkaline and dilute acid. The highest ethanol yield in 0.75% NaOH concentration for 30 minutes is 26.1% g ethanol/ g algae [3], while 0.5% (v/v) of sulphuric acid treatment for 15 min produces 52% g ethanol/ g algae [23]. This study proved that ethanol production from dilute acid pre-treatment dominates over alkaline for almost double the yield. Moreover, in terms of economic aspect, pre-treatment cost by using dilute acid treatment on corn stover (AU$33.8 per litre ethanol) is lower than using lime catalyst (AU$41.5 per litre ethanol) (Figure A.1.6). Therefore, dilute acid as a catalyst is considered to be the most outstanding method for microalgal biomass pre-treatment.

Fermentation.The proposed technology for glucose to ethanol conversion is conducted in two different methods, SHF or SSF. Both methods involve saccharification by enzyme amylase and cellulose for glucose released. *Saccharomyces cerevisae* is used as biocatalyst to ferment glucose to ethanol.

Purification.Although purification is not included in the scope of study, the process will still be discussed by adopting purification process done by Aden et al.[33]. Here, dilute ethanol is concentrated in 2 stages of distillations; namely beer column to remove water and $CO_2$and rectifying column to concentrate ethanol. CO_2 will further be absorbed in scrubber, while molecular sieve recovers most of the ethanol from watercarried out at rectifying overhead.

3.5 Economic Analysis

From the bioethanol production model that has been developed, the plant economic shall be considered based on two major cost analysis; capital and operating cost. Moreover, the project viability relies on sales revenues as much as it does on capital and operating costs. The sales revenues determined the return of

investment on the plant, which is greatly controlled by market demand of sold products and their value.

To be noted, all the prices that are taken into account are converted into 2010/2011 index price and the unit price adopted are in Australian Dollar (1 USD = 0.945 AUD).

3.5.1 Capital Cost Estimates

The primary purpose for developing a process design found in the earlier chapter is to estimate the costs of each unit operation in accordance to the capacity of the design project. Since there is no precedent assessment of bioethanol production from microalgae in a large scale, the study is based on a factored estimate with probable accuracy up to $\pm$ 30 percent [14]. The total capital investment is classified into fixed capital and working capital expenditures.

3.5.1.1 Major Equipment Cost (MEC)

Major equipment cost assessed cost of equipment that is considered important and expensive, such as pump, batch reactors, centrifuge, and cooler/heater. The equipment cost is estimated using data provided by Peters and Timmerhaus[14]. The value will be fitted by scaling the equipment according to the design size.

The overall cost estimates in the purification stage are based on a technical report completed by Aden et al.[33], which involves four major equipment namely, the beer column, rectification column, molecular sieve dehydrator and scrubber. This is done so that the rough production cost burdened by purification stage is also known.

3.5.1.2 Fixed Capital Investments and Annual Fixed Capital

Fixed capital investment (FCI) comprises by 2 major costs; direct and indirect cost. Direct cost is all costs paid to the contractors in the initial stages of the project, including electrical system, piping installation, land and buildings and any costs associated with auxiliary facilities, utilities and administration. While indirect costs, on the other hand, count for engineering and supervision,

construction expenses, contractor's fee and contingency. Each component mentioned above are listed in table 3.2 with the estimated factors adopted from Molina Grima et al.[34].

In the case of land costs required under cultivation, each system was modelled based on the cost of two large agricultural properties situated in Gippsland, Victoria. The location of the plant is selected to ensure free supply of carbon dioxide and it is assumed that Gippsland where Australia's most carbon intensive power generation is, provides a suitable location for this purpose in the form of flue gas [32].

Table. 3.2 Component list on total fixed capital investment

	Item	Factor
Direct cost		
1.	Major Equipment Cost (MEC)	MEC
2.	Installation costs	0.3 MEC
3.	Instrument and control	0.1 MEC
4.	Piping	0.3 MEC
5.	Electrical	0.1 MEC
6.	Buildings	0.3 MEC
7.	Yard Improvement	0.1 MEC
8.	Service facilities	0.2 MEC
9.	Land	0.06 MEC
Indirect cost		
10.	Engineering and supervision	0.25 MEC
11.	Construction expenses	0.1 Σ items 1-9
12.	Contractor's fee	0.05 Σ items 1-9
13.	Contingency	0.06 0.1 Σ items 1-12
Total fixed capital investment (FCI)		**Σ items 1-13**

Moreover, portion of fixed capital investment that goes to fixed capital per year will be counted in cost for depreciation, property tax, insurance, debt service, and purchase tax. In this study, depreciation is estimated in 10 years of project lifetime. No debt services incurred in this work as 100% capital equity is taken.

The breakdown of factors correspond to each fundamental is presented in table 3.3

Table 3.3 Component list on annual fixed capital

Item		Factor
14.	Depreciation (straight line)	(Σ items 1-8, 10-13)/ 10 years
15.	Property tax	0.01 depreciation
16.	Insurance tax	0.006 depreciation
17.	Debt service	none
18.	Purchase tax	0.06 (Σ items 1-12)/ 10 years
Total fixed capital year		**Σ items 1-13**

3.5.1.3 Working Capital

An initial working capital amounting to 5% of the total capital investment was used to account for:

- Raw materials and supplies carried in stock
- Finished products in stock and semifinished products in the process of being manufactured
- Accounts receivable
- Cash kept on hand for monthly payment of operating expenses
- Accounts and taxes payable

3.5.2 Operating Cost Estimates

The operating cost estimate the cost burdened for production cost. This cost consists of variable operating cost (VOC) and fixed operating cost (FOC).

3.5.2.1 Variable Operating Cost

Variable operating costs are reliant on the production rate of the plant. Electricity consumption and raw materials constitute the major costs in production operation. Raw material costs are defined by the amount used in a year production, which price is dependent on the market price. Electricity consumed to operate equipment is estimated based on the equipment power corresponded to its running hour in a

year production. However, energy required in biomass production (cultivation and dewatering) was estimated based on the data from Davidson et al.[32]. Their study has taken into consideration the energy requirements for mixing; by airlift pump used in photobioreactors or paddle wheel in raceway ponds.

3.5.2.2 Fixed Operating Cost

Fixed operating costs include labour and supervision, maintenance materials, operating supplies, plant overheads, property taxes, and contingency. These costs are generally independent of production rate and listed in table 3.4.

Labour cost accounts for a total of 12 shift operators, 4 engineers, and 2 supervisors employed at standard Australian pay rates. Corresponding pay rates also apply for 2 non-process labourers working during normal operation, 4 security guards working on a two-shift swing, and 2 administrators.

Table 3.4 Component list on fixed operating cost

	Item	**Factor**
1.	Labour	
2.	Supervision	0.2 (item 1)
3.	Maintenance	0.04 MEC
4.	Operating supplies	0.004 raw material cost
5.	Plant overhead	0.55 Σ items 1-3
6.	Tax	0.016 (VOC + Σ items 3-4)
7.	Contingency	0.05 VOC
Total fixed operating cost (FOC)		**Σ items 1-7**

3.5.3 Cost of Ethanol Production

The production cost of ethanol from microalgae is based on the stages mentioned earlier in this report, by comparing the costs of three different cultivation systems, two dual-stage dewatering processes and finally between SHF and SSF. The total of this cost is the sum of fixed capital year cost and total of operating cost. Subtracting the production cost with total volumetric of ethanol produced per year

will present the cost of production burdened in each litre of ethanol. Furthermore, the lowest of figure resulted will determine the most applicable and cost effective technology scheme.

3.5.4 Profitability

Cash flow analysis predicts the profitability of the project. The following analyses are based on the project plant life of 10 years with interest rate at 10%, scheduled as:

1) In the first and second year (t = 1, 2): Design &Construction is 90% (30 % at year one and 60% at the second year) of fixed capital cost
2) In the third and (t = 3): Commissioning & start of production with 10% of fixed capital cost, 100% working capital, 100% fixed operating cost and 30% of variable operating cost.
3) In the fourth year (t=4), 70% of operating production. 100% of fixed operating cost and 70% of variable operating cost.
4) From fifth year on (t = 6 - 10): 100% normal operating production. 100% fixed operating cost + 100% variable operating cost.

As the production starts to run, the revenue from the selling of product and by-product can be achieved. From the net discounted cash flow, the payback period and the net present value of end project lifetime is estimated. IRR is found using trial error by alternating the interest rate that gives net cash flow year ten equal to zero. ROR is obtained by dividing the net cumulative cash flow at end of project by the project lifetime and original investment.

3.5.5 Sensitivity Analysis

Sensitivity analysis is performedby altering economic component of the cash flow at $\pm$20% variation. Such components that is analysed are the sensitivity on sales price, production rate, fixed and variable operating cost, capital investment, and interest rate.

3.6 Bioethanol production with lipid extraction as part of the process

An additional scenario of process production is made considering lipid extraction is part of biomass production which economic aspect also plays big role in

determining the net production cost of bioethanol production. Prior pre-treatment, wet biomass from dewatering will directly delivered to lipid extraction department to collect the valuable oil content in the microalgae strain. As mentioned earlier in table 3.1 that *Chlorococcum sp.*has lipid content of 19.3 %wt in its intracellular. This oil derived, which is the main feedstock of biodiesel production, is sometime preferred to be extracted before ongoing to the bioethanol production process. Hence the estimation of total bioethanol production has to include this process.

The technology implemented for lipid extraction is based on solvent extraction with two stages process of lipid extraction with ethanol 96 %(v/v) and lipid purification using hexane and water. 90 % of crude lipid is assumed collected in this process. The design of this stage will refer from previous study by Davidson et al. Furthermore,the economic parameters on this process will also be estimated based on section. 3.5.1 – 3.5.3.

CHAPTER 4
RESULT AND DISCUSSION

4.1 Process Design

The following section illustrates the process design of bioethanol production from microalgae strain in large-scale production. In general, the information required for design purposes were taken based on literature, where further rough calculation has been carried out to estimate the design processes of entire bioethanol production. The overall Process Flow Diagram of bioethanol production is depicted in Appendix A.2. Database is summarized in Appendix A.3 and sample calculation is appended in Appendix A.4.

4.1.1 Biomass production

The study of biomass production from microalgae strain has been carried previously by Davidson et al[32]. In brief, strains are pumped from the inoculum storage tank (RP T-2) to raceway pond and medium growth, such as nutrients and seawater is added to enhance culture growth (see PFD A.2.1). A mixture of flue gas and air are injected to provide sufficient gas exchange within the culture pond. Similar mechanisms are adopted for both photobioreactor in horizontal and external loop (see PFD A.2.2 and A.2.3). From the closed chamber, the strains are pumped to solar array, before recycling to the column. Table 4.1 shows the design unit of each cultivation method along withthe number of microalgae cultivated. It is assumed that only 80% of the biomass is harvested from cultivation. The detail design specification can be referred to Table A.4.1.1

Table 4.1 Unit design on cultivation

Variable	Horizontal Tubular Reactor	External Loop Reactor	Raceway Pond
Biomass produced/ batch (tonne)	757.5	568.75	947.5
Unit Description	132 parallel tubes; tube length of 80m. Tube diameter 0.06m	4m airlift, 80m tubular run; Tube diameter 0.06m	1050m^2 pond; 14m wide and 75m long; 0.2m depths.

Dewatering is done in dual stage process. The cultures biomass is pumped to an open tank flocculants (DR E-1) where 40 mg/L of chitosan flocculants loading with 90% efficiency is used (Table A.4.2.1). Table 4.2 describes the characteristic of flocculation in different cultivation method. Further detailed calculation on flocculation specification can be referred to Table A.4.2.2-4. Following flocculation, biomass slurry is transferred to a disc stack centrifugation (DR E-2). The power consumed during centrifugation is estimated at 1.5 MW for horizontal PBR, 1.6 MW for external loop PBR, and 115 MW for raceway pond (table A.4.2.5-7).

Table 4.2 Flocculation Characteristic [32]

	Horizontal PBR	**External Loop PBR**	**Raceway pond**
Flocculants per tank (kg)	240	260	1600
Biomass concentration before (kg/m^3)	4.525	3.8	0.585
Biomass concentration after flocculation (kg/m^3)	36.5	30.4	4.7

4.1.2 Pre-treatment

Dry microalgal biomass that is attained is delivered to further biomass-to-ethanol process. Treatment with dilute sulphuric acid at high temperature will decompose crystalline structure of cell wall and solubilise cellulose. The hydrolysate liquid and solid are flashed to atmospheric temperature to vent out most of water through steam. Following blowdown tank, the remaining hydrolysate is separated from the solid biomass by allowing settling using centrifuge. The liquid has to be neutralized by raising pH to 5 in order to provide an appropriate environment for enzyme hydrolysis. Dilution water containing soluble polymeric sugar is then sent to SHF or SSF..

On the basis of 33 batches of a 10 days production per year, around $1515x10^3$ kg of dried, fine biomass in a batch will be pre-treated. Due to large amount of dry biomass, each batch is distributed into an hourly cycle, where only 6313 kg of dry biomass per cycle is transported though belt conveyer (PT BC-1) to pre-treatment tank PT E-1. Dry biomass is treated with 0.5 % v/v of sulphuric acid in 10 g/L

biomass loading for 15 minute residence time at temperature of 160 °C and 6 atm (see PFD A.2.5). This condition parameter are implemented based on the highest ethanol yield using dilute sulphuric acid pre-treatment from a study by Harun et al.[23]. Apparently, 3% v/v of sulphuric acid that has been adopted by Harun et al.[23]is relatively high for large-scale production. This concern arises as the price for sulphuric acid is expensive, which incurs high operational cost, and corrosion factor on the equipment becomes considerarion. On the other hand, Carvalheiro et al. [25] proposed that 0.5-1.5% of sulphuric acid treatment at 121-160°C is the most favourable condition for industrial application. Furthermore, water from water storage tank PT T-2 is also pumped to the tank to attain biomass loading. As the sulphuric acid treatment require high temperature environment, pre-heated saturated steam of 162°C and 6.03 atm is directly injected to the jacketed tank. A total of six 185 m^3-pre-treatment tanks PT E-1 are required to accommodate each cycle of biomass. In addition, it has been assumed that 100% of carbohydrate (32.52 %w/w) and starch (11.32 %w/w) contents are digested by sulphuric acid activity. Table 4.3 and table A.4.3.4 summarize the condition and the sizing of pre-treatment tank PT E-1, respectively. The amount of monomeric sugar and starch that is converted from biomass cellulose is presented in Table 4.4.

Table 4.3 Pre-treatment tank PT E-1 conditions

Acid concentration	0.5 % v/v
Biomass loading	10 kg/m^3
Pressure	6 atm
Temperature	160 °C
Residence time	15 min
pH	2.93
No. of cycle	240
Biomass per cycle	6313 kg/hr

From pre-treatment tank, the liquid hydrolysate and solid is continued to blow down tank PT E-2 to be flash cooled to 1 atm with residence time of 10 min. About 26.2 % w/w of 100°C steam is vapoured out and used to pre-heat the beer column feed on PR HX-1. The condensed steam will later be sent to waste water treatment.

Table 4.4 Monomeric sugars converted from cellulose per batch

Component	Composition (kg)
Total carbohydrate	**850,910**
Xylose	144,546
Mannose	73,788
Glucose	230,606
Galactose	437,48
Starch	**171,515**

The remaining solid and hydrolysed is then pump for solid liquid separation using solid bowl centrifuge (PT E-3) with 3.28 kg/s of solid capacity and 750 KW of energy requirement [14]. By allowing settling time for 30 min, it is expected that 95% of the solid material is being removed, leaving $111x10^3$ m^3 of total hydrolysate containing soluble cellulose. The remaining biomass residue is collected and stored for further commercial use.

Following the solid-liquid separation, the hydrolysate with pH of 3.61 has to be neutralized by NaOH by generating reaction as followed;

$$H_2SO_4 + 2\ NaOH \rightarrow Na_2SO_4 + 2\ H_2O \qquad \text{(Rxn 4.1)}$$

The addition of a total of 4.45 kg of NaOH pellet for every batch will bring up the hydrolysate pH to 5 [13]. The occurring reaction form a sodium sulphate that acts as a buffer to maintain pH condition; hence further NaOH addition in enzyme hydrolysis in unnecessary.

Three neutralization tanks (PT E-4) with capacity of 235 m^3 will accommodate theacidic hydrolysate per cycle. The residence time for neutralizing is around one hour [34]. Agitator (PT A-2) with approximate power requirement of 98.5 W/ m^3 is introduced to each tank to aid homogenization.

After done with pre-treatment section, the neutralized hydrolysed is temporarily stored to 18600 m^3 of hydrolysate storage tank PT T-3 before it is sent to enzyme hydrolysis (SHF E-3) or saccharification tanks (SSF E-2).

4.1.3 Separate Hydrolysis and Fermentation

The total of liquid hydrolysate from pretreatment to SHF will be divided into 4 trains due to large number of volume (see PFD A.2.6). The capacity of single enzyme hydrolysis tank (SHF E-3) is approximately 1600 m^3, where 20 of tanks are available. Cellulase and amylase are conveyed through SHF BC-1 and SHF BC-2, respectively. In each hydrolysing tank, the cellulase and amylase loading are kept at 0.02 and 0.01 g enzyme/ g substrate, respectively [10]. The tank environment is maintained at pH 5 and temperature 40°C for 72 hour of estimated residence time [10]. Mixing (SHF A-3) is available in every tank with power requirement of 19.4 watt/m^3. During enzyme hydrolysis, approximately 90% of the complex sugar in carbohydrate and starch are hydrolysed to hexose sugars. Table 4.5 summarized the condition of enzyme hydrolysis SHF E-3 tank. Upon completion, the hydrolysed liquid is cooled to 37°C before sending it to fermentation tank and the second train from PT T-3 is delivered to hydrolysis tank.

Table 4.5 Enzyme Hydrolysis tank condition SHF E-3

Temperature	40 °C
pH	5
Residence time	72 hour
Cellulase loading	0.02 g/g substrate
Amylase loading	0.01 g/g substrate
Approximate sugar converted	90 %

Before entering the fermenter tank, *Saccharomyces cerevisae* is cultivated in yeast jacketed fermenter vessel SHF E-1 that holds 3.2 m^3 of volume (refer PFD A.2.6). The yeast seed loading is kept at 5 g/L, while 3 %v/v of Luria broth and supporting medium ammonium chloride, ammonium sulphate, and magnesium sulphate are supplied at concentrations of 2 g/L, 1 g/L and 0.8 g/L. Yeast is treated at aerobic condition at temperature 37°C for one day. Oxygen exchange is supplied by vigorous mixing (SHF A-1), with estimated agitation power of 14 watts. Moreover, since yeast is very susceptible to contamination, tank sterilization is necessary prior to yeast seeding. This can be achieved by steam injection at 160°C for 15 min to the vessel jacket. Table 4.6 lists the condition yeast fermentation tank SHF E-1.

Table 4.6 Yeast fermentation tank condition SHF E-1

Size of vessel	3.2 m^3
No. of tank required	1
Temperature	37°C
Residence time	1 hour
Condition	Aerobic

Saccharomyces cerevisae is used as biological catalyst to convert the fermentable sugar into ethanol and carbon dioxide according to followed reaction;

$$C_6H_{12}O_6 \rightarrow 2\ C_2H_5OH + 2\ CO_2 \qquad \text{(Rxn. 4.2)}$$

Fermentation takes place in 20 of 1600 m^3 vessel jacketed fermentation tank SHF E-2 and split into 4 trains (refer PFD A.2.6). As the fermentation is carried out in anaerobic condition, nitrogen is purged to vent out the air inside the vessel followed by similar sterilisation as mentioned before. Finally, the pre-treated yeast seed, hydrolysed hydrolysate and nutrients are fed to the vessel and the fermentation is accomplished in 48 hours [13] at constant temperature of 37°C. 54 KW of agitator is given to each tank. Moreover, about 10 gram of sodium hydroxide has to be added to act as a buffer to maintain the bioactivity of yeast at pH 7. This is done due to the CO_2 produced during conversion that gives an acidic effect to solution.

It is assumed that the biocatalyst activity of *Saccharomyces cerevisae* can only achieve 90% conversion of hexose sugar. Pentose sugar- xylose, is not converted. After all trains of the fermentation are completed, approximately 210 tonnes of dilute ethanol (40 % w/w) produced is stored at beer well SHF T-1 and 220 tonnes of gaseous carbon dioxide is released per batch. The CO_2 off gas is condensed and sent to scrubber PR E-5. The ethanol carried in the CO_2 is scrubbed and returned to beer well. Summary of fermentation tank SHF E-2 condition and sizing are given in table 4.7 and table A.4.5.5, respectively.

Table 4.7 Fermentation tank SHF E-2 conditions

Temperature	37°C

pH	7
Residence time	48 hour
Condition	Anaerobic

4.1.4 Simultaneous Saccharification and Fermentation

SSF is carried out in 20-jacketed fermentation tanks SSF E-2, each has capacity of 1600m^3 (refer PFD A.2.7). This process is divided into 4 trains from a single batch biomass. The fermentation tanks sizing is presented in table B.4.6.6.

