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Techno-Economic Analysis of Protein Concentrate Produced by Flash Hydrolysis of Microalgae

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TECHNO-ECONOMIC ANALYSIS OF PROTEIN CONCENTRATE PRODUCED BY FLASH HYDROLYSIS OF MICROALGAE

by

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ABSTRACT

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Process simulation and techno-economic analysis of 95wt.% protein concentrate from microalgae has been performed using SuperPro Designer v. 9.0. This work, first of its kind, is focused on the economic analysis of protein concentrate that includes processes such as microalgae cultivation, harvesting, protein extraction and drying steps. A baseline capacity of 160 MT/day protein concentrate production on commercial basis has been analyzed. This throughput requires 336 MT/day dry algae (54 wt.% protein). The amount of carbon dioxide required to grow this quantum of algae is estimated to be 648 MT/day, which is produced from an *in situ* 21 MW power plant run by approximately 12 MT/h natural gas (methane).

The economic feasibility study has been performed for the entire process. It became clear that decreasing the amount of water of the microalgae biomass slurry to the flash hydrolyzer reduces the fixed capital investment (FCI) and the annual operating cost (AOC). The baseline production of protein concentrate reveals the following results: FCI: $264 million; AOC: $145 million; capital recovery: $180 million/year for 15 years; unit cost of production: $2.86/kg protein depending on the algae slurry density; minimum selling price: $4.13/kg protein; power requirement: 19.5 MW; Land requirement: 7177 acres; water: 15576 MT/day (4.1MGD).

Further analysis revealed that the major contributors to the financial statue of this work is contingent on the algae slurry going to the flash hydrolysis, protein content of the microalgae, pond depth for algae cultivation, and algae productivity.

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NOMENCLATURE

AFC	Annualized Fixed Cost
AOC	Annual Operating Cost
DC	Direct Cost
DCF ROI	Discounted Cash Flow Return on Investment
DO	Dissolved Oxygen
FCI	Fixed Capital Investment
FC	Fixed Capital
FH	Flash Hydrolysis
IC	Indirect Cost
IRR	Internal Rate of Return
MEA	Monoethanolamine
NPV	Net Present Value
OC	Other Cost
PAR	Active Radiation
PB	Payback Period
PBR	Closed Photobioreactors
PC	Purchase Cost
PE	Photosynthetic Efficiency
ROI	Return on Investment
TAC	Total Annualized Cost
TCI	Total Capital Investment

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CHAPTER 1

1.1 Background and Justification

Microalgae research has been in the forefront of the energy research since the inception of fossil-fuel-depletion awareness. This particular biological entity has received attention because of its high levels of oil, protein and carbohydrate. Moreover, it has remarkable potential utilization of poor quality land and water, and acts as a deep sink for carbon dioxide from energy-producing sector such as coal-fired power plants [Quinn et al., 2014].

Besides its high energy content, microalgae, which are classified as heterogeneous organisms possess both food and biological ingredients. Phaeophyceae are well-known for having an important class of phenolic compounds and phlorotannins, which are strong sources of bioactivities including antioxidant, anti-inflammatory, antidiabetic, anti-proliferative or antibacterial effects [Sánchez-Camargo et al., 2015].

Scenedesmus obliquus, another kind microalgae, contains considerable level of astaxanthin (3, 30-dihydroxy-b, β-carotene-4, 40-dione) that is a natural ketocarotenoid pigment which has been widely used in feed as colorant. It has superior antioxidative activity, potential inhibitory action to the proliferation of some cancer cells and correlation with the enhancement of T-cell activity in human [Qin et al, 2008]. Tibbetts et al. reported a general composition of microalgae: ash (5–17 %), moderate to high carbohydrate (18–46 %), crude protein (18– 46 %), high crude lipid (12- 48 %), and energy (19–27 MJ kg^{-1}). Other reporters quoted that microalgae has high protein content of 39-71% dry mater; pigments, and other bioactive constituents like dietary fibres (as high as 74.6% on dry basis) in some species; carotenoids, carbohydrates, omega-3 fatty acids, which have tremendous use in the pharmaceutical industries [Balasubramanian et al., 2011]. Furthermore, Chlorella vulgaris has been reported to have the following essential amino acid (%wt), the framework of proteins: aspartic acid 8.6, threonine 5.5, serine 4.4, glutamic acid 10.3, proline 5.0, glycine 7.0, alanine 10.7, valine 6.7, methionine 2.6, cysteine 1.3, isoleucine 3.4, leucine 8.2, tyrosine 4.4, phenylalanine 6.0, histidine 1.6, lysine 5.4, arginine 7.4, tryptophan 0.2, ammonia 1.3 [Ursu et al., 2014]. Table 1 highlights composition of different species of microalgae. Generally, the high contents of proteins in microalgae have undoubtedly proven that it is a prominent candidate for the production of peptide or protein concentrate for both food and pharmaceutical industries apart from being bioenergy source.

Despite the immense bioactive components in the algae, less research has gone into extracting these component. To extract these essential components, for example, protein, the cell wall of the algae needs to be breached. Because of this, bead milling [Doucha et al. 2008, & Lee et la. 2011], ultrasonication [Furuki et al.2003, Gouveia et al. 2009, Gerde et al.2012], microwave radiation [Zheng, et al. 2011], enzymatic treatment [Fleurence et al. 1999, Sari et al. 2013], cell homogenizer [Mendes-Pinto et al. 2001] and high-pressure cell disruption [Jubeau et al. 2012] have been reported.

However, all these methods are tedious and time consuming. Garcia-Moscoso et al. reported of protein extraction from microalgae via flash hydrolysis (FH) at the laboratory level. This method proved to be most efficient in extracting the protein from the microalgae in 10 seconds. Flash hydrolysis in subcritical water (below 374^{o}C and 22.1 MPa) extracted proteins efficiently and produced lipid-rich biofuel intermediates from wet microalgae (*Scenedesmus sp.*) in a continuous flow process. However, there has been no studies related to techno-economic analysis of protein extraction.

In this study, the results from the laboratory level studies were used to develop a techno-economic analysis of FH process when *Scenedesmus obliquus* (17 % lipid, 23 % Carbohydrate, and 54% protein) was used as a feedstock. Flash hydrolysis is carried out in a subcritical water at temperature below 374°C and pressure 22.1 MPa. This work focuses on modelling an industrial and commercial protein concentrate suitable for the both food and pharmaceutical industries, beginning from the algae cultivation to the protein extraction stage, and finally perform economic analysis.

Table 1.General Composition of Different Algae (% of dry matter)

Algae	Protein	Carbohydrate	Lipids
Anabaena cylindrical	43-56	25-30	4-7
Aphanizomenon flos-aquae	62	23	3
Chlamydomonas rheinhardi	48	17	21
Chlorella pyrenoidosa	57	26	2
Chlorella vulgaris	51-58	12-17	14-22
Dunaliella salina	57	32	6
Euglena gracilis	39-61	14-18	14-20
Porphyridium cruentum	28-39	40-57	9-14
Scenedesmus obliquus	50-56	10-17	12-14
Spirogyra sp.	6-20	33-64	11-21
Arthrospira maxima	60-71	13-16	6-7
Spirulina platensis	46-63	8-14	4-9
Synechococcus sp.	63	15	11

Source: Becker, 2007

Table 1.1 General Productivity of different Microalgae

Algal species	Raceway pond depth (m)	Biomass Productivity ($g/m^2/day$)
Scenedemus sp.	0.2	17
Tetraselmis MUR 233	0.2	29.6 (max. -37.5
Spirulina Platensis	0.3	8.2 (max -13.95)
Nannochloropsis sp	0.15-0.2	14.1
Dictyosphaerium sp.	0.3	5.8

Source: Kumar et al. 2015.