Similar to SHF, SSF vessel involves nitrogen and steam injection to the vessel to meet sterile and anaerobic requirements. The yeast is grown in yeast fermenter tank SSF E-1, which specification can be referred to table 4.8 After the enzymes, yeast, nutrients, and hydrolysate are added to the tank, the temperature is kept at 37°C, which is optimum temperature for yeast, and pH 5. The agitation power for SSF (SSF A-2) is around 54 KW. The total estimated residence time for one fermentation train is 72 hours. During this time, it is assumed that only 70% of fermentable sugar can be saccharide, where only glucose is fermented by *Saccharomyces cerevisae*. The total conversion of glucose to ethanol is 90 percent. This results a total of 160 tonnes of ethanol and 153 tonnes of CO_2 per batch. Table 4.9 summarized the condition of SSF fermentation tank.

In addition, approximately 23% w/w of dilute ethanol is stored in beer well SSF T-1 prior purification. CO_2 off gas (99.8 % w/w) is captured and condensed in SSF HX-1, followed by CO_2 absorption in PR E-5. Recovered ethanol in CO_2 off gas from scrubbing will be collected back to beer well SSF T-1.

Table 4.8 Yeast fermentation tank condition SSF E-1

Size of vessel	3.2 m^3
No. of tank required	1
Temperature	37°C
Residence time	1 hour
Condition	Aerobic

Table 4.9 Fermentation tank SSF E-2 conditions

Temperature	37°C
pH	5
Residence time	72 hour
Condition	Anaerobic

4.1.5 Purification

The dilute beer from fermentation will be passed to distillation and molecular sieve to recover up to 90 percent of raw dilute ethanol to a concentrated ethanol. First beer column removes most of dissolve CO_2, water and soluble material, while the second column, rectifying column, produce concentrated ethanol. Molecular sieve dehydrate the remaining water vapour from rectifying column overhead, resulting pure ethanol as final product, while the some of the regenerated water-ethanol mixture is recycled back to distillation column. Scrubber is applied to recover ethanol carried in the off gas from fermentation tank and beer column overhead. Since product recovery scheme is not part of the scope to be simulated, the estimated design for industrial scale bioethanol purification is adjusted based on the study of Aden et al.[33], which focus on bioethanol production from lignocellulosic biomass.

Before entering the beer column, the dilute raw ethanol is heated to 95°C using flashed steam from blow down tank PT E-2 in PR HX-1, followed by a further heating to 100°C with reboiled stream for bottom beer column (see PFD A.2.8).

The beer column (PR E-1), which operates at 2 atm, consists of 32 trays,reflux ratio at 3, and diameterof 4.4 meter. Overhead of beer column contains around 88 %w/w CO_2 and 12 %w/w ethanol with trace of water is absorbed through scrubber PR E-5 and the recovered ethanol is stored to beer well, while 90% of water is removed at bottom. Overall, 99 % of ethanol from beer feed is drawn from side column and recycled to rectifying column PR E-2.

Around 97 % w/w of CO_2 from fermentation tank off gas and 88 % w/w of CO_2 from beer overhead containing ethanol enter water scrubber (PR E-5) (see PFD A.2.9). Scrubber itself consists plastic packing with 4 theoretical stages. With the

water from storage tank is fed at top scrubber, the CO_2 is absorbed and vented at top scrubber. Approximately 99 % of the CO_2 is collected and stored in CO_2 storage tank as co-product.

Rectifying column PR E-2 operates at 2 atm. 93.4 %w/w and 98.4 %w/w ethanol in water mixture from beer column and dehydration, respectively, enter rectifying column. The ethanol vapour from beer column is fed at top tray and the recovered mixture from molecular sieve is fed in column mid. Around 99% of the total ethanol fed to the column flows to overhead in 0.3 %w/w of water mixture. 93% water content from rectifying feed is collected to rectifying bottom and recycled to water storage.

The ethanol vapour in overhead continues to molecular sieve PR E-4 and E-5 for water removal. Two molecular sieve chambers (PR E-3 and E-4) will work correspondently every 8 hours for bed regeneration. About 95 % of the carried water will be removed by dehydration, which is recycled back to rectifying column feed. At the overhead, 80 % of purified and recovered ethanol vapour can be regarded as 99 % pure bioethanol and stored in bioethanol well after condensation. The summary of purification calculation is given Appendix B.4.7

4.1.5.1 Product and Co-product Recovery

From the entire process of the bioethanol production from microalgae, there are four components that can be considered as major outputs; namely bioethanol, as the main product, CO_2 and biomass residue as the co-product, and water and solid material as waste product. As the pure bioethanol is sold to the fuel market, CO_2 is also considered to be recycled back for microalgae cultivation or sold as carbon credit to other industries. Microalgal biomass residue that has been separated by centrifugation during pre-treatment can be sold as animal feed. Since the biomass residue is high in lipid content, it can be utilized as feedstock for biodiesel production, which can be added to plant revenue. Water output is treated and recycled back to water well, while solid waste material is disposed to waste treatment. Table 4.10 summarized the major output produced in both SHF and

SHF. Detailed calculation on components recovery can be referred in Appendix B.4.7.

Table 4.10 Summary of product, co-product, and waste products produced in overall ethanol production using SHF and SSF.

Material	SSF (kg/ batch production)	SHF (kg/ batch production)
Bioethanol	188,417	235,822
CO_2	152,309	247,812
Biomass residue	850,909	850,909
Water	9,981	9,981
Solid material	342,513	213,062

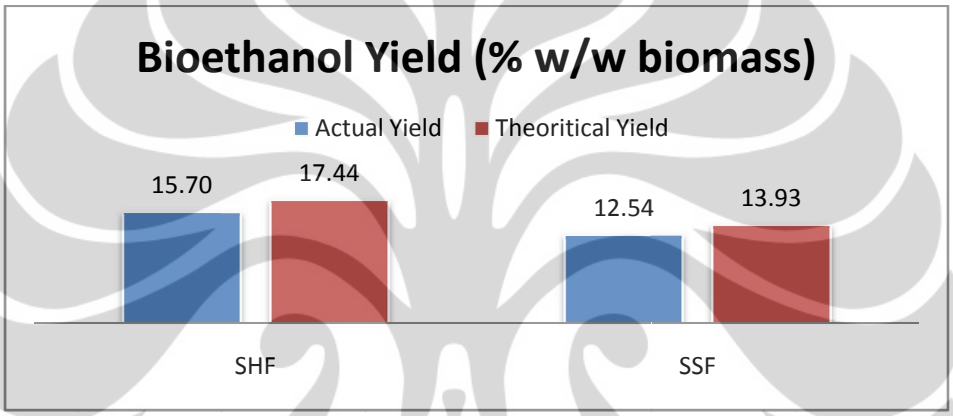


Figure 4.1 Annual bioethanol yields in weight per weight of biomass

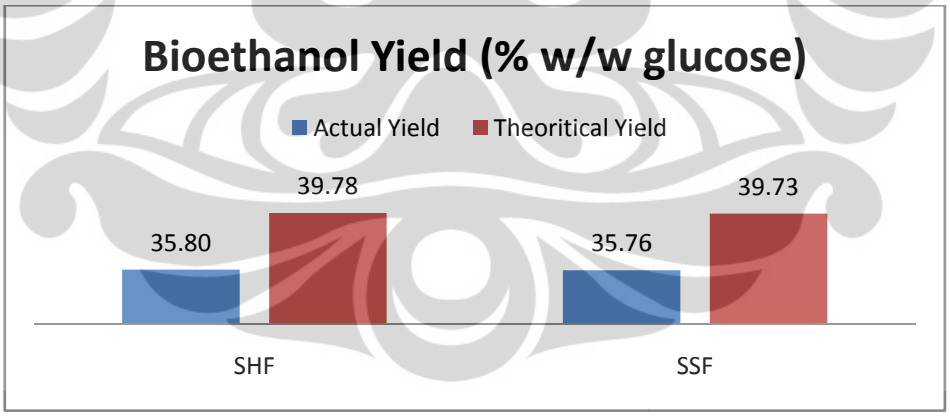


Figure 4.2 Annual bioethanol yields in weight per weight of reducing sugar

Summarizing the overall end product of bioethanol in the entire production, SHF performed better over SSF in terms of bioethanol yield. From figure 4.1, around 15.7% of gram mass content of dry biomass can be converted into bioethanol by implementing SHF. This yield drops by 4% when fermentation technology is

alternated to SSF because the operating condition does not support higher sugar conversion. The difference between SHF and SSF bioethanol yield might not be drastic, but this will greatly influence total production cost per litre or bioethanol produced which will be discussed in the following chapter. In comparing between actual and theoretical yield, the estimated bioethanol yield is slightly lower by 1% from the maximum conversion biomass to bioethanol, regardless the fermentation method applied. Similarly, annual bioethanol converted per gram of fermentable sugar released is roughly equal in SHF and SSF (Figure 4.2). This points that fermentation by *Saccharomyces c*. in any methods applied are able to convert most of the glucose fed.

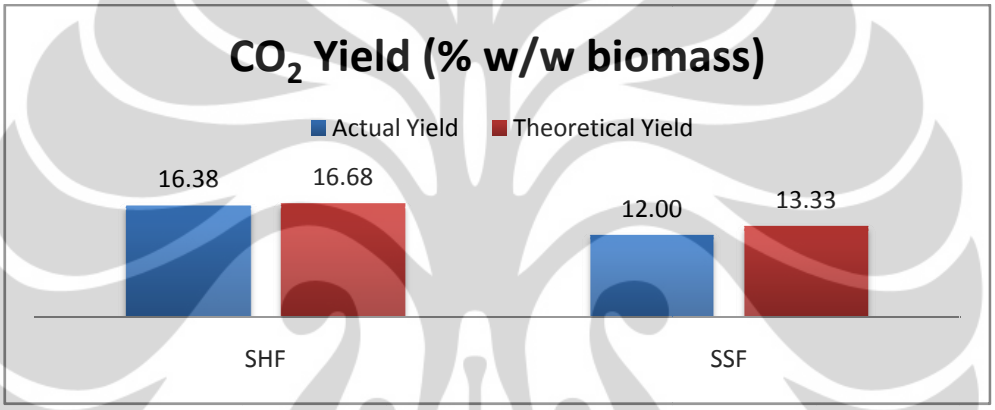


Figure 4.3 Annual CO_2 produced in weight per weight of biomass

Co-product carbon dioxide, on the other hand, annually yields 4 % lower in SSF implementation (figure 4.3). This can be a good prospect for SSF if minimum emission is main consideration of the CO_2 out to atmosphere. However, higher CO_2 capture can accommodate microalgae cultivation needs for carbon dioxide gas exchange.

4.2 Economic Model

The economic analysis is done to further evaluate microalgae as an alternative feedstock for bioethanol production in terms of economic viability. The industrial costs of processing must be cheaper than those for processing existing feedstock, in particular lignocellulosic-based materials. This part of evaluation includes capital cost estimation, operating cost estimation, cash flow and sensitivity

analysis. The estimation methods and results are explained in this section, with the calculations found under Appendix B.

4.2.1 Capital cost

Figure 4.4 shows the capital cost distribution across all systems, it is clear that the capital cost for cultivation pre-dominates the costs for other stages by almost 50% or more due to the large capacity of biomass production and complexity of equipment in photobioreactors. Therefore, it is suffice to assume that the technologies involved in biomass production (cultivation and dewatering) plays a major role in deciding the overall cost of the bioethanol plant. Pre-treatment and fermentation assuming constant cost for all systems, is more apparent in raceway pond as compared to the two photobioreactor options.

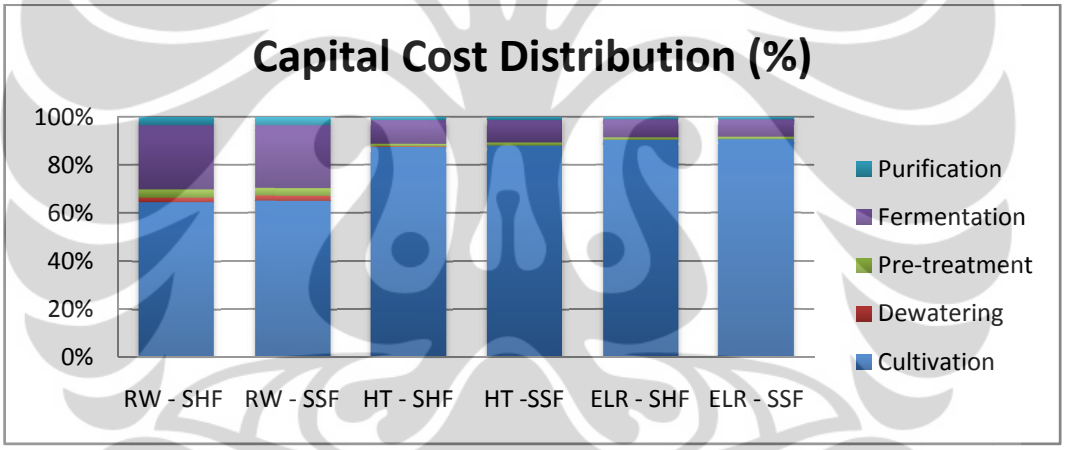


Figure 4.4. Breakdown of capital cost based on section of plant

Table 4.11 summarises the capital investment for all systems. Raceway pond using SSF option for fermentation achieves the lowest cost at AU$1183 million whereas External Loop Reactor working under the same setting would bear a total cost of AU$4191 million. As mentioned earlier, the cultivation cost for ELR is too great that other cost saving options in downstream processes are unable to offset the cost incurred.

Moreover, table 4.12 summarized the total annual capital. Reflecting to the total capital investment, the lowest annual capital is also found in raceway pond

cultivation with SSF for fermentation. This has proven that RW-SSF is low in capital cost as iteliminates the cost spent for enzyme hydrolysis.

Table 4.11. Capital cost comparison for different design process configuration

Plant Area	SHF (AU$ million/yr)			SSF (AU$ million/yr)		
	RW	HT	ELR	RW	HT	ELR
Cultivation	$734.10	$3,617.23	$2,721.25	$2,721.25	$734.10	$3,617.23
Dewatering	$23.26	$5.40	$4.82	$4.82	$23.26	$5.40
Pre-treatment	$37.65	$37.65	$37.65	$37.65	$37.65	$37.65
Fermentation	$305.40	$305.40	$293.96	$305.40	$293.96	$293.96
Purification	$38.12	$38.12	$38.12	$38.12	$38.12	$38.12
Total FCI	**$1,138.54**	**$3,107.24**	**$4,003.80**	**$1,127.10**	**$3,095.81**	**$3,992.36**
Total WC	**$56.93**	**$155.36**	**$200.19**	**$56.35**	**$154.79**	**$199.62**
Total Capital Investment	**$1,195.46**	**$3,262.61**	**$4,203.99**	**$1,183.45**	**$3,250.60**	**$4,191.97**

Table 4.12. Annual fixed capital cost in each stage

Plant Area	SHF (AU$ million/yr)			SSF (AU$ million/yr)		
	RW	HT	ELR	RW	HT	ELR
Cultivation	$77.08	$290.80	$386.15	$77.08	$290.80	$386.15
Dewatering	$2.70	$0.56	$0.63	$2.70	$0.56	$0.63
Pre-treatment	$3.97	$3.97	$3.97	$3.97	$3.97	$3.97
Fermentation	$33.60	$33.60	$33.60	$29.28	$29.28	$29.28
Purification	$1.05	$1.05	$1.05	$1.05	$1.05	$1.05
Total Fixed capital year	**$118.40**	**$329.98**	**$425.39**	**$114.07**	**$325.65**	**$421.07**

4.2.2 Operating Cost Estimates

In the use of a dual-stage dewatering process, Chitosan was the preferred flocculants with costs estimated at USD 11/kg [35]. Pure industrial sulphuric acid is priced at AUD 300/m^3 based on Alibaba express. The costs of cellulose and amylaseused in hydrolysis calculations are based on 15-20 cents USD/gallon bioethanol and 2-4 cents USD/gallon[2]. Yeast from *Saccharomyces*

*cerevisiae*ispriced at USD86/kg based on Sigma Aldrich.In terms of the cost associated with wastewater treatment, it was assumed that 20% of the annual broth throughput required treatment [32].

In this economic model, several key assumptions were made with regards to water use, funding, and exchange rate. Since *Chlorrococcum sp*. is a marine algae, seawater which is free of chargeis used throughout the production line with the exception of fresh water used during yeast fermentation. All funding assumes to be acquired through venture capital; therefore debt service has been eliminated in cost estimation.

Table 4.13 summarized the total cost incurred for production operational based on operating and fixed operating cost. Further detailed calculation can be referred in appendix B.6

Table 4.13. Variable operating cost breakdown on each stage at different plant

Plant Area	**SHF (AU$ million/yr)**			**SSF (AU$ million/yr)**		
	RW	**HT**	**ELR**	**RW**	**HT**	**ELR**
Cultivation	$24.07	$10.81	$10.29	$24.07	$10.81	$10.29
Dewatering	$38.60	$4.95	$5.94	$38.60	$4.95	$5.94
Pre-treatment	$13.99	$13.99	$13.99	$13.99	$13.99	$13.99
Fermentation	$3.82	$3.82	$3.82	$0.40	$0.40	$0.40
Purification	$9.20	$9.20	$9.20	$9.20	$9.20	$9.20
Total VOC	**$89.68**	**$42.77**	**$43.24**	**$86.26**	**$39.35**	**$39.82**
Total FOC	**$119.38**	**$107.73**	**$125.43**	**$114.85**	**$64.85**	**$65.35**
Total OC	**$209.06**	**$150.50**	**$168.67**	**$201.11**	**$104.20**	**$105.16**

4.3 Cost of Ethanol Production

4.3.1 Cultivation

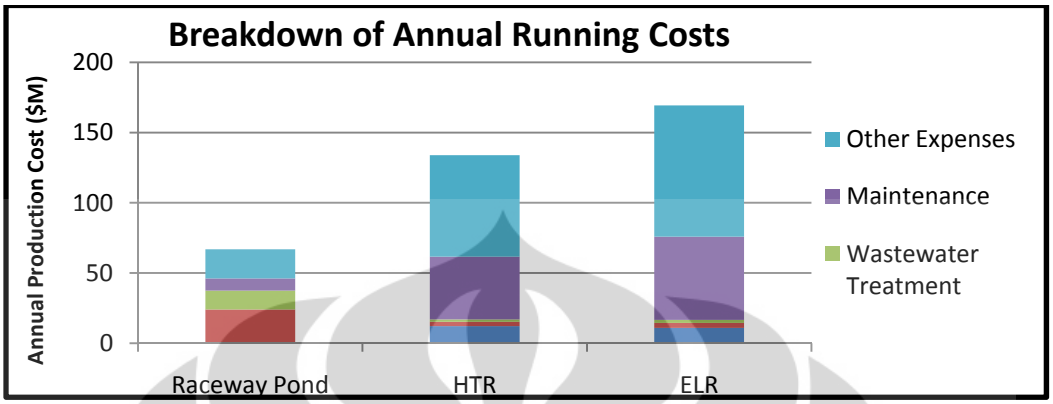


Figure 4.5 Breakdown of annual running cost for cultivation options[5]

According to Davidson et al.[32], raceway pond was found to be the cheapest biomass production system (AU$ 3/kg) followed by horizontal tubular reactor at AU$ 10/kg and finally external loop reactor at AU$ 13/kg. At this given production capacity, the cost required to build a much complex photobioreactor (HTR or ELR) are definitely higher as compared to those required in raceway pond. A breakdown of the annual operating cost incurred by each system is illustrated in figure 4.5

From figure 4.5, wastewater treatment and culture medium are the major cost contributors in raceway pond. Its lower volumetric productivity has resulted in greater culture medium requirement and larger fluid volume for processing in wastewater treatment. On the contrary, photobioreactors have higher electricity and maintenance costs. The energy consumption in both photobioreactors is greater as compared to energy used to operate paddle wheel in raceway pond configuration. Davidson et al. have also considered the monetary lost incurred due to higher risk of contamination in raceway pond. The study has shown that it requires more than 70% of algae to be contaminated for horizontal tubular reactor to be the cheaper cultivation option.

4.3.2 Dewatering

The cost of dewatering process for raceway pond is approximately 15times larger than photobioreactors. This is mainly due to large processing volume in raceway

pond which requires longer harvesting period and consequently greater energy costs [32]. As can be seen from Figure 4.6, the capital cost of dewatering is less significant compared to its operating cost. Based on the findings done by Davidson et al., the reliability of standalone centrifuge option or centrifuge coupled with flocculation has deemed it more suitable as a raceway pond-dewatering alternative. The study also suggested a single-stage centrifugal dewatering process for photobioreactor cultivation systems. However the study mentioned above has failed to make comparisons with dual-stage flocculation and filtration. In the case of raceway pond, upon conducting similar calculations based on the data provided by Davidson et al., the costs of standalone centrifugation and both dual-stage dewatering options (Flocculation+Centrifuge, Flocculation + Filtration) are comparable, with the latter lower in energy cost.