1.2 Research Aims and Objectives

The following are the research aims and objectives:

- To study the simulation of production of concentrated protein from microalgae through flash hydrolysis.
- To conduct techno-economic analysis of production of concentrated protein from microalgae at industrial/commercial scale.
- To discuss the inherent bottlenecks that hinder the feasibility of this novel process.

1.3 Limitations

The research does not incorporate the packaging, distribution, and transportation cost of the final product, the protein concentrate. Hence the prices employed in this work are not a true reflection of the protein concentrates on the market. Two kinds of power sources are suggested in this work: on-site power production and outside power. The cost of infrastructure for natural gas transportation and electricity transmission to the plant site was not included.

CHAPTER 2

OVERVIEW OF MICROALGAE

2.1 Algae Cultivation

Algae are considered as potential feedstock candidates with higher productivity per unit land area as compared to traditional lignocellulosic biomass [Griffiths and Harrison, 2009]. They possess greater control of nutrient use, ability to receive and metabolize concentrated carbon dioxide from industrial sources, and consequently avoid competition with arable crops [Lardon et al., 2009]. There are many types of algal culture systems that have been built or proposed. Table 2 delineates different algal cultivation methods. Besides, the table shows the respective yields and cost of production quoted from literature.

Table 2 Algal culture systems, types, yields, and cost of production estimates.

System	Types	Yield (dry Mg/ha/yr)	Cost $/kg dry biomass
Ponds	Open Raceway Circular with Mixing Large Open	7-135[a]	0.6-3.80[b]
Closed Photobioreactors (PBR)	Tubular Flat Panel Column	70-150	0.47[b]-34[d]
Emerging Technologies	Open thin-layer panel Polymer bags Immobilized bed	Not reported	25-600[e]

a-Moheimani and Borowitzka(2006); b-Chisti (2007); d-Grima etal.(2003); e-Borowitzka (1999).

Though open pond systems require large acres of land and water to thrive, it is the least capital intensive. Conversely, closed system (PBR) lend itself to greater process control, but possesses higher capital cost. Emerging technologies on the other hand might offer better options to ponds and PBR, maybe yielding lower capital cost and higher cellular densities, giving total lower production cost. However, components such as polymer bags and immobilized bed are difficult to estimate. Besides, algae productivity and harvesting could be higher, but inherent hurdles such as high material cost, difficult scale-up, and proper strain identification for immobilized growth restrict the implementation and progress of emerging technologies [Katrina et al., 2012]. Due to the high capital demand open raceway pond are employed in this study.

2.2 Water Resources

The success of microalgae cultivation is contingent partly on reliable water supply. Due to the continuous evaporation of water from the open pond, make-up water needs to be supplied. PBR also require water for cooling purposes. There have been propositions that low competitive water, such as seawater and brackish water, could be used for algae cultivation. However, these sources require pre-treatments which results in high energy demand for the whole process. Moreover, water recycling has the potential of reducing consumption and nutrient loss, but it comes with greater risk of bacteria-fungi-virus infection and inhibition. Additionally, non-inhibitors such as organic and inorganic chemicals and remaining metabolites from destroyed algae cells are found in the recycled water [Slade et al., 2013]. In this work, it is assumed that 98% of the water for algae cultivation is recycled.

2.3 Land Use and Location

Marginal land use has been suggested to be one of the advantages of algae cultivation since this limits its competition for food production. However, topographic and soil constraints limit the construction of raceway pond systems since they require flat terrain. Moreover, soil porosity calls for the need to line these ponds with polymeric and sealing materials thereby increasing cost of construction. Apart from land use, solar radiation required for algae cultivation is determined by the location of the pond. For practical purposes, suitable pond locations are warm countries near the equator (see Table 3) where insolation is not less than 3000 h/yr [Slade et al., 2013].The average amount of solar radiation that reaches our planet every second, E_{solar}, is about 1367 W/m^2, defined as the solar constant [Holtermann et al., 2011]. From Table 3, $q^n_{solar,max}$ is maximum irradiation intensity; q^n_{solar} is the annual amount of irradiation; v is intensity of irradiations which describes the regularity of solar irradiation.

Table 3 Solar irradiation data for different locations

Location	$q^n_{solar,max}$ (W/m^2)	q^n_{solar}(kWh/m^2)	v
Bergen(Norway)	828	785	-
Helsinki(Finland)	906	970	0.1338
Stuttgart(Germany)	974	1126	0.1419
Madrid(Spain)	-	1657	0.1943
Lisbon(Portugal)	1010	1726	-
Rabat (Morocco)	-	1837	0.2076
Sahara Desert	-	2350	-

Source: Holtermann et al., 2011

Weyer et al reported that photosynthetic productivity is contingent on the intensity of the solar irradiation. Only the light within the wavelength range of 400 to 700 nm, known as photosynthetically active radiation (PAR), can be used by plants and algae, which practically means that only 40 to 45% of total solar energy can be utilized for photosynthesis. Another group reported that the theoretical maximum for photosynthetic efficiency (PE) is between 8 and 11% of the total solar energy [Brennan and Owende, 2010; Hindersin et al., 2013].

However, typical PE values of cultivated microalgae is reported to be in the range of 4 to 7% under optimized condition [Doucha and Livansky, 2006, 2009; Hase et al., 2000; Morita et al., 2002].

2.4 Nutrients

The necessary nutrients for algae growth are primarily nitrogen, phosphorus, and potassium [Slade et al., 2013]. Others are calcium, magnesium and sulfur, which are necessary because biological molecules do not consist of carbon and water only. By assimilation, inorganic nutrients are converted into organic compounds to form part of the organisms' biomass [Holtermann et al., 2011]. Moreover, since dry algal biomass consist of 7 wt% nitrogen and 1 wt% phosphorus, fertilization has become highly indispensable [Wijffels et al., 2010]. Ammonium nitrate and phosphate is used in this work as it contains nitrogen and phosphorus that essential nutrients for plants.

2.5 Algae Bloom

Excess nutrients input can lead to excessive algae growth. This will lead to deficiency of oxygen leading algae decomposition and eutrophication. Dead algae eventually precipitate and finally settle at the bottom of the water or any natural water or lake forming colloidal nutrient [Chipman et al., 2010, Sun et al., 2014]. Gao et al. and Zhu et al. reported that algae bloom causes a higher availability of P, Fe, and S. Sharp et al. and Shen et al. reported that algae bloom changes the physical and biological outlook of the benthic environment such as DO, pH, and Eh, particulate matter, which eventually affect the nutrient cycling.

2.5 Carbon Dioxide

Carbon content in microalgae emanates from atmospheric carbon dioxide. Moreover, there is a direct relationship between biomass output and CO_2 consumption [Holtermann et al., 2011]. For example, 1 ton dry algae containing 50% carbon by mass consumed 1.83 ton CO_2. However, in reality, CO_2 supply will be several factor of 1 tonne. For raceway pond, the outgassing is a function of the depth, friction coefficient of the lining, mixing velocity, pH and alkalinity. Depending on the operating conditions, the theoretical efficiency can range from 20% to 90% [Weissman et al., 1998]. Practically, the efficiency of CO_2 fixation in open raceways may be less than 10%; roughly 35% for thin layer cultivation the efficiency of CO_2 [Slade et al., 2013]; approximately 75% in closed tubular PBRs [Acie´n et al., 2012].

The source of CO_2 can adversely affect the overall production cost of the process. It can also affect the choice of location of plant. CO_2 supply from flue gas has been reported to cheaper than using raw CO_2. It takes 3.7GJ of energy to absorb 1 tonne of CO_2 from flue gas using monoethanolamine (MEA). About 370 kg of CO_2 is released during CO_2 absorption and regeneration of the MEA solution [Lam et al., 2012]

It is also advisable to site the flue gas source close to the algae production site. Since the cost of separating the CO_2 from the flue gas is costly, the flue gas is directly fed to the algae pond. This injection does not affect the algae growth because the algae can tolerate the contaminants in the flue gas [Slade et al., 2013]. In this work CO_2 is supplied by flue gas from an in situ power production that is powered by natural gas. Natural gas has been chosen because its flue gas is cleaner than that of coal that contains a lot of heavy metals (e.g. mercury), which can eventually contaminate the protein concentrate.