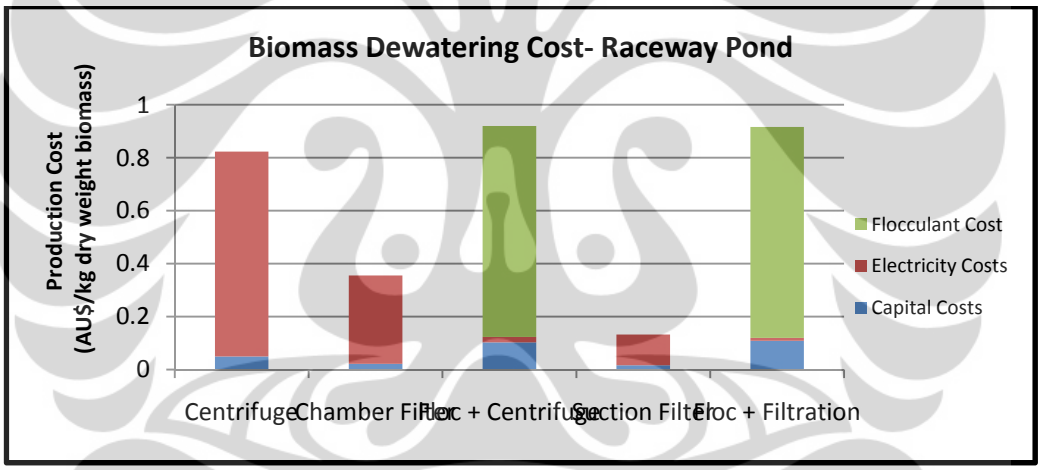


Figure 4.6. Biomass Dewatering Costs-Raceway Pond

Referring to Figure 4.6, the cost of standalone filtration is significantly lower in comparison to other dewatering alternatives. As mentioned previously, this option raises concerns when applied in a large scale due to membrane clogging and formation of compressible filter cakes, which consequently led to higher operating costs and possible addition to filtration units to maintain throughput.

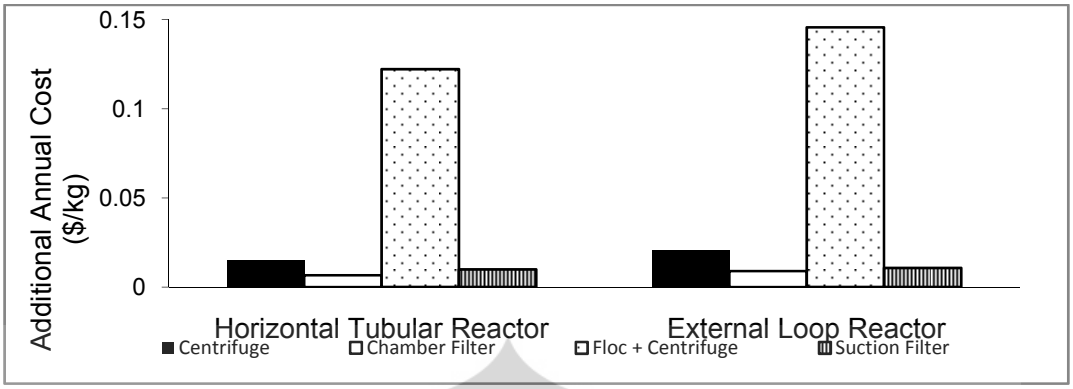


Figure 4.7 Biomass dewatering cost- HTR and ELR[5]

The graph also shows that dual-stage dewatering process incurs higher cost due to high flocculants cost. However, process development in this technology is believed to achieve greater impact in cost reduction, thus revolutionise bioethanol production [15]. In addition, the preceding flocculation process has led to a significant decrease in energy consumption in centrifugation by approximately 98%. In the case of dewatering methods used in photobioreactors, centrifugation clearly achieves the lowest additional annual costs at AU$ 0.01-0.02/kg (figure 4.7). Therefore, dual stage dewatering unit (Flocculation + Filtration) and single-stage centrifugation would be the suitable technology for harvesting microalgae in raceway pond and photobioreactors respectively.

4.3.3 Pre-treatment

From Figure 4.4, pre-treatment accounts for as high as 3% of the total production cost in raceway pond configuration giving AU$2.85 per litre of ethanol produced (on the basis of total production cost). Although not obvious in the economic summary, studies have shown reduction in production cost by coupling pre-treatment with enzymatic hydrolysis to increase ethanol yield [36].

4.3.4 SHF and SSF

By comparing the fermentation methods and assuming the costs are constant for all systems, SHF has a higher production cost at AU$59 million as compared to SSF at AU$47 million due to greater equipment costs. However, due to higher

ethanol yield in SHF, it resulted in lower ethanol cost of approximately AU$ 5.97/litre bioethanol for SHF production stage in raceway pond. A cost comparison between the different systems can be seen in Figure 4.8 and Figure 4.9 below, where ethanol cost for SSF production is approximately AU$5.93/L bioethanol.

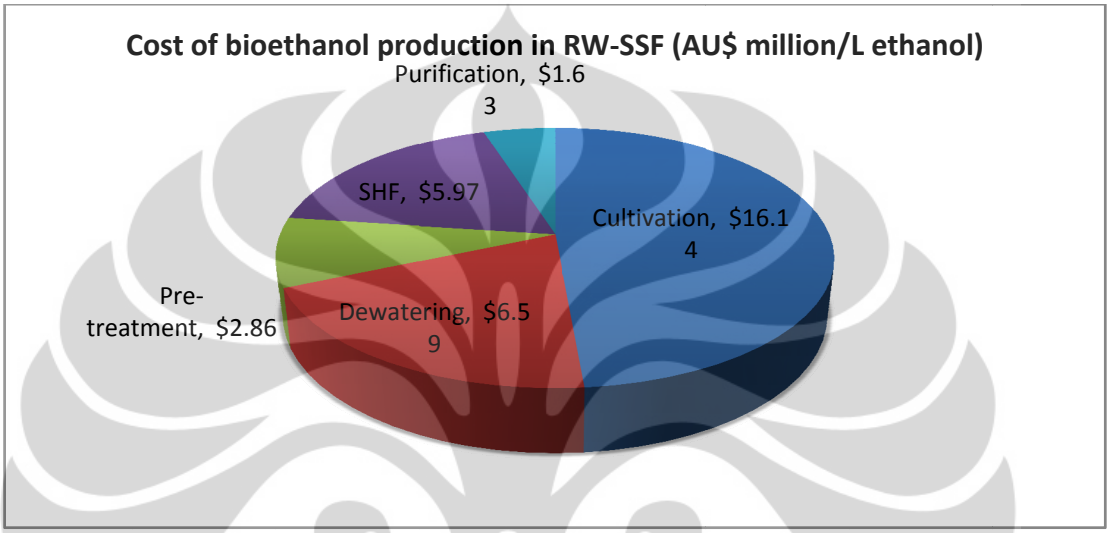


Figure 4.8 Cost of bioethanol production at each individual stages- SHF

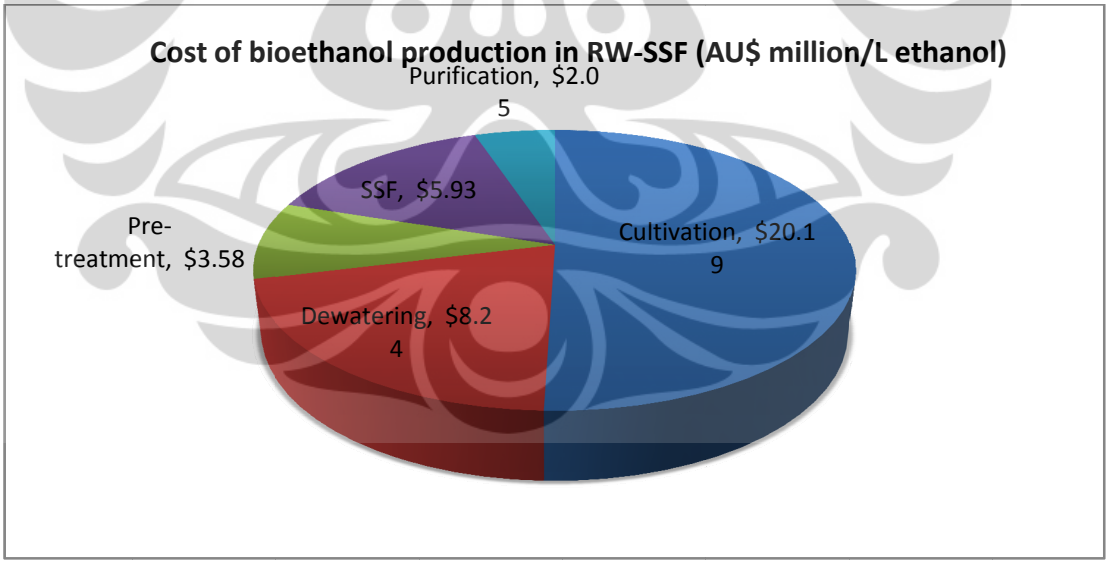


Figure 4.9 Cost of bioethanol production at each individual stages- SSF

Having said that, SSF remains the preferred option for fermentation due to the reasons such as preventing inhibitive reactions as mentioned in Section 2.6.

Unfortunately this study did not consider the effects of enzymatic inhibitions and unwanted products formation from SHF; however its impact on cost can only be deduced to incur greater loss. Ultimately, in view of the cost effectiveness of the two alternatives; taking into consideration of possible contamination, SSF would be the better option and with little difference in terms of ethanol cost, improvements on ethanol yield will undoubtedly lower its ethanol cost as a whole.

4.3.5 Purification

Product recovery in the NREL/Chem Systems studies is based on distillation technology [33]. Given that distillation is a mature technology, it has granted a lower impact on cost and energy consumption in the production of bioethanol as can be seen in Figure 4.4, contributing less than 10% of the total capital cost.

4.3.6 Overall Production Cost

Figure 4.8 and figure 4.9 illustrate the ethanol cost at each stages while the overall production cost of ethanol for all comparable systems is shown in Table 4.14. Each system is represented in terms of the overall cost per litre of bioethanol produced. With annual bioethanol produced at 9.9 million litre and 7.9 million litre in SHF and SSF, respectively, production cost of ethanol per litre ranging from AU$ 33.20 to AU$ 67 depends on the system used.

As mentioned in the earlier chapter, dual-stage dewatering is recommended for raceway pond to accommodate the large volume of medium circulating throughout the system. Centrifugation, on the other hand is more cost effective for dewatering culture from photobioreactors. SSF is a cheaper option per se; however, when looking at an entire process configuration, SHF has a total production cost lower than SSF.

Table 4.14 Summary of design process configuration

Cultivation	RW	HT	ELR	RW	HT	ELR
	$159.18	$388.78	$502.55	$159.14	$355.19	$452.65
Dewatering	Flocculation &Centrifugation			Flocculation &Centrifugation		
	$64.99	$7.10	$8.32	$64.97	$6.49	$7.49
Pre-treatment	Acid Hydrolysis			Acid Hydrolysis		
	$28.25	$23.14	$22.76	$28.25	$21.14	$20.50
Fermentation	SHF			SSF		
	$58.90	$48.25	$47.45	$46.70	$34.96	$33.89
Purification	Purification Distillation			Purification Distillation		
	$16.12	$13.21	$12.99	$16.12	$12.07	$11.70
Tot. Production cost	**$327.46**	**$480.48**	**$594.06**	**$315.18**	**$429.85**	**$526.23**
AU$/L bioethanol	**$33.20**	**$39.99**	**$48.71**	**$54.55**	**$60.23**	**$66.78**

Based on findings from table 4.14, with different technologies scheme on various stages ranging from cultivation to purification, the most cost effective process configuration is raceway pond cultivation, flocculation and centrifugation dewatering, acid hydrolysis pre-treatment, separate hydrolysis and fermentation, and purification with annual production cost of AU$ 159, 65, 28, 58, 16 million per year, respectively. The net production cost on the followed scheme is AU$ 327 million per annum, resulting lowest production price rate at AU$ 33.20 per litre of bioethanol produced among other technologies configuration.

4.4 Profitability

4.4.1.1 Revenue

As mentioned in section 4.16, the mircoalgal bioethanol production generates biethanaol, biomass residue and CO2 as product and co-product. These materials are expected to be sold to the market. Biomass residue that high in lipid content is great for biodiesel biomass. On the other hand, most of the CO_2 produced will be consumed in the cultivation hence the carbon permit can be sold. The total annual

profit of a 100% normal operation is summed in table 4.15. For bioethanol, a selling price of $100 per litre is set in order to achieve return in investment during project lifetime.

Table 4.15 Net annual profits

Product and co-product	Quantity	Selling price (AUD) / units	Revenue
Bioethanol	9.9 million litre	100 / L	$986,330,268.00
Biomass	28000 tonnes	30 / kg	$2,430,000.00
Carbon credit	81000 tonnes	25 / tonnes	$2,025,000.00
Total			$990,785,268.00

4.4.1.2 Cash Flow Analysis

Based on all economic components that have been estimated and the set revenue incomes, discounted cash flow analysis on 10 years project lifetime at interest rate at 10 % is established (figure 4.10). Income tax rate at 30 % is taken based on Australian Taxes Office on company tax rate. Details on the cash flow computation are shown in appendix C. According to the figure, the project is still under debt in the first 5 years of project time. This is due to large capital cost for project initiation, construction, and commissioning where highest expenditure burdened on the project was at the fourth year.

The key financial indicator that has been commonly used in profitability evaluation are internal rate of return (IRR), net present value (NPV), discounted pay back period, rate of return (ROR). These indicators are summarized in table 4.16.

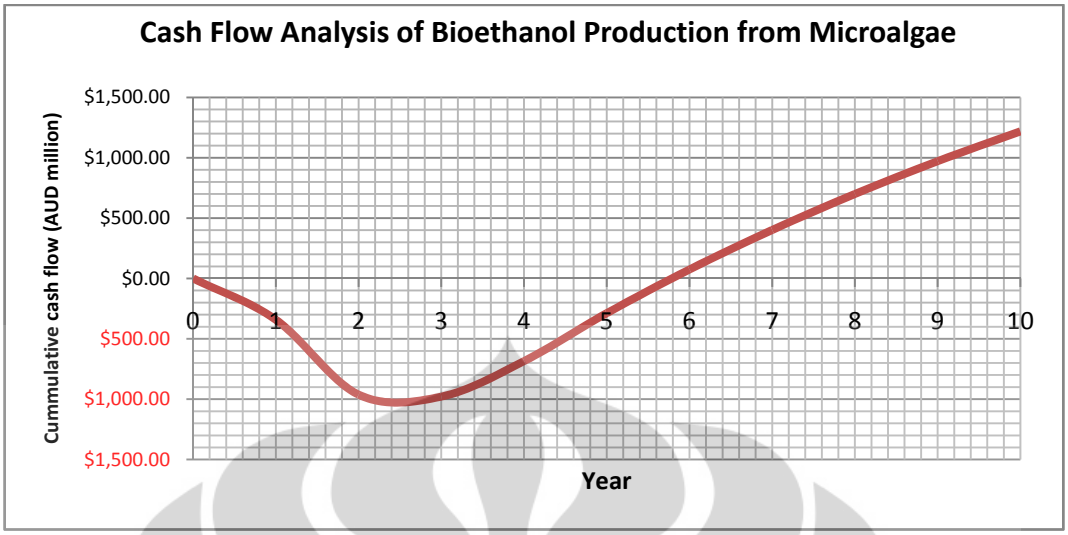


Figure 4.10 Discounted Cash Flow

Table 4.16 key indicator on profitability analysis

Interest rate	10 %
IRR	29.95 %
ROR	11 %
Discounted pay back period	5.8 years
NPV	$1,217,978,615.55

Internal rate of return point out the maximum rate that the project can afford. If IRR is lower than the minimum interest rate, the project is rejected, as there will be no return of investment received by the end of project life. IRR for this project is found at 29.95 %. This shows that the IRR is much greater that minimum interest rate hence according to this basis, the project would be accepted. Similarly, the value of rate of return from the project is expected to be greater than minimum interest rate. With ROR at 11 %, the project is favourable on this basis.

Net present value (NPV) of a project is the sum of the all discounted future cash flow, presented to the present worth. As long as NPV >0, project is desirable, which is the case for this design which has NPV at 1.2 billion dollar.

In addition, the project can be accepted if the investment will be returned before the end of project lifetime. Although 5.8 years of pay back time is rather long period for the debt to be paid off, the project is still favourable on this basis.

The price of bioethanol is set at AU$100 in order to obtained profits to cover project investment's debt and due to on high company income tax at 30%. Based on this scenario, the project would satisfy all key financial indicators. However, this selling price is relatively high for fuel market price. When selling price of bioethanol is adjusted to market price, there will be no return on investment. Therefore, the process is not feasible in economic based in profitability analysis.

4.5 Sensitivity Analysis

The sensitivity analysis was done on sales price, production rate, variable and fixed operating cost, capital investment and interest rate. The test created is based on the effect of variation on each component by ±20%. Figure 4. 11 – 4.15 show the results obtained.

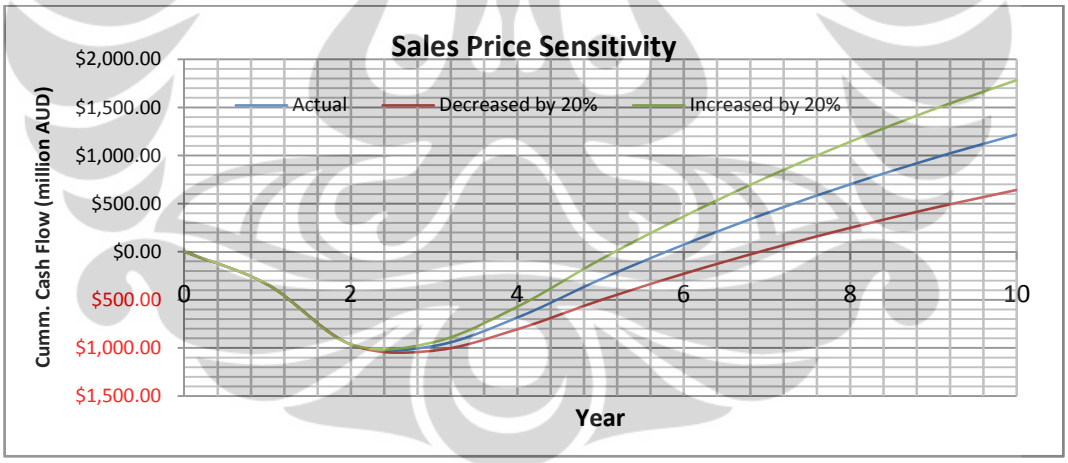


Figure 4.11 Sensitivity on sales price

Based on figure 4.11 and 4.12, variation on sales price and production rate has the same trend. As it has been expected that the payback time will be obtained much earlier when product rate and products price increase by 20 %. However, return on investment will only be covered at the year 7 if both components are decreased to 20 %.