2.6 Microalgae Harvesting

Being one the technological steps in microalgae recovery from dilute algae culture, microalgae harvesting has been reported to have tasked the financial aspect in the bioenergy domain. Harvesting step contributes to 20-30% of the cost of microalgae production [Rawat et al., 2011]. The micro size of the algae grown in a very dilute culture (concentration less than 1 g/L) is the reason behind the high cost of harvesting [Danquah et al., 2009; Molina et al., 2003]. Moreover, microalgae possess a negative surface charge and their cells have algogenic organic matter that render them stable in a dispersed condition [Danquah et al., 2009]. Presently, there is no single economically viable and efficient microalgae harvesting method in the algae industry [Christenson et al., 2011]. However, combination of two or more harvesting methods can reduce cost of production [Schlesinger et al., 2012]. Table 4 highlights different methods, advantages and disadvantages of algae harvesting.

2.7 Microalgae Conversion

Microalgae is presently basically a source for biofuels, but to a lesser extent for bioactive components, such as protein. There are two general techniques for microalgae conversion: thermochemical and biochemical [Tsukahara and Sawayama, 2005]. While thermochemical conversion utilizes heat to decompose organic compounds in the algae, biochemical conversion employs microorganisms to produce biofuel. Thermochemical conversion can be subdivided into gasification, liquefaction, pyrolysis, and direct combustion. Biochemical on the other hand can be subdivided to into anaerobic digestion, alcoholic fermentation and photobiological hydrogen production. Other emerging conversion technologies are transesterification (acid/base) catalysis and photosynthetic microbial fuel cell are under research [Tan et al., 2015].

Table 4 Microalgae harvesting methods

Harvesting method	Advantages	Disadvantages
Chemical Coagulation/ flocculation	• Simple and fast method. • No energy requirement.	• May be expensive and toxic to algae. • Culture medium recycle is limited.
Auto and bioflocculation	• Inexpensive • Culture medium recycle is permitted. • Non-toxic to microalgae biomass.	• Cellular composition is affected. • Possibility of microbiological contamination.
Gravity Sedimentation	• Simple and inexpensive.	• Time-consuming. • Possibility of biomass deterioration. • Low concentration of algal cake.
Flotation	• Feasibility for large scale application. • low cost • Low space requirement. • Short operation time.	• Requires the use of flocculants. • Not applicable to marine algae harvesting.
Electrical based processes	• Applicable to wide variety of algae sp. • Do not require the use of flocculants.	• Poorly disseminated. • High energy and equipment cost.
Filtration	• High recovery efficiency. • Allow separation of shear sensitive sp.	• Possibility of fouling/clogging. • Require regular membrane change. • High cost of pumping and membrane.
Centrifugation	• Fast method. • High recovery efficiency • Suitable for almost all microalgae sp.	• Expensive. • High energy requirement. • Possibility of cell damage to high shear forces. • Suitable for high-valued product recovery.

Source: Ana et al., 2015

2.8 Liquefaction

This reaction converts wet algae biomass to bio-oil at low temperature and high pressure by either employing a catalyst or not. In this case energy intensive drying process is unnecessary. This process does not only convert lipids but also carbohydrates and protein in the algae into biocrude oil [Biller and Ross, 2011]. The efficiency of thermochemical liquefaction is contingent on reaction temperature, retention time, catalyst, and composition of biomass (liquid, carbohydrate, and protein) [Yang et al., 2004].

Hydrothermal liquefaction is a type of liquefaction process that employs subcritical water at medium temperature of 280-370^{o}C and pressure range of 10-25 MPa. This also converts wet biomass into liquid biocrude as the main product. (Patil et al., 2008).

2.9 Anaerobic Digestion

Anaerobic digestion is one of the biochemical conversion methods that utilize microorganisms to convert algae biomass into biogas comprising CH_4 and CO_2 with small amount of H_2S. Biogas has energy content of 20-40% of the original lower heating value of the biomass. The optimum moisture content of biomass that is suitable for anaerobic digestion is in the range of 80-90% dry weight [Brennan and Owende, 2010].The three anaerobic stages are: *hydrolysis* of polysaccharides; *fermentation* (sugar to alcohol acetic acid, volatile fatty acid and mixture of H_2 and CO_2); *methanogenesis* (conversion of gas mixture into CH_4 (60-70%) and CO_2 (30-40%)) [Cantrell et al., 2008)]. Knowing the carbon, hydrogen, oxygen and nitrogen content of the biomass, the theoretical production of methane can be illustrated by the following stoichiometric equation [Ward et al. 2014]:

$$(C_aH_bO_cN_d) + \left(\frac{4a-b-2c+3d}{4}\right)H_2 \longrightarrow \left(\frac{4a+b-2c-3d}{8}\right)CH_4 + \left(\frac{4a-b+2c+3d}{8}\right)CO_2 + dNH_3$$

where a, b, c, and d equal the carbon, hydrogen, oxygen, and nitrogen contents on molar basis respectively. Methane yield (litres/g (VS) destroyed) is found as follows:

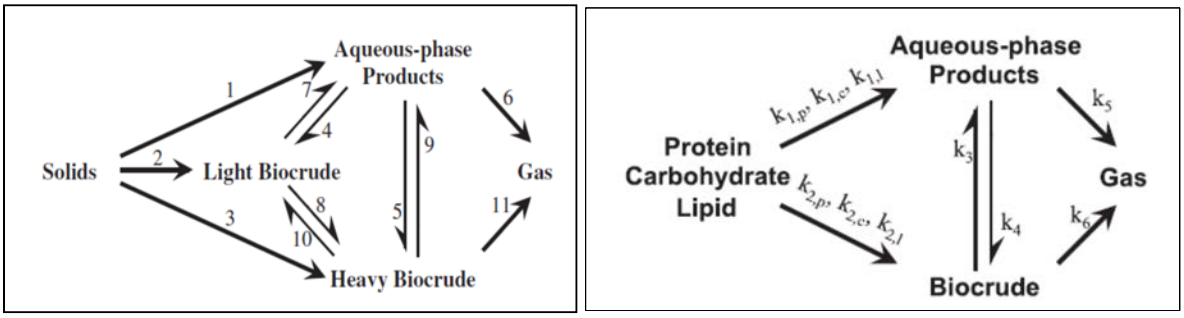
$$\left(\frac{4a+b-2c-3d}{12a+b+16c+14d}\right) * V_m$$

Where V_m is the molar volume (22.14 L/mol) at 0°C and 1 atm.

2.10 Flash Hydrolysis

From the preceding discussion on the different kinds of reactions, the end products are biofuels (diesel, methane, hydrogen, ethanol, etc.).

Flash Hydrolysis is a reaction between subcritical water (temperature less than 374°C and pressure less than 20 MPa) and algae in a continuous plug-flow reactor with residence time of few seconds. This reaction does not produce biocrude oil; it does produce protein-laden aqueous phase and solid phase (containing biofuel intermediate). The short time does not allow the conversion of protein and carbohydrate to convert to biocrude oil as reported by Valdez and Savage, 2013 and Valdez et al., 2014. Subjecting the microalgae biomass to liquefaction condition for a long time eventually convert most of protein, carbohydrate, and lipid into biocrude oil as shown in Figure 1. This means that to produce more protein in the aqueous phase, the reaction time should be in seconds as reported by Garcia-Moscoso et al.