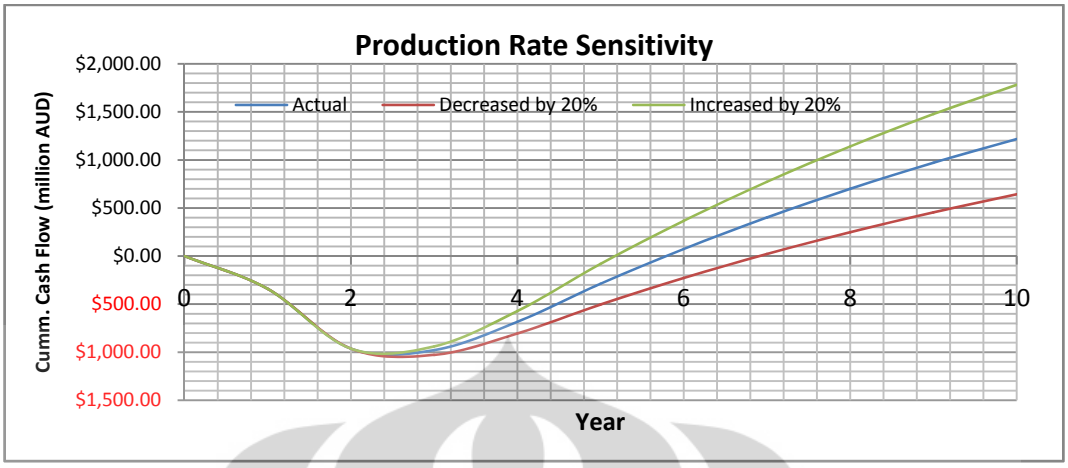


Figure 4.12 Sensitivity on production rate

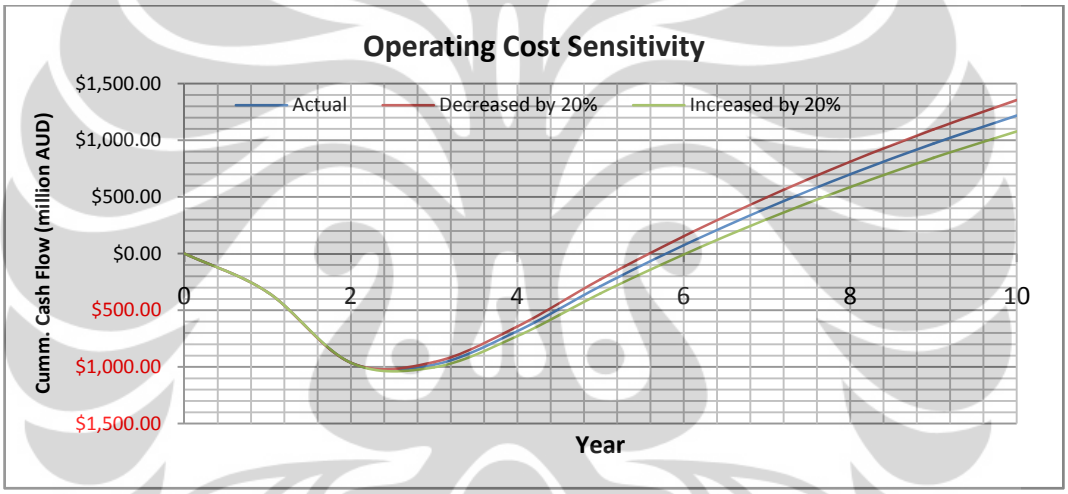


Figure 4.13 Sensitivity on operating cost

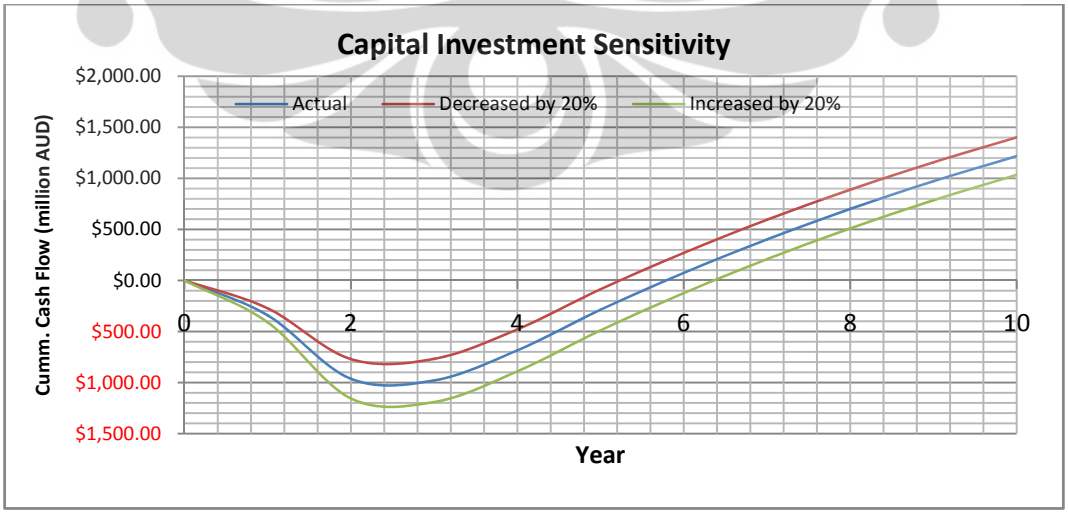


Figure 4.14 Sensitivity on capital investment

On other hand, sensitivity analysis on both variable and fixed operating cost (figure 4.13) does not show significant difference on the variant. A 20 % decrease on these components will shift the actual NPV cash flow slightly to the left.

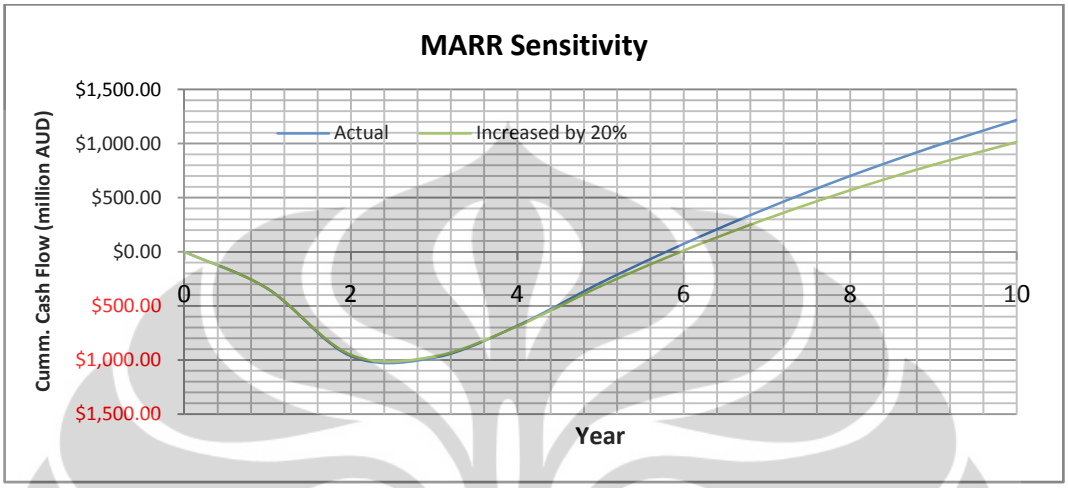


Figure 4.15 Sensitivity on MARR

Alteration on capital investment (figure 4.15) by 20 % indicates differences not only on project payback time, but also on NPV at end of project and maximum debt. If the capital cost increased by 20%, NPV is very smaller and project debt is greater, but the all parameters on cash flow are still acceptable.Figure 4.16 show that the increase on MARR by 20% will not give much changes on actual cash flow. It might affect smaller NPV but the difference is not significant.

Based on overall finding in sensitivity analysis, the cash flow is insensitive to any alteration on economic component hence the project is safe and viable.

4.6 Lipid extraction as part of bioethanol production

An addition scenario of net bioethanol production cost is to include lipid extraction in process. Briefly, Lipid extraction involves reaction with ethanol at 96 %(v/v) ethanol in agitated tank for 20 hours, where 90 % of crude oil is recovered from total oil content in the microalgae strain [32]. As the crude oil still containing both saponofiable (glucolipids and acylglycerols for biodiesel

feedstock) and unsaponifiable lipid, water and hexane is added as lipid purifying agent. From 1,515 tonnes of biomass per batch production, around 263 tonnes of crude oil is collected, while biomass residue (1,252 tonnes/ batch) still high in carbohydrate and starch is further sent to pre-treatment.

Table 4.17 Economic parameters on lipid extraction [32]

Fixed Capital Investment	$14,852,515.59
Variable Operating Cost	$92,490,381.54
Fixed Capital Year	$2,171,939.07

The economic model regarding this process is summed in table 4.17, where these values will be added to the net production cost of bioethanol. As result, the bioethanol production, in process configuration of raceway pond cultivation and separate hydrolysis and fermentation, yield at AU$ 33.71 per litre bioethanol produced. Detailed calculation can be referred to table C.6.3 in appendix.

Based on the result found, the addition of lipid extraction into bioethanol production will only affect the production cost by 50 cent per litre bioethanol produced. In term of the profitability analysis, since capital investment and operation cost involving lipid extraction only differ by roughly 1.2 % and 3 %, respectively, the cash flow parameters are also not significantly influenced.

CHAPTER 5
CONCLUSION

The production of bioethanol from microalgal biomass is viable technologically and has vast potential for continued advancements and large-scale benefits. From raw culture to bioethanol product, the production has to on go 6 processes; namely cultivation, dewatering, pre-treatment, enzyme hydrolysis, fermentation, and purification. The project lifetime is 10 years with 330 days of production, which is divided into 10 days production per cycle. The study has assessed different technology schemes throughout the processes. The technology that has been chosen is based on the most applicable methods and the most cost effective based on the production cost burdened and the bioethanol produced. Following are the process scheme.

- Cultivation and dewatering belong to biomass production stage. This study has been previously assessed by Davidson et al. [32] and data obtained has been cited appropriately. Among the three cultivation options studied, raceway pond is the most cost effective method requiring simple construction and little maintenance. The annual production cost for raceway pond (AU$ 160 million) is significantly lower as compared to photobioreactor horizontal tubular (AU$ 355 million) and external loop (AU$ 452 million), thus reinforcing its feasibility. The analysis on harvesting of microalgae shows that dual-stage dewatering with flocculation followed by centrifugation with production cost at $64 million per year is a more attractive method as the former technique greatly reduces the energy and cost requirement for centrifugation.
- Dilute acid pre-treatment approach has been studied upon repeatedly for its effectiveness to improve the enzymatic digestibility of cellulose. By improving the yield of sugar to be fermented will ultimately increase the ethanol yield or ethanol production rate, thereby reducing the overall cost of bioethanol production to $28 million per annum.
- In relation enzymatic hydrolysis and fermentation, there are two different methods evaluated; separate hydrolysis and fermentation (SHF) and

simultaneous saccharification and fermentation (SSF). SHF gives rise to an annual ethanol production cost of AU$ 59 million, whereas the SSF gives AU$ 47 million.

- Going to the purification, process has been scaled based on former study by Eden et al. [33]. This was done in order to achieve a rough figure on the production cost for product recovery. As a result, approximately AU$ 16 million of production rate per year is estimated on both study of SHF and SSF.
- The overall production cost per annum by applying SHF is found at AU$ 327 million, while SSF is lower at AU$ 315 million. Although SHF seems to exert higher production cost, it yields much greater bioethanol product at 10 million litres per annum. SSF, on other hand, produces around 8 million litre per annum. Adding to this, applying SHF into mircoalgal bioethanol production scheme is considered viable as it gives cheaper production price (AU$ 33/ litre) than SSF (AU$ 40/ litre).

Once the proposed process has been determined, it has to get through an evaluation for its feasibility in economic perspective. To investigate such criteria, test on profitability and sensitivity analysis has been assessed.

- Data obtained on profitability investigation using cumulative cash flow analysis has concluded that project on the bioethanol production from microalgae can be feasible to all key financial indicators at bioethanol selling price AU$100 per liter, biomass residue AU$ 30 per kg and carbon credit AU$ 25 per tonne. Adopting minimum interest rate 10%, the pay back period is attained at 5.8 years. Moreover, with ROR (11%) and IRR (29.95%) greater that minimum interest rate, the project is considered favourable at the respective set sales price.
- Findings from the sensitivity analysis shows that the proposed project are viable. Although changes on economic components, such as sales price, production rate, capital investment, operating cost, and MARR, is altered by ± 20%, the cash flow economic indicators will still be achieved

The overall findings from this study indicate that bioethanol production from microalgae at a scale of 50,000 tonnes per year is technologically feasible but not economically. Although the project is favourable according to sensitivity analysis, the set sales price of microalgal bioethanol at AU$ 100 per litre is extremely high and will not compete bioethanol (from other biomass) market price. Adjusting product price to market price will also cause great economic loss on the project.

Further study should be done by aiming on the cost reduction of some processes, including cultivation and dewatering cost as it exerts highest capital and production cost among all stages. Moreover, extending the project lifetime to 20-30 years is a wise view, as it allows longer time constrain on investment return. Future improvements will expect a better economic perspective on microalgal bioethanol production to be more competitive to available biofuel in the market.

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APPENDIX A

Appendix A.1. Technology Selection

Table A.1.1 Protein and carbohydrate content in various algae strain[4]

Algae strains	Proteins	Carbohydrates
Scenedesmus obliquus	50–56	10–17
Scenedesmus quadricauda	47	-
Scenedesmus dimorphus	8–18	21–52
Chlamydomonas rheinhardii	48	17
Chlorella vulgaris	51–58	12–17
Chlorella pyrenoidosa	57	26
Spirogyra sp.	6–20	33–64
Dunaliella bioculata	49	4
Dunaliella salina	57	32
Euglena gracilis	39–61	14–18
Prymnesium parvum	28–45	25–33
Tetraselmis maculate	52	15
Porphyridium cruentum	28–39	40–57
Spirulina platensis	46–63	8–14
Spirulina maxima	60–71	13–16
Synechoccus sp.	63	15

Table A.1.2 Flocculation efficiency in different algae strain [5]

Algae Class	Algal Species	Cell Density (cells / mL)	Flocculation Method	Flocculation Efficiency (%)
Cryptophyceae	R. salina	(1-2) x 10^6	pH	85-90
Bacillariophyceae	A.septentrionalis	(1-1.5) x 10^6	Ferric & pH	85-95
	C. calcitrans	(1-2) x 10^7	pH	95-99
	C. mulelleri	(2-3) x 10^6	pH	95-97
	N. closterium	(1-2) x 10^6	pH	90-95
	T. pseudonana	(2-4) x 10^6	Ferric & pH	80-95
	Skeletonema sp.	(3-6) x 10^6	pH	95-98
Eustigmatophyceae	N. oculate	(1-2) x 10^7	pH	<30
Prasinophyceae	T. suecica	(4-8) x 10^5	pH	85-95
Prymnesiophyceae	Isochrysis sp.	(3-4) x 10^6	Ferric & pH	<30

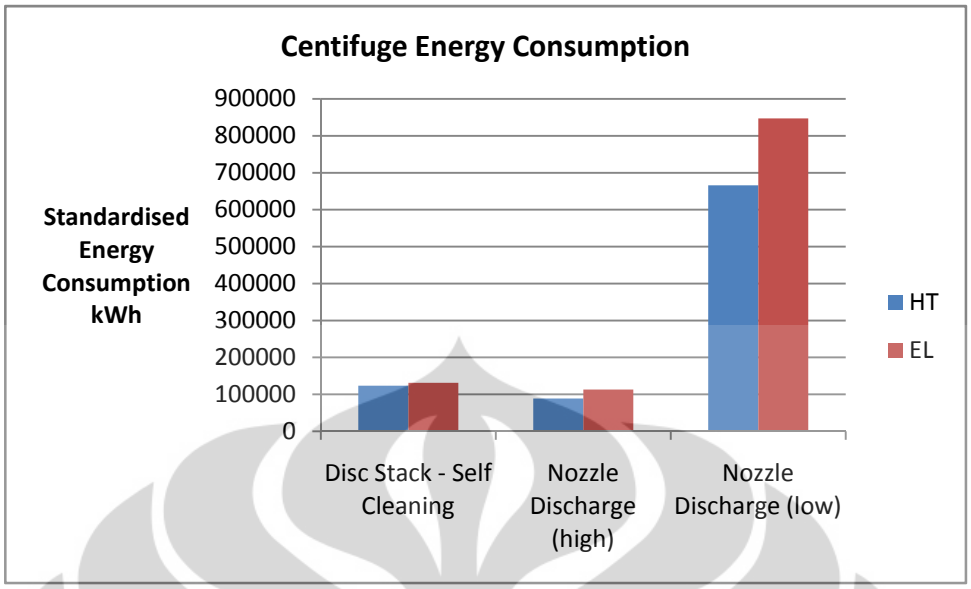

Figure A.1.1 Energy Consumption for the Best Standalone Centrifugation Dewatering Options [5]

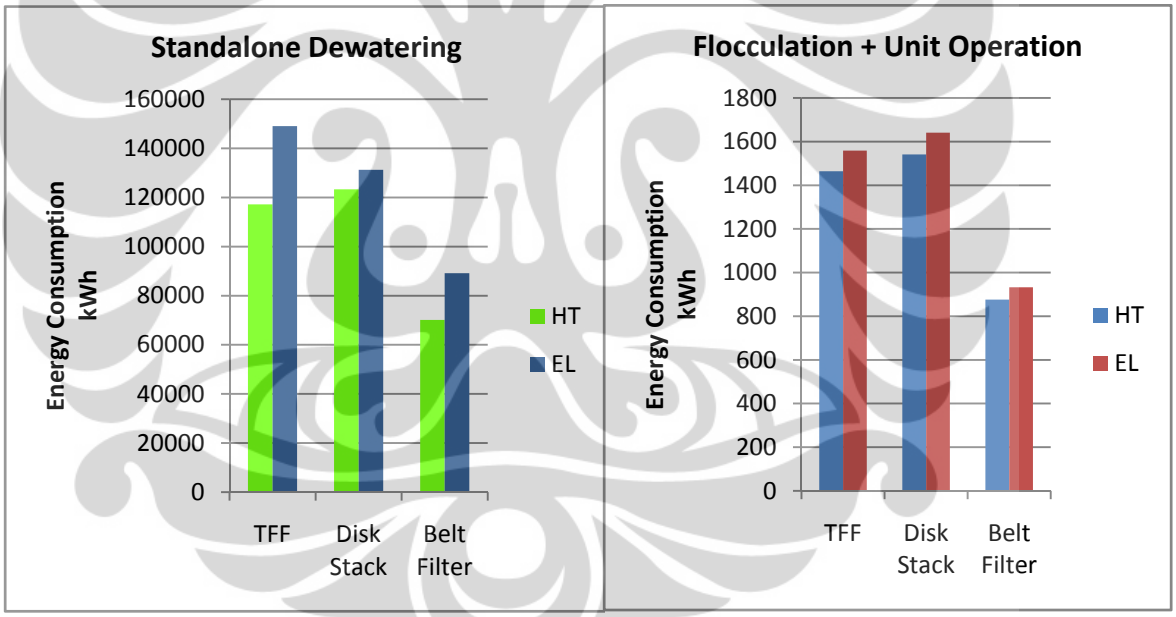

Figure A.1.2 Energy requirement in standalone dewatering (right) and dual-stage dewatering with flocculation (left)[5]

Table A.1.3: Capabilities of algae strains to be concentrated by Chitosan[5]

Species	Optimum chitosan dosage ($mg\,L^{-1}$)	Culture pH	pH after chitosan addition	Cell recovery (%)
Chaetoceros muelleri	150	8.06	5.03	95
Chaetoceros calcitrans	80	7.29	5.27	80
Skeletonema costatum	80	8.66	5.42	70
Thalassiosira pseudonana	40	8.29	6.31	90
Tetraselmis chui	40	7.69	6.03	80
Pavlova lutheri	80	7.28	5.30	80
Tahitian *Isochrysis*	40	7.43	6.26	90

Table A.1.4. Algae Drying Methods [6]

Method	Advantages	Limitations	Remarks
Drum-drying	Fast and efficient	Cost intensive	Ruptures cellulosic cell walls, sterilises the product, not suitable for *Spirulina*
Spray-drying	Fast and efficient	Cost intensive	Sterilises the product, breakage of cellulosic cell walls not always guaranteed
Sun-drying	Very low fixed capital and no running costs	Slow process, weather dependent	Biomass may ferment, sterilisation not possible, does not break cellulosic cell walls
Solar-drying	Low capital cost	Weather dependent	Does not break cellulosic cell walls, sterilisation not possible
Cross-flow-drying	Faster than sun- and solar- drying, cheaper than drum-drying	Requires electricity	Does not break cellulosic cell walls, sterilisation not possible
Vacuum-shelf drying	Gentle process	Cost intensive	Does not break cellulosic cell walls, product becomes hygroscopic, sterilisation not possible, preserves cell constituents
Freeze-drying	Gentle process	Slow process, Cost intensive	Does not break cellulosic cell walls, sterilisation not possible, preserves cell constituents

Table A.1.5 Summary of different pre-treatment method on corn stover [7]

Method	Catalyst	Time / Temp (°C)	Glucose yield (%) / Xylose yield (%)
Alkali	Lime ($Ca(OH)_2$)	4 weeks/ 55	92.0/52.8
Dilute Acid-1	0.49 % H_2SO_4	20 min/ 160	91.6/ 91.2
Dilute Acid-2	5 % H_2SO_4	90 min/ 120	54.6/100
AFEX*	Concentrated NH_3	5 min/ 90	96.0/ 77.7
ARP*	NH_3	10 min/ 170	90.0/ 41.1
Steam-1*	H_2SO_4	5 min/190	73.6/61.0
Steam-2*	SO_2	5 min/ 190	90.0/84.0

*These methods are not discussed in this work, as it is generally focus on lignocellulosic feedstock (corn stover). The result data is presented for comparison purposes only.

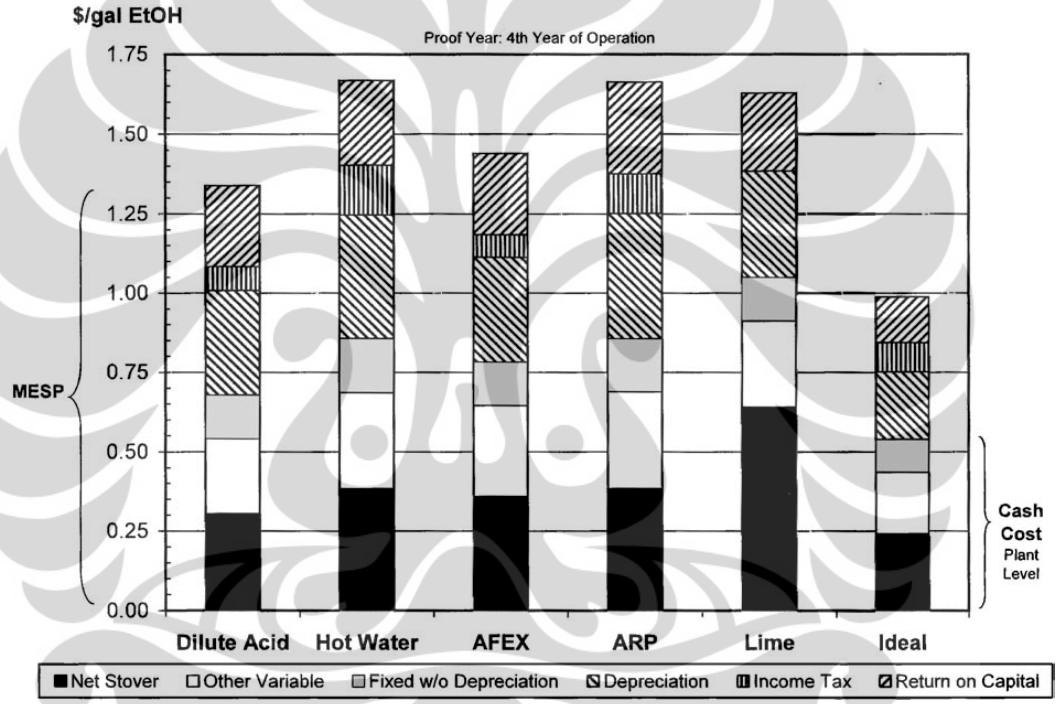

Figure A.1.6 Cost of different pre-treatment on corn stover [8]

Appendix A.2. Process Flow Diagram

Appendix A.2.1 Cultivation: Raceway Pond

Appendix A.2.2 Cultivation: External Loop Photobioreactor

Appendix A.2.3 Cultivation: Horizontal Tubular Photobioreactor

Appendix A.2.4 Dewatering: Flocculation and Centrifugation

Appendix A.2.5 Pre-treatment: Acid Hydrolysis

Appendix A.2.6 Separate Hydrolysis and Fermentation

Appendix A.2.7 Simultaneous Saccharification and Fermentation

Appendix A.2.8 Purification: Beer Column, Rectifying Column, Molecular sieve

Appendix A.2.9 Purification: Scrubber

Appendix A.3. Design Database

Equipment tag	Process Equipment	Design capacity
PT BC-1	Pneumatic solids conveying-belt conveyor - biomass	1.5 kg/s at 20 m distance
PT T-1	Water storage tank	1515 m^3
PT T-2	Sulphuric acid storage tank	770 m^3
PT E-1	Jacketed acid PT Reactor	185 m^3
PT P-1	Pump-sulphuric acid inlet	centrifugal, 384 m^3/hr
PT P-2	Pump-water inlet	centrifugal, 1530 m^3/hr
PT P-3	Pump- water for steam	centrifugal, 152 m^3/hr
PT P-4	Pump- blowdown tank inlet	centrifugal, 1592 m^3/hr
PT H-1	Water heater for steam	Direct fired heater, 5900KW
PT E-2	Blow down tank	110 m^3
PT P-5	Pump- S/L separator inlet	centrifugal, 743 m^3/hr
PT E-3	Centrifuge	solid bowl , SS, 750 KW, 3.2 kg/s solid
PT P-6	Pump- Neutralisation tank inlet	715 m^3/hr
PT E-4	Neutralisation Reactor tank	235 m^3
PT BC-2	Pneumatic solids conveying-belt conveyor - NaOH	20m distance
PT P-7	Pump- to E.H/SSF storage	centrifugal, 715 m^3/hr
PT P-8	Pump to EH /SSF	Centrifugal, 1788 m^3/hr
PT T-3	Storage tank	18600 m^3
PT A-1	Pre-treatment tank agitator	29.55 W/m^3
PT A-2	Neutralisation tank agitator	98.5 W/m^3
SHF E-3	Enzyme bioreactor tank	1600 m^3
SHF BC-2	Pneumatic solids conveying equipment - cellulase	
SHF BC-3	Pneumatic solids conveying equipment - amylase	
SHF P-2	Pump- Biomass outlet to fermentation	centrifugal, 1788 m^3/hr
SHF HX-1	Cooler- Bioethanol inlet to fermentation	185 m^3/ hr
SHF E-1	Jacketed yeast fermentation tank	3.2 m^3
SHF E-2	Jacketed fermentation tank	1600 m^3
SHF T-1	Ethanol storage tank	5600 m^3
SHF BC-1	Pneumatic solids conveying equipment - NaOH	
SHF P-3	Pump- Ethanol outlet to ethanol tank	Centrifugal, 2682m^3/hr
SHF P-4	Pump- Ethanol outlet to purification	Centrifugal, 2682m^3/hr
SHF P-1	Pump- water for steam	Centrifugal, 12m3/hr)
SHF A-1	Yeast fermentation agitator	0.014 KW
SHF A-2	Ethanol fermentation agitator	53.95 KW
SHF A-3	Enzyme hydrolysis agitator	53.95 KW
SHF H-1	Water heater for steam	Direct fired heater, 6135KW, CS, 202 kPa
SSF E-1	Jacketed yeast fermentation tank	3.2 m^3

SSF E-2	Jacketed fermentation tank	1600 m^3
SSF T-1	Ethanol storage tank	1600 m^3
SSF BC-1	Pneumatic solids conveying equipment - cellulase	
SSF BC-2	Pneumatic solids conveying equipment - amylase	
SSF P-2	Pump- fermentation tank outlet	Centrifugal, 2682 m^3/hr
SSF P-3	Pump- ethanol storage tank outlet	Centrifugal, 2682 m^3/hr
SSF P-1	Pump- water for steam	Centrifugal, 12 m^3/hr
SSF H-1	Water heater for steam	Direct-fired heater, 6135KW, CS, 202 kPa
SSF A-1	Yeast fermentation agitator	0.01 KW
SSF A-2	Fermentation tank agitator	53.95 KW
PR E-1	Beer column	32 trays, 4.37 m diameter, 2atm
PR E-2	Rectifying Column	0.29 m dia. Rect, 0.1 m dia. Strip. 60trays
PR HX-1	Beer column feed HX-1	Shell and tube
PR HX-2	Beer column feed HX-2	plate frame
PR HX-4	Beer column reboiler	101407 KW
PR HX-6	Rectifying column reboiler	10362 KW
PR HX-3	Beer column condenser	5547 KW
PR HX-5	Rectifying column condenser	44785 KW
PR E-3 E-4	Mol sieve (9 pieces)	
PR E-5	Vent scrubber	Plastic packings, 7.6 m, 4 theoretical stages
PR P-1	Beer column bottom pump	Centrifugal, 1146 m^3/hr
PR P-2	Beer column reflux pump	Centrifugal, 2.72 m^3/hr
PR P-3	Rectification column bottoms pump	Centrifugal, 34.92 m^3/hr
PR P-4	Rectification column reflux pump	Centrifugal, 99.11 m^3/hr
PR P-4	Scrubber bottoms pump	Centrifugal, 15.64 m^3/hr
PR DR-1	Beer column reflux drum	Centrifugal, 1.30 m^3/hr
PR DR-2	Rectification column reflux drum	Centrifugal, 49.61 m^3/hr

Appendix A.4. Process Design Calculation

Appendix A. 4. 1 Cultivation Calculations and Data

Basis Calculation: 50000 tonnes of microalgae biomass per year
Assumptions:

- Facility operates 330 days per year
- Various dilution rates are considered for different reactor configurations. Each reactor is designed to produce 151.5tonne of dry weight biomass per day (ie. 50000/330=151.5tonne/d)
- Only 80% of biomass is harvested from the reactors at any given time

Table A.4.1.1 Comparison of photobioreactor and raceway production methods[32]

Variable	Horizontal Tubular Reactor	External Loop Reactor	Raceway Pond
Annual Biomass Production (t)	50000	50000	50000
Biomass Required per Batch (t)	757.5	568.75	947.5
Biomass Extracted per batch (t)	606	455	758
Biomass Concentration (kg/m3)	4.525	3.8	0.585
Dilution Rate (1/d)	0.25	0.333333333	0.2
Area Required per cultivation unit (m^2)	947*	12*	1050*
Area Per Unit (m^2)	1263	16	1400
Total Cultivation Area (m^2)	5284365*	8980263*	8098291*
Total Area (m^2)	7047680	11973684	10797721
Total Cultivation Area (ha)	528	898	810
Total Area (ha)	705	1197	1080
No. Of Units Required	5580	748355	7713
Unit Description	132 parallel tubes; tube length of 80m. Tube diameter 0.06m	4m airlift, 80m tubular run; Tube diameter 0.06m	1050m^2 pond; 14m wide and 75m long; 0.2m depth.
Total Tubing Length(m)	58925967	59868421	N/A
Cultivation Areal Productivity ($kg/m^2.d$)	0.036	0.021	0.023
Total Areal Productivity ($kg/m^2.d$)	0.027	0.016	0.018
Volumetric Productivity ($kg/m^3.d$)	1.131	1.267	0.117
Volume per Cultivation unit (m^3)	30	0.2	210
Total Volume (m^3)	167403	149671	1619658
Approximate Annual CO_2 Consumption (t)	92000	92000	92000^
Energy Dissipation (W/ m^3)	60-170	60-170	
Energy Dissipation (kWh/ per unit)			3.24~

~ Energy consumption accounts for the paddle wheel and all pumping.
# Assuming 95% CO_2 capturing efficiency.
^ Assuming 90% CO_2 utilisation efficiency.

* ¾ of the total area of all cultivation units is taken up by either the tubing or pond area.

Appendix A. 4. 2 Dewatering Calculations and Data

Table A.4.2.1 Concentration Flocculants required for dewatering [5]

Flocculation Method	Efficiency	Efficiency (%)	Specific Flocculants Added	Amount of Flocculants Added
pH& Polyelectrolyte	High	90	LT-25 & NaOH	0.5 mg/L & 1M diluted 1:2 (per 500 L batch)
Ferric Chloride & Polyelectrolyte	High	80	LT-25 & FeCl3.6H2O	0.5 mg/L & 0.1M (per 500 L batch)
Chitosan	High	95	Chitosan	150mg/L
Chitosan	High	90	Chitosan	40mg/L
pH& Polyelectrolyte	Low	30	LT-25 & NaOH	0.5 mg/L & 1M diluted 1:2 (per 500 L batch)
Ferric Chloride & Polyelectrolyte	Low	30	LT-25 & FeCl3.6H2O	0.5 mg/L & 0.1M (per 500 L batch)
Chitosan	Low	70	Chitosan	80mg/L

Table B.4.2.2 Flocculation Data for Horizontal Tubular Reactor [5]

Flocculant	Reduction In Culture Volume(m3)	Volume Remaining (m3)	New Concentration (kg/m3)	Concentration Factor Required
pH& Polyelectrolyte	120530.2	13392.2	36.2	13.8
Ferric Chloride & Polyelectrolyte	107137.9	26784.5	18.1	27.6
Chitosan	127226.3	6696.1	72.4	6.9
Chitosan	120530.2	13392.2	36.2	13.8
pH& Polyelectrolyte	40176.7	93745.7	5.2	96.7
Ferric Chloride & Polyelectrolyte	40176.7	93745.7	5.2	96.7
Chitosan	93745.7	40176.7	12.1	41.4

Table A.4.2.3 Flocculation Data for External Loop Tubular Reactor[5]

Flocculants	Reduction In Culture Volume(m3)	Volume Remaining(m3)	New Concentration (kg/m3)	Concentration Factor Required
pH& Polyelectrolyte	107763.1	11973.7	30.4	16.4
Ferric Chloride & Polyelectrolyte	95789.4	23947.4	15.2	32.9
Chitosan	113750.0	5986.8	60.8	8.2
Chitosan	107763.1	11973.7	30.4	16.4
pH& Polyelectrolyte	35921.0	83815.8	4.3	115.1
Ferric Chloride & Polyelectrolyte	35921.0	83815.8	4.3	115.1
Chitosan	83815.8	35921.0	10.1	49.3

Table A.4.2.4 Flocculation Data for Raceway Pond[5]

Flocculants	Reduction In Culture Volume(m3)	Volume Remaining(m3)	New Concentration (kg/m3)	Concentration Factor Required
pH& Polyelectrolyte	1166153.8	129572.6	4.7	106.8
Ferric Chloride & Polyelectrolyte	1036581.2	259145.3	2.3	213.7
Chitosan	1230940.2	64786.3	9.4	53.4
Chitosan	1166153.8	129572.6	4.7	106.8
pH& Polyelectrolyte	388717.9	907008.5	0.7	747.9
Ferric Chloride & Polyelectrolyte	388717.9	907008.5	0.7	747.9
Chitosan	907008.5	388717.9	1.6	320.5

Table A.4.2.5 Centrifugation Data for Horizontal Tubular Reactor [5]

Concentration Factors Required For Second Stage Dewatering After Flocculation							
		High	High	High	High	Low	Low
		LT-25 & NaOH	LT-25 & FeCl3.6H2O	Chitosan	Chitosan	LT-25 & NaOH	Chitosan
Concentration After Flocculation (kg/m3)		36.2	18.1	72.4	36.2	5.2	12.1
Concentration Factor Required for Second Stage Centrifugation							
Conc. Factor							
Disc Stack - Self Cleaning	120.0	0.1	0.2	0.1	0.1	0.8	0.3
Nozzle Discharge (high)	150.0	0.1	0.2	0.0	0.1	0.6	0.3
Nozzle Discharge (low)	20.0	0.7	1.4	0.3	0.7	4.8	2.1
Decanter Bowl	11.0	1.3	2.5	0.6	1.3	8.8	3.8
Standardised Energy Consumption after Flocculation							
		High	High	High	High	Low	Low
		LT-25 & NaOH	LT-25 & FeCl3.6H2O	Chitosan	Chitosan	LT-25 & NaOH	Chitosan
Conc Factor After Flocc.		36.2	18.1	72.4	36.2	5.2	12.1
Culture Volume After Flocc.		13392.2	26784.5	6696.1	13392.2	93745.7	40176.72
Centrifugation Energy Consumption (kWh)							
Disc Stack - Self Cleaning		1541.5	6165.9	385.4	1541.5	75531.7	13873.2
Nozzle Discharge (high)		1109.9	4439.4	277.5	1109.9	54382.9	9988.7
Nozzle Discharge (low)		8323.9	33295.6	2081.0	8323.9	407871.4	74915.2
Decanter Bowl		134527.8	538111.1	33631.9	134527.8	6591861.0	1210750.

Table A.4.2.6 Centrifugation Data for External Loop Tubular Reactor [5]

Concentration Factors Required For Second Stage Dewatering After Flocculation							
		High	High	High	High	Low	Low
		LT-25 & NaOH	LT-25 & FeCl3.6H2O	Chitosan	Chitosan	LT-25 & NaOH	Chitosan
Concentration After Flocc	(kg/m3)	30.4	15.2	60.8	30.4	4.3	10.1
Concentration Factor Required for Second Stage Centrifugation							
Conc. Factor							
Disc Stack - Self Cleaning	120.0	0.1	0.3	0.1	0.1	1.0	0.4
Nozzle Discharge (high)	150.0	0.1	0.2	0.1	0.1	0.8	0.3
Nozzle Discharge (low)	20.0	0.8	1.6	0.4	0.8	5.8	2.5
Decanter Bowl	11.0	1.5	3.0	0.7	1.5	10.5	4.5
Standardised Energy Consumption after Flocculation							
		High	High	High	High	Low	Low
		LT-25 & NaOH	LT-25 & FeCl3.6H2O	Chitosan	Chitosan	LT-25 & NaOH	Chitosan
Conc Factor After Flocc.		30.4	15.2	60.8	30.4	4.3	10.1
Culture Volume After Flocc.		14967.1	29934.2	7483.6	14967.1	104769.7	44901.3
Centrifugation Energy Consumption (kWh)							
Disc Stack - Self Cleaning		1641.1	6564.5	410.3	1641.1	80415.3	14770.2
Nozzle Discharge (high)		1181.6	4726.5	295.4	1181.6	57899.0	10634.5
Nozzle Discharge (low)		8862.1	35448.4	2215.5	8862.1	434242.8	79758.9
Decanter Bowl		143225.8	572903.3	35806.5	143225.8	7018066.0	1289032.5

Table A.4.2.7 Centrifugation Data for Raceway Pond [5]

Concentration Factors Required For Second Stage Dewatering After Flocculation							
		High	High	High	High	Low	Low
		LT-25 & NaOH	LT-25 & FeCl3.6H2O	Chitosan	Chitosan	LT-25 & NaOH	Chitosan
Concentration After Flocc	(kg/m3)	4.7	2.3	9.4	4.7	0.7	1.6
Concentration Factor Required for Second Stage Centrifugation							
Conc. Factor							
Disc Stack - Self Cleaning	120.0	0.9	1.8	0.4	0.9	6.2	2.7
Nozzle Discharge (high)	150.0	0.7	1.4	0.4	0.7	5.0	2.1
Nozzle Discharge (low)	20.0	5.3	10.7	2.7	5.3	37.4	16.0
Decanter Bowl	11.0	9.7	19.4	4.9	9.7	68.0	29.1
Standardised Energy Consumption after Flocculation							
		High	High	High	High	Low	Low
		LT-25 & NaOH	LT-25 & FeCl3.6H2O	Chitosan	Chitosan	LT-25 & NaOH	Chitosan
Conc Factor After Flocc.		4.7	2.3	9.4	4.7	0.7	1.6
Culture Volume After Flocc.		129572.6	259145.3	64786.3	129572.6	907008.5	388717.9
Centrifugation Energy Consumption (kWh)							
Disc Stack - Self Cleaning		115360.3	461441.1	28840.1	115360.3	5652653.0	1038242
Nozzle Discharge (high)		83059.4	332237.6	20764.8	83059.4	4069910.1	747534.5
Nozzle Discharge (low)		622945.4	2491781.7	155736.4	622945.4	30524326.1	5606509
Decanter Bowl		10067804.9	40271219.8	2516951.2	10067804	493322442.0	90610245

Appendix A. 4. 3 Pre-treatment Calculations and Data

Basis: 50000 tonnes of microalgal dry biomass per year

Acid pretreatment

Assumptions:

- Each batch deliver 1515151 kg of dry biomass to pre-treatment for 10 days production
- 100 % of carbohydrate and starch content in the biomass is digested by acid hydrolysis
- The biomass entering acid pre-treatment tank is spitted into cycles.
- Each cycles has 63131 kg/ hr of dry biomass

Table A.4.3.1 Pre-treatment tank PT E-1 conditions

Acid concentration	0.5 % v/v
Biomass loading	10 kg/m^3
Pressure	6 atm
Temperature	160 °C
Residence time	15 min
pH	2.93
No. of cycle	24
Biomass per cycle	6313.1 kg/hr

Total volume of water required per batch = 1515151 kg / (10 kg/m^3) = 151515 m^3
Total volume of sulphuric acid used = 0.05 * 151515 m^3 = 761.38 m^3
Density of Sulphuric acid = 1840 kg/ m^3
Total volume = 152276.53 m^3

Mass component of *Chlorococum sp*. :

Table A.4.3.2 Monomeric sugars converted from cellulose per batch

Component	Composition (kg)
Total carbohydrate	**850910**
Xylose	144546
Mannose	73788
Glucose	230606
Galactose	43748
Starch	**171515**

Component balance in acid pre-treatment tank PT E-1:

Table A.4.3.3 Component balance in PT E-1

	IN	OUT
	mass (kg)	mass (kg)
Biomass*	1515151.51	850909.09
Xylose*	-	144545.45
Mannose*	-	73787.88
Glucose*	-	230606.06

Galactose*	-	43787.88
Starch*	-	171515.15
water	150909090.40	150909090.40
acid	1400944.11	1400944

*mass component that is dissolved in acid solution

Acid hydrolysis vessel sizing:

Table A.4.3.4 Sizing of Pre-treatment tank PT E-1*

Biomass input per cycle	6313.1 kg/ hr
Volume of water and sulphuric acid	634.5 m^3
Size of vessels	185
No. of vessels	6
No. of Continuous cycles	240

*The condition of Pre-treatment shown is based on one cycle dry biomass input per batch.