(a) (b)

Figure1.Reaction network for the hydrothermal liquefaction of *Nannochloropsis sp.*

Garcia-Moscoso et al. reported a flash hydrolysis in a continuous-flow reactor whereby the protein from microalgae (*Scenedesmus sp.*) biomass was hydrolyzed in a very short residence time (few seconds). In their work, flash hydrolysis was conducted at different temperatures (240, 280, and 320°C) and in three different residence times (6, 9 and 12 sec.). They concluded that the maximum yield of protein in the aqueous phase was approximately 82 % at 320°C in 6 seconds. Based on this work, it has been proposed that extraction and concentrating protein from microalgae is feasible and environmentally friendly on commercial scale as compared to solvent extraction methods.

CHAPTER 3

COST OVERVIEW

3.0 Overview of Process Economics

Process economics is an indispensable component of every new product or process design. In designing a new process or product, many of the technical and environmental decisions are strongly influenced by economic factors. It is therefore necessary to discuss the economic aspect of this simulation. The knowledge of economics will assist in evaluating the feasibility of the process, making improvements, comparing alternatives, making design and operating decisions, etc. These decisions can be made based on (1) cost and estimate of operation; (2) depreciation; (3) break-even analysis (i.e. total production cost equals process revenue); (4) time value of money; (5) profitability analysis [El-Halwagi and Mahmoud, 2012].

3.1 Cost Types and Estimation

There are two basic cost used to make decisions in process economics: *capital investment* and *operating cost*. The total capital investment (TCI) or capital cost is the money required to purchase and install the plant and its accessories and to provide the requisite expenses needed to start the process operations. With the plant in operation, the money required to continue or run the operation is known as the *operating cost*. The basis for estimating these two costs are (i) capital (fixed, working, and total), (ii) equipment, (iii) operating, and (iv) production (total annualized cost). The fixed capital investment/cost is the money required to pay for the processing equipment and the auxiliary units, acquiring and preparing land, civil structures, facilities, and control systems. On the other hand, the working capital is the money needed to pay for the operating expenses until the product is sold. It also includes the money needed to stockpile raw materials. The following outlines the estimation of fixed capital cost. The fixed capital (FC) is estimated based on total equipment purchase cost (PC).

FC = Direct Cost (DC) + Indirect Cost (IC) + Other Cost (OC). The rest of the cost analysis can be found in the appendix.

The word estimate implies, there is a level of uncertainty in most cost estimates. These uncertainties emanate from the method or source of cost acquisition. The most commonly used methods are (i) manufacture's quotation, (ii) computer-aided tools, (iii) capacity ratio with exponent, (iv) updates using cost indices, (v) factor based on equipment cost, (vi) empirical correlations, and (vii) turnover ratio [El-Halwagi and Mahmoud, 2012].

Capacity ratio with exponent can be evaluated from the following equation.

$$FCI_B = FCI_A \left(\frac{Capacity_B}{Capacity_B}\right)^x$$

where FCI_B and FCI_A are the fixed capital investments of plant B and A respectively, and $Capacity_B$ and $Capacity_A$ are the capacities (for example, flow rate of main product) of plants B and A respectively, the exponent *x* is usually less than 1 (taken to be 0.6-0.7). This relation can also be applied to equipment estimation if the sizes of the equipment are known and the cost of one them is known.

Cost estimates are made and reported for a given time. With inflation and price fluctuation, it is necessary to account for fixed capital cost as a function of time. Cost indices are very useful in adjusted cost estimates based on time. Updates using cost indices can be applied using the following equation.

$$FCI_{t2} = FCI_{t1} \left(\frac{Cost\ index\ at\ time\ t2}{Cost\ index\ at\ time\ t1}\right)$$

Where FCI_{t1} and FCI_{t2} are the cost of plant/equipment at times t1 and t2 respectively.

Methods used to estimate the capital cost in the present study are the capacity/size ratio and updates using cost indices. These are imbedded in the SuperPro Designer software.