Blow down tank

Assumption:

- The input is based on 6313.1 kg/ hr of dry biomass per cycle
- 26.2%w/w of steam has been removed to atmosphere pressure
- residence time 10 minutes

Table A.4.3.5 Component balance of blow down tank PT E-2 per cycle of entering biomass

Each cycle	IN	OUT	
		Vapor (100 C)	Liquid hydrolysate
	mass (kg)	mass (kg)	mass (kg)
Biomass*	5318.18	-	5318.18
Complex sugar*	4151.52	-	4151.52
Sulfuric acid	8756	-	8756
Water	943181.815	218818.1811	724363.6339

*mass component that is dissolved in solution

Table A.4.3.6 Component balance of blow down tank PT E-2 in 24 cycles or a batch of entering biomass

Each batch	IN	OUT	
		Vapor (100 C)	Liquid hydrolysate
	mass (kg)	mass (kg)	mass (kg)
Biomass*	850909.09	-	850909.09
Complex sugar*	664242.42	-	664242.42
Sulfuric acid	1400944	1400944	1400944
Water	150909090.40	35010908.97	115898181.42

*mass component that is dissolved in solution

S-L separation

Assumption:

- All of inhibitors generated after acid pre-treatment are removed
- 95 % of the solid material (Biomass residue) is separated
- Residence time = 30 min

Table A.4.3.7 Component balance of centrifuge PT E-3 per cycle

each cycle	IN	OUT (95% separation)	
		S	**L**
	mass (kg)	mass (kg)	mass (kg)
Biomass*	5318.18	5318	0
Complex sugar*	4151.52	0	4152
water	943181.81	36218	8318
Sulfuric acid	8756	438	688145

Table A.4.3.8 Component balance of centrifuge PT E-3 per batch

Each batch	IN	OUT (95% separation)	
		S	**L**
	mass (kg)	mass (kg)	mass (kg)
Biomass*	850909.09	850909.09	0
Complex sugar*	664242.42	0	664242.42
water	115898181	5794909	110103272
Sulfuric acid	1400944	70047	1330897

Total solid loading in centrifuge = 3.28 kg/s of solid
Number of centrifuge required = 3
Volume inlet per batch = 117125.02 m^3
Volume out solid = 5856.25 m^3 (5818 m^3 water, 38.07 m^3)
Volume out liquid = 111268.77 m^3 (110545 m^3 water, 723.31 m^3)

Neutralisation

Assumption:

- pH is increased to 5 by adding NaOH pellets
- residence time = 1 hour

Initial H_2SO_4 concentration = (1.840 kg/L) / (98 g/mol) * (723.31/ 111268) = 0.122 M

	Initial condition (kM)	Final condition (kM)
pH	3.61	5.00
pOH	10.39	9.00
[H+]	2.44 E-04	1.00 E-05
[OH-]	4.10 E-11	1.00 E-09

NaOH added = (1.00 E-09 - 4.10 E-11) * 111. 268 L/ (40 g/L) = 4.451 kg

Neutralisation reaction occurred:

H2SO4 + 2NaOH = Na2SO4 +2 H2O			
H2so4 (kmol)	2NaOH (solid) (kmol)	Na2SO4 buffer (kmol)	H2O (kmol)
13580.5807	0.1113		
0.11	0.1113	0.1113	0.2225
13580.4694	**0.0000**	0.1113	0.2225

[Na_2SO_4] in solution = 0.1113 kmol / 111.268 L = 1 uM

Table B.4.3.9 Component balance of neutralisation tank PT E-4 per cycle

Each cycle	IN		OUT	
	vol (m3)	mass (kg)	vol (m3)	mass (kg)
Complex sugar	-	4152	-	4152
Liquid hydrolysate	695	-	-	-
NaOH (powder)	-	0.028	-	-
1uM Na_2SO4 in hydrolysate	-	-	695	-

Table A.4.3.10 Component balance of neutralisation tank PT E-4 per batch

Each batch	IN		OUT	
	vol (m3)	mass (kg)	vol (m3)	mass (kg)
Complex sugar	-	664242.42	-	664242.42
Liquid hydrolysate	111268.77	-	-	-
NaOH (powder)	-	4.451	-	-
1uM Na_2SO4 in hydrolysate	-	-	111268.77	-

Table A.4.3.11 Sizing of Neutralisation tank PT E-4

Volume of liquid hydrolysate	695 m^3
Size of vessels (incl. headspace)	235 m^3
No. of vessels	3
No. of Continuous cycles	240

Hydrolysate storage tank

Assumption: half of the total amount of the hydrolysate in watch batch is collected in storage tank PT T-3. The other half is sent to SHF E-3 or SSF E-2

Hydrolysate storage tank capacity = 18600 m^3

No. ofHydrolysate storage tank required = 3

Appendix A. 4. 4 Enzyme Hydrolysis Calculations and Data

Assumption:

- 4 train process
- Each process carries half amount of hydrolysate from pre-treatment in a batch
- Conversion of complex sugar to simple sugar is 90 %

Table A.4.4.1 Total sugar converted in E. H tank SHF E-3 per batch

A batch – in 2 trains	IN (kg)	OUT (kg)	
		Converted	Unconverted
Complex sugar - carbohydrate			
hexose	348181	313363	34818
pentose	144546	130091	14455
Complex sugar - starch	171515	154364	17152

Table A.4.4.2 Enzyme Hydrolysis tank condition SHF E-3

Temperature	40 °C
pH	5
Residence time	72 hour
Cellulase loading	0.02 g/g substrate
Amylase loading	0.01 g/g substrate
Approximate sugar converted	90 %

Table A.4.4.3 Sizing of Enzyme hydrolysis vessel SHF E-3 per train

Total hydrolysate volume	27817m^3
Size of vessels	1600 m^3
No. of vessels	20
No. of Continuous train	4

Cellulose preparation

Concentration of complex sugar per train = 5.97 kg/ m^3

Concentration of complex sugar per tank = 0.298 kg/ m^3

Cellulose loading = 0.02 kg/kg substrate

Cellulose required = 0.02 kg/kg substrate * 0.298 kg/ m^3 *27817m^3 = 166 kg

Amylase preparation

Concentration of starch per train = 1.54 kg/ m^3

Concentration of starch per tank = 0.08 kg/ m^3

Amylase loading = 0.01 kg/kg substrate

Amylase required = 0.01 kg/kg substrate * 0.08 kg/ m^3 *27817m^3 = 85 kg

Appendix A. 4.5 SHF Calculations and Data

Assumption:

- 4 train process
- Each process carries half amount of hydrolysate from pre-treatment in a batch
- Yeast *Saccharomyces cerevisae*only ferment hexose sugar
- 90 % of hexose sugar is converted to ethanol
- pH is incread to 7 by adding NaOH

Yeast preparation

Table B.4.5.1 Medium and supporting medium concentration in yeast seeding tank (SHF E-1)

Yeast loading	5 g/L
Ammonium sulfate	1 g/L
Ammonium chloride	2 g/L
Magnesium sulfate	0.8 g/L
LB broth	3 %v/v

Amount of Ammonium chloride per train = 1600 gram
Amount of potassium chloride per train = 800 gram
Amount of magnesium sulphate per train = 640 gram
Amount of yeast per train = 1000 gram
Amount of LB broth per train = 33 L
Vol of water required = 1000 mL (1L of distilled water per 1 kg of yeast)

Table A.4.5.2 Yeast fermentation tank condition SHF E-1

Size of vessel	3.2 m^3
No. of tank required	1
No. of train	4
Temperature	37°C
Residence time	1 hour
Condition	Aerobic

Fermentation

Ethanol conversion (in mass kg)

	$C_6H_{12}O_6$	2 C_2H_5OH	2 CO_2
100% conversion	517000	264244	252756
90% conversion	465300	237820	248160
residue	31020		

Initial pH = 5 ; $[H^+]$= 10e-5 $[OH^-]$= 10e-9
Final pH = 7 ; $[H^+]$= 10e-7 $[OH^-]$ = 10e-7
NaOH added = 200 gram

Table A.4.5.3 Fermentation tank SHF E-2 products

Fermentation tank	Produced in SHF (kg/ hr)	
	off gas to scrubber	to beer well
Etoh	5809.17	209345.37
water	200.28	99941.61
CO2	220243.42	4265.67
Solid material	0.00	205462.82

Table A.4.5.4 Fermentation tank SHF E-2 conditions

Temperature	37°C
pH	7
Residence time	48 hour
Condition	Anaerobic

Table A.4.5.5 Sizing of Fermentation tank SHF E-2

Total inlet volume	27817m^3
Size if vessel	1600 m^3
No. of vessel required	20
No of train	4
Approximate sugar fermented	90 %

Table A.4.5.6 Mass balance on Yeast fermentation tank

Mass Balance	Mass In (kg)	Mass Out (kg)
Complex Sugar	664242.4176	66424.24176
Amylase	42.82186905	42.82186905
Cellulase	492.72728	492.72728
Simple Sugar	0	597818.1758
Water	110823696.6	110823696.6

Table A.4.5.7 mass balance of fermentation tank per tank

Mass Balance Fermentation	Mass In	Mass Out
	Amount	Amount
Sugar	664242	31020
Enzyme (Cellulase+Amylase)	251	261
Yeast Inoculum	0.100	0.100
Yeast medium	3.040	3.040
Ethanol	0	237820
CO2	0	248160
NaOH	0.2	0

Appendix A. 4.6 SSF Calculations and Data

Assumption:

- 4 train process
- Each process carries half amount of hydrolysate from pre-treatment in a batch
- Only 70% percent of complex sugar is converted to simple sugar by enzyme activity
- Yeast *Saccharomyces cerevisae*only ferment hexose sugar
- 90 % of hexose sugar is converted to ethanol
- No NaOH addition. pH at 5

Table A.4.6.1 Total sugar converted in fermentation tank SSF E-2 per batch

A batch – in 2 trains	IN (kg)	OUT (kg)	
		Converted	Unconverted
Complex sugar - carbohydrate			
hexose	348181	243727	104455
pentose	144546	101182	43364
Complex sugar - starch	171515	120061	51455

Yeast preparation

Table A.4.6.2 Medium and supporting medium concentration in yeast seeding tank (SSF E-1)

Yeast loading	5 g/L
Ammonium sulfate	1 g/L
Ammonium chloride	2 g/L
Magnesium sulfate	0.8 g/L
LB broth	3 %v/v

Amount of Ammonium chloride per train = 1600 gram
Amount of potassium chloride per train = 800 gram
Amount of magnesium sulphate per train = 640 gram
Amount of yeast per train = 1000 gram
Amount of LB broth per train = 33 L
Vol of water required = 1000 mL (1L of distilled water per 1 kg of yeast)

Table A.4.6.3 Yeast fermentation tank condition SSF E-1

Size of vessel	3.2 m^3
No. of tank required	1
No. of train	2
Temperature	37°C
Residence time	1 hour
Condition	Aerobic

Ethanol conversion (in mass kg)

	$C_6H_{12}O_6$	2 C_2H_5OH	2 CO_2
100% conversion	413061	211120	201941
90% conversion	371755	190008	181747
residue	41306		

Table A.4.6.4 Fermentation tank SHF E-2 products

Fermentation tank	Produced in SHF (kg/ hr)	
	off gas to scrubber	to beer well
Etoh	4305.22	155147.50
water	200.28	99941.61
CO2	149622.12	2897.88
Solid material	0.00	335653.18

Table A.4.6.5 Fermentation tank SSF E-2 conditions

Temperature	37°C
pH	5
Residence time	72 hour
Condition	Anaerobic

Table A.4.6.6 Sizing of Fermentation tank SSF E-2 per train

Total inlet volume	27817m^3
Size if vessel	1600 m^3
No. of vessel required	20
No of train	4
Approximate sugar converted	70 %
Approximate sugar fermented	90 %

Table A.4.6.7 Mass balance on Fermentation tank SSF E-2

	IN	**OUT**
	mass (kg)	mass (kg)
complex sugar released	664242.42	-
cellulose	43	43
amylase	493	493
ethanol	-	190008
CO2 (gas)	-	181747
Unconverted sugar	-	41306
Yeast Inoculum	0.10	0.10
Yeast medium	3.04	3.04

Appendix A. 4.7 Purification Calculations and Data

Purification from SHF

Table A.4.7.1 Mass balance on fermentation tank SHF E-2

Fermentation tank SHF E-2	Produced in SHF (kg/ hr)	IN (kg/hr)	OUT (kg/hr)
		off gas to scrubber	to beer well
Etoh	215154.54	5809.17	209345.37
water	100141.89	200.28	99941.61
CO2	224509.09	220243.42	4265.67
Solid material	205462.82	0.00	205462.82

Table A.4.7.2 Mass balance on Beer Well SHF T-1

Beer well SHF T-1	IN (kg/hr)		OUT (kg/hr)
	from SHF	from scrubber	to beer column
Etoh	209345.37	6295.00	215640.37
water	99941.61	200.28	100141.89
CO2	4265.67	224.42	4490.09
Solid material	205462.82	0.00	205462.82

Table A.4.7.3 Mass balance on Beer Colum PR E-1

Beer column PR E-1	IN (kg/hr)	OUT (kg/hr)		
	from beer well	overhead to scrubber	to Rectifying column	water treatment
Etoh	215640.37	517.54	213483.97	1638.87
water	100141.89	0.00	10014.19	90127.70
CO2	4490.09	4175.79	314.31	0.00
Solid material	205462.82	0.00	0.00	205462.82

Table A.4.7.4 Mass balance on Scrubber PR E-5

Scrubber PR E-5	IN (kg/hr)		OUT (kg/hr)	
	from off gas fermenter	from beer colum	CO2 capture	Beer well
Etoh	5809.17	517.54	31.63	6295.08
water	200.28	0.00	0.00	200.28
CO2	220243.42	4175.79	224194.78	224.42
Solid material	0.00	0.00	0.00	0.00

Table A.4.7.5 Mass balance on Rectifying Colum PR E-2

Rectifying column PR E-2	IN (kg/hr)		OUT (kg/hr)	
	from beer column	from dehydration	to dehydration	to water recycle
Etoh	213483.97	53633.69	266952.66	165.00
water	10014.19	641.62	674.65	9981.16
CO2	314.31	0.00	0.00	314.31
Solid material	0.00	0.00	0.00	0.00

Table A.4.7.6 Mass balance on Molecular Sieve PR E-3 E-4

MOL sieve PR E-3 E-4	IN (kg/hr)	OUT (kg/hr)	
	from rectifying column	to rectifying column	ETOH storage
Etoh	266952.66	53633.69	213318.97
water	674.65	641.62	33.03
CO2	0.00	0.00	0.00
Solid material	0.00	0.00	0.00

Purification from SSF

Table A.4.7.7 Mass balance on fermentation tank SHF E-2

Fermentation tank SSF E-2	Produced in SHF (kg/ hr)	IN (kg/hr)	OUT (kg/hr)
		off gas to scrubber	to beer well
Etoh	159452.73	4305.22	155147.50
water	100141.89	200.28	99941.61
CO2	152519.9995	149622.12	2897.88
Solid material	335653.18	0.00	335653.18

Table A.4.7.8 Mass balance on Beer Well SHF T-1

Beer well SSF T-1	IN (kg/hr)		OUT (kg/hr)
	from SHF	from scrubber	to beer column
Etoh	155147.50	4665.00	159812.50
water	99941.61	200.28	100141.89
CO2	2897.88	156.00	3053.88
Solid material	335653.18	0.00	335653.18

Table A.4.7.9 Mass balance on Beer Colum PR E-1

Beer column PR E-1	IN (kg/hr)	OUT (kg/hr)		
	from beer well	overhead to scrubber	to Rectifying column	water treatment
Etoh	159812.50	383.55	158214.38	1214.58
water	100141.89	0.00	10014.19	90127.70
CO2	3053.88	2840.11	213.77	0.00
Solid material	205462.82	0.00	0.00	205462.82

Table A.4.7.10 Mass balance on Scrubber PR E-5

Scrubber PR E-5	IN (kg/hr)		OUT (kg/hr)	
	from off gas fermenter	from beer colum	CO2 capture	Beer well
Etoh	4305.22	383.55	23.44	4665.33
water	200.28	0.00	0.00	200.28
CO2	149622.12	2840.11	152309.77	152.46
Solid material	0.00	0.00	0.00	0.00

Table A.4.7.11 Mass balance on Rectifying Colum PR E-2

Rectifying column PR E-2	IN (kg/hr)		OUT (kg/hr)	
	from beer column	from dehydration	to dehydration	to water recycle
Etoh	158214.38	39748.28	197840.38	122.28
water	10014.19	641.62	674.65	9981.16
CO2	213.77	0.00	0.00	213.77
Solid material	0.00	0.00	0.00	0.00

Table A.4.7.12 Mass balance on Molecular Sieve PR E-3 E-4

MOL sieve PR E-3 E-4	IN (kg/hr)	OUT (kg/hr)	
	from rectifying column	to rectifying column	ETOH storage
Etoh	197840.38	39748.28	158092.10
water	674.65	641.62	33.03
CO2	0.00	0.00	0.00
Solid material	0.00	0.00	0.00

APPENDIX B- ECONOMIC ASSESSMENT

Appendix B.1- Cultivation Cost Data

Table B.1.1 Comparative economics of open ponds and closed photobioreactors (PBRs) (use of algae)

Parameter	Relative advantage
Capital/operating cost	Open ponds<< PBRs
Biomass concentration	Open ponds< PBRs
Oxygen inhibition	Open ponds> PBRs
Contamination risk	Open ponds< PBRs
Water losses	Open ponds≈ PBRs
Carbon dioxide losses	Open ponds≈ PBRs
Process control	Open ponds≈ PBRs
Space required	Open ponds≈ PBRs

Table B.1.2. Comparison Cost for different cultivation methods (Past yr report)

Parameters	Cultivation System		
	Raceway Pond	Horizontal Tubular Reactor	External Loop Reactor
Total Delivered Equipment Cost (M$ AUD)	216	837	1,483
Total Capital Investment (M$ AUD)	734	2721	15
Energy Cost	692	8676	7757
Annual Production Cost (M$ AUD)	139	495	649
Cost per kg of wet Biomass produced (AUD)	2.7	10	13

Appendix B.2- Dewatering Cost Data

Table B.2.1 Dewatering cost summary (dual stage, flocculation+filtration)

Raceway Pond

Basis	Total Batch Volume (m3)	Day to process batch	Hours processed each day	Time to flocculate	Batches per tank	Tank Volume (m3)	Effective Volume of each tank	Number of Tanks	*Time to flocculate = 1 hr to fill + 1 hr to mix + 1 hr to settle + 1 hr to pump out
Tanks	1295726	5	24	4	30	40000	1200000	1	

Energy Consumption

	Energy / impeller (kW)	Mixing Time (Hrs)	Energy Use per tank (kWh)	Tank mixing rate	Hours Flocculated over	Energy Consumed (kWH)	Price of Electricity (US$/kWh)	Number of Batches	Total Energy Use	Total Cost
Mixing	4000	1	4000	0.27	120	129572	0.06	66	8551795	**513108**

	Unit Cost (2008 $/kWh)	Energy Consumed (kWh/m3) per volume processed	Total Volume Processed(m3)	Standardisation	Batches processed annually	Centrifuge Total Energy Use (kWh)	Total Cost
Suction Filter	0.06	0.1	16196.58	1.33	66	**142758**	**8565**

First Stage	Unit Cost ($ in 2002)	Index (2002)	Index(2008)	Scaling	Unit Cost ($/m3 in 2008)	Number of Centrifuges Required	Total Price
Biomass Feed Pumps (S.S, Centrifugal, 3600m3/h)	43200	697.8	902.1	1.29	55848	22	1228655
Flocculation Tank (40000m3)	3400000	353	603.4	1.709	5811785	1	5811784
							7040440
Second Stage							
Harvested Biomass Conveyer Belt	7100	438.4	620	1.414	10041	1	10041
Suction Filter	765000	438.4	620	1.414	1081888	1	1081889
							1091930

Energy Costs	521673	
Flocculant Cost	37627897	
Annual Cost	5128473	
Total Cost	**3278043**	

Raw material (Flocculant)	Unit Price (US$/kg)	Kg required per tank	Cost per tank	Number of tanks	Cycles per tank	Batches per year	Total Price
Chitosan	11	1600	17600	1	30	66	**37627897**

Total and Annual Fixed Capital for Dewatering	% of MEC	Cost ($)	
Delivered Equipment Cost	1	8132370	

Delivered Equipment Installation	0.3	2439711		
Instrumentation and Control (installed)	0.1	813237		
Piping (installed)	0.3	2439711		
Electrical Systems (installed)	0.1	813237		
Buildings (including services)	0.3	2439711		
Yard Improvements	0.1	813237		
Service Facilities	0.2	1626474		
Total Direct Plant Cost		**19517688**		
Engineering and Supervision	0.25	2033092		
Construction Expenses	0.1	1951769		
Contractors Fees	0.05	975884		
Subtotal		24478434		
Contingency	0.06	1468706		
Total Indirect Plant Cost		**6429452**		
Fixed Capital Investment		**25947140**		
Annualised Costs				
Depreciation		2594714	10.00	depreciation period (years)
Goods and Services Tax		626280	0.10	percentage of FCI
Contingency		1907479	0.05	
Annualised Cost		**5128473**		

Table B.2.2 Dewatering cost summary (dual stage, flocculation+filtration)

Horizontal Tubular Reactor

Basis	Total Batch Volume (m3)	Day to process batch	Hours processed each day	Time to flocculate	Batches per tank	Tank Volume (m3)	Effective Volume of each tank	Number of Tanks Necessary	**Time to flocculate = 1 hr to fill + 1 hr to mix + 1 hr to settle + 1 hr to pump out
Tanks	133923	4	24	4	24	6000	144000	0.930	

Energy Consumption

	Energy per impeller (kW)	Mixing Time (Hrs)	Energy Use per tank (kWh)	Tanks mixing at once	Hours Flocculated over	Energy Consumed (kWH)	Price of Electricity (US$/kWh)	Number of Batches	Total Energy Use	Total Cost
Mixing	600	1	600	0.233	120	16740.33	0.06	82.5	1381077	**82865**

	Unit Cost (2008 $/kWh)	Energy Consumed (kWh/m3) per volume processed	Total Volume Processed(m3)	Standardisation	Batches processed annually	Centrifuge Total Energy Use (kWh)	Total Cost
Suction Filter	0.06	0.1	1674.03	0.172	82.50	**2384**	**143**

First Stage	Unit Cost ($ in 2002)	Index (2002)	Index(2008)	Scaling	Unit Cost ($/m3 in 2008)	Number of Centrifuges Required	Total Price
Biomass Feed Pumps (S.S, Centrifugal, 3000m3/h)	40800	697.8	902.1	1.292	52745	4	210981
Flocculation Tank (6000m3)	660000	353	603.4	1.709	1128170	1	1049219
							1260200
Second Stage							
Harvested Biomass Conveyer Belt	7100	438.4	620	1.414	10041	1	10041
Suction Filter	765000	438.4	620	1.414233577	1081888	1	1081889
							1091930

Energy Costs	83008
Flocculant Cost	4,861392
Annual Cost	1117934
Total Cost	**6062334**

Flocculant	Unit Price (US$/kg)	Kg required per tank	Cost per tank	Number of tanks	Cycles per tank	Batches per year	Total Price
Chitosan	11	240	2640	1	24	82.5	**4861392**

Total and Annual Fixed Capital for Dewatering	% of MEC	Cost ($)
Delivered Equipment Cost	1	2352130
Delivered Equipment Installation	0.3	705639

Instrumentation and Control (installed)	0.1	235213		
Piping (installed)	0.3	705639		
Electrical Systems (installed)	0.1	235213		
Buildings (including services)	0.3	705639		
Yard Improvements	0.1	235213		
Service Facilities	0.2	470426		
Total Direct Plant Cost		**5645112**		
Engineering and Supervision	0.25	588032		
Construction Expenses	0.1	564511		
Contractors Fees	0.05	282256		
Subtotal		7079911		
Contingency	0.06	424795		
Total Indirect Plant Cost		**1859594**		
Fixed Capital Investment		**7504705**		
Annualised Costs				
Depreciation		750471	10.00	depreciation period (years)
Goods and Services Tax		120243	0.10	percentage of FCI
Contingency		247220	0.05	
Annualised Cost		**1117934**		

Table B.2.3 Dewatering cost summary (dual stage, flocculation+filtration)

External Loop Reactor

Basis	Total Batch Volume (m3)	Day to process batch	Hours processed each day	Time to flocculate	Batches per tank	Tank Volume (m3)	Effective Volume of each tank	Number of Tanks Necessary
Tanks	119737	3	24	4	18	6500	117000	1

Energy Consumption

	Energy per impeller (kW)	Mixing Time (Hrs)	Energy Use per tank (kWh)	Tanks mixing at once	Hours Flocculated over	Energy Consumed (kWH)	Price of Electricity (US$/kWh)	Number of Batches	Total Energy Use	Total Cost
Mixing	650	1	650	0.256	120	19956.14	0.06	110	2195175	**131711**

	Unit Cost (2008 $/kWh)	Energy Consumed (kWh/m3) per volume processed	Total Volume Processed(m3)	Standardisation	Batches processed annually	Centrifuge Total Energy Use (kWh)	Total Cost
Suction Filter	0.06	0.1	1496.71	0.205592105	110.00	**3385**	**203.09**

First Stage	Unit Cost ($ in 2002)	Index (2002)	Index(2008)	Scaling	Unit Cost ($/m3 in 2008)	Number of Centrifuges Required	Total Price
Biomass Feed/Outlet Pumps (3250m3/h)	42000	697.8	902.1	1.292	54297	4	217187
Flocculation Tank (SS, 6500m3)	700000	353	603.4	1.7093484	1196544	1	1224533
							1441720
Second Stage							
Harvested Biomass Conveyer Belt	7100	438.4	620	1.414	10041	1	10041
Suction Filter	765000	438.4	620	1.414	1081888	1	1081889
							1091930

Energy Costs	131914
Flocculant Cost	5795263
Annual Cost	1240280
Total Cost	**7167456**

Flocculant	Unit Price (US$/kg)	Kg required per tank	Cost per tank	Number of tanks	Cycles per tank	Batches per year	Total Price
Chitosan	11	260	2860	1	18	110	**5,795,263**

Total and Annual Fixed Capital for Dewatering	% of MEC	Cost ($)
Delivered Equipment Cost	1	2533649

Delivered Equipment Installation	0.3	760095		
Instrumentation and Control (installed)	0.1	253365		
Piping (installed)	0.3	760095		
Electrical Systems (installed)	0.1	253365		
Buildings (including services)	0.3	760095		
Yard Improvements	0.1	253365		
Service Facilities	0.2	506730		
Total Direct Plant Cost		**6080759**		
Engineering and Supervision	0.25	633,412		
Construction Expenses	0.1	608,076		
Contractors Fees	0.05	304,038		
Subtotal		7626,285		
Contingency	0.06	457577		
Total Indirect Plant Cost		**2003103**		
Fixed Capital Investment		**8083862**		
Annualised Costs				
Depreciation		808386	10.00	depreciation period (years)
Goods and Services Tax		13,535	0.10	percentage of FCI
Contingency		296359	0.05	
Annualised Cost		**1240280**		

Table B.2.4. Comparison cost for different cultivation methods [5]

Parameters	Cultivation System		
	Raceway Pond	Horizontal Tubular Reactor	External Loop Reactor
Centrifugation			
Total Delivered Equipment Cost ($ AUD x10^3)	418	294	297
Fixed Capital Investment ($ AUDx10^3)	1,333	938	948
Energy Cost (AU$ million)	36.5	0.61	0.87
Annualised Cost (M$ AUD)	2.3	0.14	0.16
Chamber Filtration			
Total Delivered Equipment Cost ($ AUD x10^3)	274	150	153
Fixed Capital Investment ($ AUD x10^3)	873	477	487
Energy Cost (AU$ million)	15.8	0.26	0.37
Annualised Cost (AUD x10^3)	1040	68	76
Suction Filtration			
Total Delivered Equipment Cost ($ AUD x10^3)	1259	1135	1138
Fixed Capital Investment ($ AUD x10^3)	4018	3623	3632
Energy Cost (AU$ million)	5.5	0.09	0.13
Annualised Cost (AUD x10^3)	769	402	405
Flocculation + Centrifugation			
Total Delivered Equipment Cost ($ AUD x10^3)	7291	1510	1692
Fixed Capital Investment ($ AUD million)	23.2	4.82	5.4
Energy Cost (AU$ million)	0.97	0.09	0.14
Annualised Cost (AUD million)	43.5	5.8	6.9
Flocculation + Filtration			
Total Delivered Equipment Cost ($ AUD x10^3)	8132	2352	2533
Fixed Capital Investment ($ AUD million)	26	7.5	8.1
Energy Cost (AU$ million)	0.52	0.08	0.13
Annualised Cost (AUD million)	43.2	6.1	7.2

Appendix B.3- Pre-treatment Cost Calculations (Acid Hydrolysis)

Table B.3.1 Acid hydrolysis cost summary

Major equipment cost										
Equipment tag	**Acid pretreatment**		Unit Cost	year	Index (at presented year)	Index (2010)	Scaling	Unit cost (2010)	Equipment Quantity	Total Price
PT BC-1	Pneumatic solids conveying- biomass	fig 12-63	$190,000.00	2002	438.4	627	1.430	$271,738.14	1	$271,738.14
PT T-1	Water storage tank	fig12-55 extrapolate (1)	$369,222.30	2002	353	617.8	1.750	$646,191.32	5	$3,230,956.61
PT T-2	Sulfuric Acid storage tank	fig 12-55 extrapolate (1)	$200,092.40	2002	353	617.8	1.750	$350,190.04	1	$350,190.04
PT E-1	Jacketed Acid PT Reactor	fig 12-55 extrapolate (2)	$267,643.00	2002	353	617.8	1.750	$468,413.16	6	$2,810,478.96
PT P-1	Pump-acid in	fig 12-23 extrapolate (3)	$7,339.05	2002	697.8	902.5	1.293	$9,491.96	2	$18,983.93
PT P-2	Pump-water in	fig 12-23 extrapolate(3)	$13,350.83	2002	697.8	902.5	1.293	$17,267.30	2	$34,534.60
PT P-3	Pump- water for steam	fig 12-23 extrapolate (3)	$4,915.55	2002	697.8	902.5	1.293	$6,357.53	2	$12,715.06
PT P-4	Pump- blowdown tank in	fig 12-23 extrapolate (3)	$12,419.54	2002	697.8	902.5	1.293	$16,062.82	2	$32,125.65
PT H-1	Water heater	fig 14-38	$140,000.00	2002	353	617.8	1.750	$245,019.83	1	$245,019.83
Blowdown										
PT E-2	blow down tank	Aden et al	$161,482.00	2000	371.1	617.8	1.665	$268,832.07	9	$161,482.00
PT P-5	Pump- S/L separator IN	fig 12-23 extrapolate (3)	$9,820.65	2002	697.8	902.5	1.293	$12,701.54	3	$9,820.65
Centrifuge										
PT E-3	Centrifuge	fig 15-47								
PT P-6	Pump- Neutralisation tank IN	fig 12-23 extrapolate	$190,000.00	2002	438.4	627	1.430	$271,738.14	3	$190,000.00
Neutralisation										
PT E-4	Neutralisation Reactor tank	fig 12-55 extrapolate								

PT BC-2	Pneumatic solids conveying (NaOH)	fig 12-63								
PT P-7	Pump- to E.H/SSF storage	fig 12-23 extrapolate	$336,033.00	2002	353	617.8	1.750	$588,105.35	3	$336,033.00
PT P-8	pump to EH /SSF	fig 12-23 extrapolate (3)	$10,000.00	2002	438.4	627	1.430	$14,302.01	1	$10,000.00
PT E-5	Storage tank	fig 12-55 extrapolate (1)	$9,604.94	2002	697.8	902.5	1.293	$12,422.55	3	$9,604.94
	Total MEC									**$12,206,592.66**

(1)y = 227.02x + 25287

(2)y = 1367.8x + 14600

(3)y = 19336 x^0.433

All figure all based on [9]

Total and Annual Fixed Capital for SHF		% MEC	Cost ($)
Major Purchased Equipment (MEC)		-	$12,206,592.66
Equipment Installation Cost		0.3	$3,661,977.80
Instrumentation and Control (installed)		0.1	$1,220,659.27
Piping (installed)		0.3	$3,661,977.80
Electrical Systems (installed)		0.1	$1,220,659.27
Buildings (including services)		0.3	$3,661,977.80
Yard Improvements		0.1	$1,220,659.27
Service Facilities		0.2	$2,441,318.53
Land		0.06	$732,395.56
Total Direct Plant Cost			**$30,028,217.95**
Engineering and Supervision		0.25	$3,051,648.17
Construction Expenses		0.1	$3,002,821.80
Contractors Fees		0.05	$1,501,410.90
Contingency		0.06	$2,255,045.93
Total Indirect Plant Cost			**$9,810,926.79**

Total Fixed Capital Investment			**$39,839,144.74**			
Direct Cost of Pre-treatment						
Raw Materials	Unit	Unit usage/d	batches per year	Unit cost(USD$/unit)	Annual Cost	
NaOH	kg	4.45	33	85	12484	Sigma aldrich
Sulfuric acid	L	1400.93	33	300	13869298	(alibaba, china, 200-400USD)
fresh water	m3	151515	33	1	5000000	
Total Raw Material Cost					**13881782**	

Equipment tag	Equipment	Energy per impeller (kW)	Mixing Time (Hrs)	Tanks mixed per day	batches per year	Energy Consumed Annually(kWH)	Price of Electricity (USD/kWh)	Annual Cost	
PT A-1	pretreatment mixing (29.55 W/m3)	37.52	24	4	33	118890	0.07	8322	
PT A-2	neutralisation mixing (98.5 W/m3)	125.58	24	4	33	397861	0.07	27850	
	Total Mixing Electricity					**516751**		**36173**	

Equipment tag	Equipment	Power Consumption (kW)	No of Pumps	Hours running per day	batches per year	Annual Energy Consumed (kWh)	Price of Electricity (USD/kWh)	Annual Cost
PT H-1	Water heater for steam	5900.00	1	24	33	4672800	0.07	$327,096.00
PT P-1	Pump-acid in	256.19	2	24	33	405799.1584	0.07	$28,405.94
PT P-2	Pump-water in	1020.27	2	24	33	1616102.686	0.07	$113,127.19
PT P-3	Pump- water for steam	101.60	2	24	33	160934.4	0.07	$11,265.41
PT P-4	Pump- blowdown tank IN	863.36	2	24	33	1367557.607	0.07	$95,729.03
PT P-5	Pump- S/L separator IN	502.00	3	24	33	1192754.59	0.07	$83,492.82
PT E-3	Centrifuge	750.00	3	12	33	891000	0.07	$62,370.00
PT P-6	Pump- Neutralisation tank IN	476.90	3	24	33	1133117.077	0.07	$79,318.20

PT P-7	Pump- to E.H/SSF storage	476.90	3	24	33	1133117.077	0.07	$79,318.20
PT P-8	pump to EH /SSF	238.45	4	24	33	755411.0259	0.07	$52,878.77
	Total Pumping Electricity					**9415948.441**		
	Total Utilities Cost							**$880,122.78**
TOTAL COST								
Fixed cost		$3,967,819.04	AUD					
Raw Material cost		$13,118,284.40	AUD					
Energy cost		$867,478.25	AUD					
Total Annual Pre-treatment cost		**$17,953,581.70**	AUD					

Appendix B.4- Fermentation Cost Calculations

Appendix B.4.1- Simultaneous Hydrolysis and Fermentation (SHF)

Table C.4.1 Enzyme Hydrolysis Cost

Major Equipment List									
Equip. tag	Enzyme Hydrolysis		Unit Cost (in 2002)	Index (2002)	Index (2010)	Scaling	Unit cost ($/m3 in 2010)	Equipment Quantity	Total Price
SHF E-3	Enzyme Bioreactor Tank	fig 12-55 (1)	$388,519.00	353.00	617.80	1.750141643	$679,963.28	20	$13,599,265.62
SHF BC-2	Pneumatic solids conveying equipment (cellulase),	fig12-60	$35,000.00	438.40	627.00	1.43020073	$50,057.03	1	$50,057.03
SHF BC-3	Pneumatic solids conveying equipment (amylase)	fig12-63	$28,000.00	438.40	627.00	1.43020073	$40,045.62	1	$40,045.62
SHF P-2	Pump- Biomass OUT to fermentation (5632m3/h)	fig12-23	$14,282.45	697.80	902.50	1.29335053	$18,472.21	4	$73,888.83
SHF HX-1	Cooler- Bioethanol IN to fermentation (185m2)	fig14-29	$6,000.00	698.70	902.50	1.291684557	$7,750.11	4	$31,000.43
	Total MEC								**25378280**
	y = 227.02x + 25287 (1)								

Total and Annual Fixed Capital for E.H	% of MEC	Cost ($)	Total Cost				
Major Purchased Equipment (MEC)	-	$13,794,257.53					
Equipment Installation Cost	0.3	$4,138,277.26					
Instrumentation and Control (installed)	0.1	$1,379,425.75					
Piping (installed)	0.3	$4,138,277.26					
Electrical Systems (installed)	0.1	$1,379,425.75					
Buildings (including services)	0.3	$4,138,277.26					
Yard Improvements	0.1	$1,379,425.75					

Service Facilities		0.2	$2,758,851.51					
Land		0.06	$827,655.45					
Total Direct Plant Cost				**$33,106,218.07**				
Engineering and Supervision		0.25	$3,448,564.38					
Construction Expenses		0.1	$3,310,621.81					
Contractors Fees		0.05	$1,655,310.90					
Contingency		0.06	$504,869.83					
Total Indirect Plant Cost				**$8,919,366.92**				
Total Fixed Capital Investment					**$42,025,584.99**			

Direct Cost of E.H								
Raw Materials	Unit cost(USD$/gallon bioethanol)	gallon of ethanol produced/50000tonnes biomass (per 330day)		Annual Cost	($/y)		*Michael, Harun	
Cellulase	0.15	2377244.67		356587			3 days for enzymatic hydrolysis	
Amylase	0.02	2377244.67		47545				
Total Raw Material Cost					**404132**			
Total Utilities Cost					**981590**			
Total Enzyme Hydrolysis Cost				USD	10954260	Exchange Rate	1 USD = 0.945 AUD	
				AUD	**10351776**			

Equip. tag			Energy per impeller	Tanks mixed per day	Mixing Time (Hrs)	Batch per yr	Energy Consumed Annually (kWh)	Price of Electricity (USD/MWh)	Annual Cost
	Enzyme Hydrolysis Mixing		53.94545608	20	24	66	1708992.0	70	119629.443

			Power Consumption	No of units	Hours running per	Batch per year	Annual Energy Consumed	Price of Electricity (USD/MWh)	Annual Cost

			(kW)		day		(kWh)		
SHF HX-1	Cooler OUT to fermentation		37310.18	5	1	66	12312359.4	$70.00	37310.18
SHF P-2	Biomass OUT to fermentation pump		238.45	20	1	66	314754.5941	$70.00	238.45

Table B.4.2 Fermentation Cost

Major Equipment List

Equipment tag	Fermentation		Unit Cost (in 2002)	Index (2002)	Index (2010)	Scaling	Unit cost ($/m3 in 2010)	Equipment Quantity	Total Price
SHF E-1	Jacketed Yeast Fermentation Tank (3.2m3)	fig13-15	$18,976.96	353	617.8	$1.75	$33,212.37	1	$33,212.37
SHF E-2	Jacketed Fermentation Tank (1600m3)	fig 12-55	$2,203,080.00	353	617.8	$1.75	$3,855,702.05	20	$77,114,041.02
SHF T-1	Ethanol Storage Tank (1400m3)	fig 12-55	$343,115.00	353	617.8	$1.75	$600,499.85	20	$12,009,997.00
SHF BC-1	Pneumatic solids conveying equipment (supporting medium powder+NaOH)	fig12-63,p576	$28,000.00	438.4	617.8	$1.41	$39,458.03	5	$197,290.15
SHF P-3	Pump- Ethanol OUT to EtOH tank (2682m3/h)	fig12-21	$23,230.83	697.8	902.5	$1.29	$30,045.60	10	$300,456.04
SHF P-4	Pump- Ethanol OUT to purification (2682m3/h)	fig12-21	$23,230.83	697.8	902.5	$1.29	$30,045.60	10	$300,456.04
SHF P-1	Pump- water for steam (centrifugal, 12m3/hr)	fig 12-23	$4,433.77	697.8	902.5	$1.29	$5,734.42	4	$22,937.67
SHF H-1	Water heater for steam (heater @6135KW, CS, 202 kPa)	fig 14-38	$28,000.00	353	617.8	$1.75	$49,003.97	5	$245,019.83
	Total MEC								**$90,223,410.12**
	y = 227.02x + 25287 (1)								
	y = 1367.8x + 14600 (2)								

Total and Annual Fixed Capital for SHF	% of MEC	Cost ($)					
Major Purchased Equipment (MEC)	-	$90,223,410.12					
Equipment Installation Cost	0.3	$27,067,023.04					
Instrumentation and Control (installed)	0.1	$9,022,341.01					
Piping (installed)	0.3	$27,067,023.04					
Electrical Systems (installed)	0.1	$9,022,341.01					
Buildings (including services)	0.3	$27,067,023.04					
Yard Improvements	0.1	$9,022,341.01					
Service Facilities	0.2	$18,044,682.02					
Land	0.06	$5,413,404.61					
Total Direct Plant Cost			**$221,949,588.90**				
Engineering and Supervision	0.25	$22,555,852.53					
Construction Expenses	0.1	$22,194,958.89					
Contractors Fees	0.05	$11,097,479.45					
Contingency	0.06	$3,350,897.45					
Total Indirect Plant Cost			**$59,199,188.32**				
Total Fixed Capital Investment				**$281,148,777.22**			

Direct Cost of SHF									
Raw Materials	Unit	Unit usage/batch	Unit cost(USD$/unit)	No of batch (2sub batches)	Annual Cost				
NaOH	kg	0.2152	85	66	1,207			Sigma aldrich	
LB broth	kg	1.666	132.6	66	14,586			Sigma aldrich	
yeast	kg	0.2	86	66	1,135			Sigma aldrich	
Ammonium Chloride	kg	1.6	77.8	66	8,216			Sigma	

								aldrich	
Potassium Sulphate	kg	0.8	187.96	66	9,924			Sigma aldrich	
Magnesium Sulphate	kg	0.64	130	66	5,491			Sigma aldrich	
Fresh Water	m3	0.069	1	66	4.532			Brennan	
Nitrogen Purge (for ethanol fermentor)	m3	122400.00	0.3	66	2,423,520			Brennan	
Total Raw Material Cost, C						**2,464,084**			
Total Utility Cost, D						**233,160**			
Total Fermentation Cost					USD	58,692,645	0.945AUD=1USD		
					AUD	**55,464,550**			

Utilities		Energy per impeller (kW)	Mixing Time (Hrs)	Tanks mixed per day	No of batch (2sub batches)	Energy Consumed Annually(kWh)	Price of Electricity (USD/kWh)	Annual Cost
SHF A-1	yeast fermentation mixing	0.01	24	1	66	21.53	0.07	1.50
SHF A-2	ethanol fermentation mixing	53.95	24	20	66	1708992	0.07	119629
		Power Consumption (kW)	No of pumps	Hours running per day	No of batch (2 sub-batches)	Energy Consumed Annually (kwh)	Price of Electricity (USD/kWh)	Annual Cost
SHF P-3	Pump- Ethanol OUT to EtOH tank	1908.06	10	1	66	1259322.753	0.07	88152.593
SHF P-4	Pump- Ethanol OUT to Purification	1908.06	10	1	66	1259322.753	0.07	88152.593
SHF P-1	Pump- water for steam	1.99E-03	4	1	66	0.524479132	0.07	0.037
SHF H-1	Water heater for steam	6135.85	4	1	66	1619863.893	0.07	113390.473

Appendix B.4.2- Separate Hydrolysis and Fermentation (SHF)

Table B.4.3 SIMULTANEOUS SACCHARAFICATION AND HYDROLYSIS (SSF)

Major equipment cost

Equip. tag	Equipment		Unit Cost (in 2002)	Index (2002)	Index (2010)	Scaling	Unit cost ($/m3 in 2010)	Equipment Quantity	Total Price
SSF E-1	Jacketed Yeast Fermentation Tank (3.2m3)	fig13-15,p628	$18,976.96	353	617.8	1.750141643	33212.37	1	$33,212.37
SSF E-2	Jacketed Fermentation Tank (1600m3)	fig 12-55 extrapolate	$2,203,080.00	353	617.8	1.750141643	3855702.05	20	$77,114,041.02
SSF T-1	Ethanol Storage Tank (1400m3)	fig 12-55 extrapolate	$343,115.00	353	617.8	1.750141643	600499.85	20	$12,009,997.00
SSF BC-1	Pneumatic solids conveying equipment (cellulase),	fig12-60,p573	$35,000.00	438.4	627	1.43020073	50057.03	1	$50,057.03
SSF BC-2	Pneumatic solids conveying equipment (amylase)	fig12-63,p576	$28,000.00	438.4	627	1.43020073	40045.62	1	$40,045.62
SSF P-2	Pump- fermentation OUT (2682m3/hr)	fig 12-23 extrapolate	$23,230.83	697.8	902.5	1.29335053	30045.60	10	$300,456.04
SSF P-3	Pump- ethanol storage tank OUT (2682m3/hr)	fig 12-23 extrapolate	$23,340.23	697.8	902.5	1.29335053	30187.09	10	$301,870.92
SSF P-1	Pump- water for steam (centrifugal, 12m3/hr)	fig 12-23 extrapolate	$4,433.77	697.8	902.5	1.29335053	5734.42	4	$22,937.67
SSF H-1	Water heater for steam (direct-fired heater @6135KW, CS, 202 kPa)	fig 14-38	$28,000.00	353	617.8	1.750141643	49003.97	4	$196,015.86
	Total MEC								**$90,068,633.54**

Total and Annual Fixed Capital for SHF	% of MEC	Cost ($)					
Major Purchased Equipment (MEC)	-	90068633.54					
Equipment Installation Cost	0.3	27,020,590					
Instrumentation and Control (installed)	0.1	9,006,863					
Piping (installed)	0.3	27,020,590					
Electrical Systems (installed)	0.1	9,006,863					
Buildings (including services)	0.3	27,020,590					
Yard Improvements	0.1	9,006,863					
Service Facilities	0.2	18,013,727					
Land	0.06	5,404,118					
Total Direct Plant Cost		**$221,949,588.90**					
Engineering and Supervision	0.25	22,517,158					
Construction Expenses	0.1	22,156,884					
Contractors Fees	0.05	11,078,442					
Contingency	0.06	16,639,279					
Total Indirect Plant Cost		**72,391,764**					
Total Fixed Capital Investment		**293,960,602**					

Direct Cost of Pre-treatment								
Raw Materials	Unit	Unit usage/d		No of batch (2sub batches)		Unit cost(USD$/unit)	Annual Cost	
Cellulase	kg/ gallon ethanol	1584831		66		0.15	237,725	
Amylase	kg/ gallon ethanol	1584831		66		0.02	31,697	
LB broth	kg	1.666667		66		132.6	14,586	
yeast	kg	0.2		66		86	1,135	Sigma aldrich

Ammonium Chloride	kg	1.6		66		77.8	8,216	Sigma aldrich	
Potassium Sulphate	kg	0.8		66		187.96	9,924	Sigma aldrich	
Magnesium Sulphate	kg	0.64		66		130	5,491	Sigma aldrich	
Fresh Water	m3	0.069		66		1	4.532	Brennan	
Total Raw Material Cost							**308,778**		
Utilities		Energy per impeller (kW)	Mixing Time (Hrs)	Tanks mixed per day	No of batch (2sub batches)	Energy Consumed Annually(kWH)	Price of Electricity (USD/kWh)	Annual Cost	
SSF A-1	Yeast Fermeneter mixing	0.01	24	1	66	15.84	0.07	1	(Aden et al. 2002)
SSF A-2	Fermenter Mixing	53.95	24	20	66	1709136	0.07	119,640	(Aden et al. 2002)
					Total Mixing Electricity	**1709151.84**		**119,641**	
		Power Consumption (kW)	No of Pumps	Hours running per day	No of batch (2sub batches)	Annual Energy Consumed (kWh)	Price of Electricity (USD/kWh)	Annual Cost	
SSF P-2	Pump-fermentation OUT	1.418	10	1	66	935.88	0.07	65.51	1.418
SSF P-3	Pump- ethanol storage tank OUT	1.579	10	1	66	1042.14	0.07	72.95	1.579
SSF P-1	Pump- water for steam	1.99E-03	4	1	66	0.524479132	0.07	0.04	1.99E-03
SSF H-1	Water heater for steam	6135.848081	4	1	66	1619863.893	0.07	113390.47	6135.848081
TOTAL COST									

Appendix B.5- Product Recovery Cost Calculations

Table B.5.1 Purification cost summary

Major Equipment List

Eqpt Tag	Equipment	Unit Cost(2000)	Index (2000)	Index (2010)	Scaling	Unit cost ($/m3 in 2010)	Equipment Quantity	Total Price
PR E-1	Beer column (	1,105,399	440.3	667.5	1.516	1675797.94	1	1675797.939
PR E-2	Rectifying Column (	1,330,413	440.3	667.5	1.5160	2016921.82	1	2016921.82
PR HX-1	Beer column feed HX-1	944,658	371.1	617.8	1.664	1572648.11	1	1572648.107
PR HX-2	Beer column feed HX-2 plate	92,562	371.1	617.8	1.664	154095.40	1	154095.4018
PR HX-4	Beer colum reboiler	774,072	371.1	617.8	1.664	1288659.88	1	1288659.88
PR HX-6	Rectifying Column reboiler	71,074	371.1	617.8	1.664	118322.60	1	118322.6009
PR HX-3	Beer column Condenser	32,614	371.1	617.8	1.664	54295.15	1	54295.1474
PR HX-5	Rectifying column condenser	233,975	371.1	617.8	1.664	389516.99	1	389516.99
PR E-3 E-4	Mol Sieve (9 pieces)	3,112,319	371.1	617.8	1.664	5181327.62	1	5181327.616
PR P-1	Beer Column Bottom pump	279,004	669	902.5	1.349	376384.32	1	376384.3199
PR P-2	Beer Column Reflux Pump	3,510	669	902.5	1.349	4735.09	1	4735.089686
PR P-3	Rectification Column Bottoms Pump	32,059	669	902.5	1.349	43248.50	1	43248.50149
PR P-4	Rectification Column Reflux Pump	35,182	669	902.5	1.349	47461.52	1	47461.51719

PR DR-1	Beer Column Relfux Drum	10,122	371.1	617.8	1.349	16850.91	1	16850.90703
PR DR-2	Rectification Column Reflux Drum	123,885	371.1	902.5	2.431	301283.25	1	301283.2458
PR E-5 E-6	Vent Scrubber	268,480	440.3	902.5	2.049	550313.88	1	550313.8769
PR P-5	Scrubber Bottoms Pump	15,878	669	902.5	1.349	21419.87	1	21419.87294
	Total MEC							**13813282.83**

Total and Annual Fixed Capital for SHF	% of MEC	Cost ($)						
Major Purchased Equipment (MEC)	-	13813282						
Equipment Installation Cost	0.3	4,143,985						
Instrumentation and Control (installed)	0.1	1,381,328						
Piping (installed)	0.3	4,143,985						
Electrical Systems (installed)	0.1	1,381,328						
Buildings (including services)	0.3	4,143,985						
Yard Improvements	0.1	1,381,328						
Service Facilities	0.2	2,762,657						
Land	0.06	828,797						
Total Direct Plant Cost			**33,151,879**					
Engineering and Supervision	0.25	3,453,321						
Construction Expenses	0.1	3,315,188						
Contractors Fees	0.05	1,657,594						
Contingency	0.06	505,566						
Total Indirect Plant Cost			**8,931,669**					
Total Fixed Capital Investment				**42,083,547**				

Direct Cost for Purification											
	Utilities	Units	Unit usage/d	sf	F	KW	Hours running/day	No of batch (2sub batches)	Annual Energy Consumed (kWh)	Price of Electricity (USD/kWh)	Annual Cost
PR HX-2	Beer column feed HX-2 plate frame	BTU/hr sf F	200	909	253	13466.431	24	33	10665413.3	0.07	746578.93
PR HX-4	Beer colum reboiler	BTU/hr sf F	178	13899	140	101407.412	24	33	80314670.9	0.07	5622026.97
PR HX-6	Rectifying Column reboiler	BTU/hr sf F	130	1089	250	10362.1375	24	33	8206812.9	0.07	574476.90
PR HX-3	Beer column Condenser	BTU/hr sf F	92	880	234	5546.5696	24	33	4392883.12	0.07	307501.82
PR HX-5	Rectifying column condenser	BTU/hr sf F	157	4146	235	44785.2417	24	33	35469911.4	0.07	2482893.80
				m3/hr		KW	24				
PR P-1	Beer Column Bottom pump	gpm	5053	1146.020		0.227655178	24	33	180.3029008	0.07	12.62
PR P-2	Beer Column Reflux	gpm	12	2.7216		5.40E-04	24	33	0.427944058	0.07	0.03
PR P-3	Rectification Column Bottoms Pump	gpm	154	34.9272		6.94E-03	24	33	5.49561983	0.07	0.38
PR P-4	Rectification Column Reflux Pump	gpm	437	99.1116		1.97E-02	24	33	15.5932116	0.07	1.09
PR P-5	Scrubber Bottoms Pump	gpm	69	15.6492		3.11E-03	24	33	2.46251656	0.07	0.17
								Total Pumping Electricity	**128384480**		
Total											**9,733,492**
Total Purification Cost						USD	10,890,418				
						AUD	**10,291,445**				

Appendix B.6- Overall Costs Summary

Table B.6.1. Cost Summary Sheet for SHF

Estimation of Capital Investment Costs					
Fixed Capital Investment (Direct+ Indirect costs)			**Raceway Pond**	**Horizontal Tubular**	**External Loop reactor**
Cultivation			$734,104,818.12	$2,721,254,314.81	$3,617,225,716.20
Dewatering			$23,262,348.78	$4,819,914.58	$5,399,071.40
Pre-treatment			$37,647,991.78	$37,647,991.78	$37,647,991.78
Enzymatic Hydrolysis	SHF		$39,714,177.82	$39,714,177.82	$39,714,177.82
Fermentation	SHF		$265,685,594.47	$265,685,594.47	$265,685,594.47
Purification			$38,122,909.75	$38,122,909.75	$38,122,909.75
Total			$1,138,537,840.72	$3,107,244,903.21	$4,003,795,461.42
Working Capital		5%	$56,926,892.04	$155,362,245.16	$200,189,773.07
Total Capital Investment			**$1,195,464,732.76**	**$3,262,607,148.37**	**$4,203,985,234.49**
Estimation of Total Product Costs					
Variable Operating Cost (Utilities+Raw Materials)			**Raceway Pond**	**Horizontal Tubular**	**External Loop reactor**
Cultivation			$24,068,648.06	$10,809,114.01	$10,290,216.21
Dewatering			$38,597,831.78	$4,951,887.17	$5,937,805.14
Pre-treatment			$13,985,762.65	$13,985,762.65	$13,985,762.65
Enzymatic Hydrolysis	SHF		$1,273,788.38	$1,273,788.38	$1,273,788.38
Fermentation	SHF		$2,551,159.86	$2,551,159.86	$2,551,159.86
Purification			$9,198,150.46	$9,198,150.46	$9,198,150.46
Total Variable Operating Cost			**$89,675,341.20**	**$42,769,862.54**	**$43,236,882.72**
Fixed Operating Cost					
Process Labour			**No. of staff**	**Salary ($/yr)**	**Cost ($/yr)**
Operators/ Shift			3		
Shift Teams			4		
Total Shift Operators			12	$45,000.00	$540,000.00
Process/Maintenance Engineer			1		
Shift Teams			2		
Total Engineers			2	$65,000.00	$130,000.00
Administration			2	$35,000.00	$70,000.00
Labour			2	$30,000.00	$60,000.00
Security			3	$25,000.00	$75,000.00

Total Operating Labour					$875,000.00
Direct Supervisory	Mollina		0.2		$9,000.00
		% Fixed Cap/% Labour	**Raceway Pond**	**Horizontal Tubular**	**External Loop**
Maintenance and Repairs	Mollina	0.04	$14,114,643.14	$38,746,509.35	$49,727,195.87
Operating Supplies	Mollina	0.004	$307,958.82	$91,919.12	$97,255.29
Plant Overheads	Mollina	0.55	$8,249,253.73	$21,796,780.14	$27,836,157.73
contigency	Mollina	0.05	$4,483,767.06	$2,138,493.13	$2,161,844.14
Property Taxes	Mollina	0.016	$1,665,567.09	$1,305,732.66	$1,488,981.34
Fixed Operating Cost (E)			**$119,380,531.04**	**$107,733,296.94**	**$125,432,317.08**
Annual Production Cost			$327,456,505.96	$480,480,733.94	$594,062,374.48
Production Rate (Ethanol)		Liter/yr	9,863,302.68		
Production Cost		AUD/L product	$33.20	$48.71	$60.23

Table C.6.2 Cost Summary Sheet for SSF

Estimation of Capital Investment Costs					
Fixed Capital Investment (Direct + Indirect costs)			**Raceway Pond**	**Horizontal Tubular**	**External Loop reactor**
Cultivation			$734,104,818.12	$2,721,254,314.81	$3,617,225,716.20
Dewatering			$23,262,348.78	$4,819,914.58	$5,399,071.40
Pre-treatment			$37,647,991.78	$37,647,991.78	$37,647,991.78
Fermentation	SSF		$293,960,602.02	$293,960,602.02	$293,960,602.02
Purification			$38,122,909.75	$38,122,909.75	$38,122,909.75
Total			$1,127,098,670.45	$3,095,805,732.94	$3,992,356,291.15
Working Capital		5%	$56,354,933.52	$154,790,286.65	$199,617,814.56
Total Capital Investment			**$1,183,453,603.97**	**$3,250,596,019.59**	**$4,191,974,105.70**
Estimation of Total Product Costs					
Variable Operating Cost (Utilities+Raw Materials)			**Raceway Pond**	**Horizontal Tubular**	**External Loop reactor**
Cultivation			$24,068,648.06	$10,809,114.01	$10,290,216.21
Dewatering			$38,597,831.78	$4,951,887.17	$5,937,805.14
Pre-treatment			$13,985,762.65	$13,985,762.65	$13,985,762.65
Fermentation	SSF		$404,986.65	$404,986.65	$404,986.65
Purification			$9,198,150.46	$9,198,150.46	$9,198,150.46

Total Variable Operating Cost			**$86,255,379.59**	**$39,349,900.94**	**$39,816,921.11**
Fixed Operating Cost					
Process Labour			**No. of staff**	**Salary ($/yr)**	**Cost ($/yr)**
Operators/ Shift			3		
Shift Teams			4		
Total Shift Operators			12	$45,000.00	$540,000.00
Process/Maintenance Engineer			1		
Shift Teams			2		
Total Engineers			2	$65,000.00	$130,000.00
Administration			2	$35,000.00	$70,000.00
Labour			2	$30,000.00	$60,000.00
Security			3	$25,000.00	$75,000.00
Total Operating Labour					$875,000.00
Direct Supervisory			0.2		$9,000.00
		% Fixed Cap/% Labour	**Raceway Pond**	**Horizontal Tubular**	**External Loop**
Maintenance and Repairs	Mollina	0.04	$13,556,681.78	$13,556,681.78	$13,556,681.78
Operating Supplies	Mollina	0.004	$297,655.72	$297,655.72	$297,655.72
Plant Overheads	Mollina	0.55	$7,942,374.98	$7,942,374.98	$7,942,374.98
contigency	Mollina	0.05	$4,312,768.98	$1,967,495.05	$1,990,846.06
Property Taxes	Mollina	0.016	$1,601,755.47	$851,267.81	$858,740.14
Fixed Operating Cost (E)			**$114,850,616.52**	**$64,849,376.28**	**$65,347,219.78**
Annual Production Cost			**$315,179,359.15**	**$429,849,580.98**	**$526,230,044.89**
Production Rate (Ethanol)		Liter/yr	**7,880,593.38**		
Production Cost		AUD/L product	**$39.99**	**$54.55**	**$66.78**

Table C.6.3 Cost Summary on net production cost incl lipid extraction

			Raceway Pond
Cultivation			$734,104,818.12
Dewatering			$23,262,348.78
Lipid extraction			$14,852,515.59
Pre-treatment			$37,647,991.78
Enzymatic Hydrolysis	SHF		$39,714,177.82
Fermentation	SHF		$265,685,594.47
Purification			$38,122,909.75
Fixed Capital Investment (Direct+Indirect costs) - A			$1,153,390,356.31
Working Capital		0.05	$57,669,517.82
Total Capital Investment			**$1,195,464,732.76**
Net production cost			
Variable Operating Cost (Utilities+Raw Materials)			**Raceway Pond**
Cultivation			$24,068,648.06
Dewatering			$38,597,831.78
Lipid extraction			$2,815,040.34
Pre-treatment			$13,985,762.65
Enzymatic Hydrolysis	SHF		$1,273,788.38
Fermentation	SHF		$2,551,159.86
Purification			$9,198,150.46
Total Variable Operating Cost (C + D)			**$92,490,381.54**
Fixed operating cost			**$119,380,531.04**
Fixed annual year			
Cultivation			$77,081,771.58
Dewatering			$2,698,432.46
Lipid extraction			$2,171,939.07
Pre-treatment			$3,967,819.04
Enzymatic Hydrolysis	SHF		$5,555,549.16
Fermentation	SHF		$28,049,018.57
Purification			$1,048,042.91
Total fixed capital per year			**$120,572,572.79**
Net produdction cost			$332,443,485.36
Production Rate (Ethanol)		Liter/yr	9863302.68
Production Cost		AUD/liter product	$33.71

Appendix B.7 Profitability Analysis

Table B.7.1 revenue list

Revenue			**Sales Price**	
Bioethanol	9,863,302.68	L	100	$986,330,268.00
Biomass-Lipid esidue	28079999.90	kg	30	$2,430,000.00
Carbon Credit	81000	ton	25	$2,025,000.00
Total Revenue			AUD	$990,785,268.00

Table B.7.2 economic parameter summary

Fixed capital Investment	$1,138,537,840.72
Working capital	$56,926,892.04
Variable operating cost	$89,675,341.20
Fixed Operating cost	$119,380,531.04

Table B.7.3 cash flow parameter

Year	%FC	%WC	%VCOP	%FCOP	%Rev
1	30%	0%	0%	0%	0%
2	60%	0%	0%	0%	0%
3	10%	100%	30%	100%	30%
4	0%	0%	70%	100%	70%
5	0%	0%	100%	100%	100%
6	0%	0%	100%	100%	100%
7	0%	0%	100%	100%	100%
8	0%	0%	100%	100%	100%
9	0%	0%	100%	100%	100%
10	0%	0%	100%	100%	100%

Table B.7.4 cumulative cash flow, Interest rate = 10%, project lifetime 10 years, government income tax = 30 %

Project Year	Fixed Capital ($US)	Working Capital ($US)	Operating Costs ($US)	Sales Revenue ($US)	Gross Profit ($US)	Depreciation ($US)	Taxable Income ($US)	Tax Paid ($US)	Cash Flow after Tax ($US)	Discounted Cash Flow ($US)	Present Value ($US)
0	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00	$0.00
1	$341.56	$0.00	$0.00	$0.00	$341.56	$0.00	$0.00	$0.00	$341.56	$341.56	$341.56
2	$683.12	$0.00	$0.00	$0.00	$683.12	$0.00	$0.00	$0.00	$683.12	$621.02	$962.58
3	$113.85	$56.93	$146.28	$297.24	$19.83	$0.00	$0.00	$0.00	$19.83	$16.39	$978.97
4	$0.00	$0.00	$182.15	$693.55	$511.40	$113.85	$397.54	$119.26	$392.13	$294.62	$684.35
5	$0.00	$0.00	$209.06	$990.79	$781.73	$113.85	$667.88	$200.36	$581.37	$397.08	$287.27
6	$0.00	$0.00	$209.06	$990.79	$781.73	$113.85	$667.88	$200.36	$581.37	$360.98	$73.71
7	$0.00	$0.00	$209.06	$990.79	$781.73	$113.85	$667.88	$200.36	$581.37	$328.17	$401.88
8	$0.00	$0.00	$209.06	$990.79	$781.73	$113.85	$667.88	$200.36	$581.37	$298.33	$700.21
9	$0.00	$0.00	$209.06	$990.79	$781.73	$113.85	$667.88	$200.36	$581.37	$271.21	$971.42
10	$0.00	$0.00	$209.06	$990.79	$781.73	$113.85	$667.88	$200.36	$581.37	$246.56	$1,217.98