3.2 Depreciation

Depreciation is the annual income tax deduction that is intended to allow a company to recover the cost of property (for example, ultrafiltration unit) over a certain recovery period [El-Halwagi and Mahmoud, 2012]. This is normally taken from the revenue before tax so that the cost of equipment can be recovered and to perpetuate the use of the property or asset. Land and working capital investment cannot be depreciated because they are recoverable in principle. There are several method to calculate depreciation: (i) linear (straight-line) method; (ii) declining-balance method; (iii) modified accelerated cost recovery system ;(iv) sum-of-years' digit method; (v) sinking fund method. The detail explanation of each is not explained in this work. However, the simplest and most commonly used is the method of the straight-line. In this work, the straight-line method was used in the SuperPro Designer.

3.3 Profitability Analysis

This can be done with or without time-value of money. Profitability criteria without time-value money are (i) return on investment (ROI) and (ii) payback period (PB). Whereas profitability criteria with time-value of money are (i) net present value (NPV); (ii) discounted cash flow return on investment (DCF ROI); (iii) discounted cash flow payback period. Another way of assessing profitability is by comparison of alternatives through (i) NPV; (ii) annual cost/revenue; (iii) total annualized cost; (iv) incremental return on investment.

ROI = $\frac{Annual\ Profit}{Total\ Capital\ investment} \times 100\%$

NPV is the cumulative value (revenues-expenses) adjusted to a reference time. It is found as NPV $=\sum_{N=0}^{N} ACF_N (1+i)^{-N}$, where ACF_N is annual cash flow for year N, i is the discount rate. NPV > 0 means investment is financially attractive; NPV = 0 means the investment is neutral; NPV < 0 means the investment is not financially attractive.

The DCF ROI also known as internal rate of return (IRR) is the value of discount rate that renders the NPV to be zero. The higher IRR value the more attractive the project [Edgar and Himmelblau, 2001]. The total annualized cost (TAC) is equal to the sum of annualized fixed cost (AFC) and the annual operating cost (AOC). This is calculated as TAC = FCI*$\frac{i(1+i)^N}{(1+i)^N-1} + AOC$. Thus the AFC is multiplied by the capital recovery factor and the result is added to the annual operating cost. This gives the annual cost needed to perpetuate the project by taking care of operating cost and capital investment.

CHAPTER 4

SIMULATIONS

4.1 Process Modelling

Scenedesmus obliquus algae was selected for this work since it is the kind grown at Old Dominion Algae Laboratory. Elemental analysis of this algae is as follows: C = 50%, H = 6.2%, N = 9.65 %, P = 1 %, and O = 32 %. This gives the empirical formula for the algae as $C_{133}H_{192}O_{32}N_{22}P$. This formula was used to calculate for the nutrient requirements. Rogers et al. reported similar empirical formula as $C_{106}H_{181}O_{45}N_{16}P$. The following were the principal assumptions made.

4.2 Essential Assumptions

- Algae strain: *Scenedesmus obliquus*
- Elemental composition of algae biomass: $C_{133}H_{192}O_{32}N_{22}P$
- Average annual areal productivity: 15 g m^{-2} d^{-1}
- Biomass Protein Content: 54 wt.%
- Daily peptide production: 160 MT/day
- Protein Extraction efficiency: 85%
- Harvested algae slurry: 20 wt. %
- Dimensions of a pond: 120 m x 10 m x 0.3 m
- Maximum culture density: 0.8 g L^{-1}
- Water recycle rate: 98%

4.3 Process Description

A simplified process flowsheet for 95 wt.% protein concentrate is as shown in Fig.2.

The process was modelled with SuperPro® Designer V. 9.0. Fig. A1 highlights the modelling of the algae cultivation, harvesting and protein extraction. The algae is grown in a raceway pond P-2 with ammonium phosphate and ammonium nitrate as the necessary nutrients. Carbon dioxide is supplied by the flue gas from an integrated cogeneration section. The growth maturity was assumed to 14 days at which the algae concentration will have reached 0.8 g/L. The culture is then pumped by pump P-6 to be filtered by the belt filter BF-101. The wash-out outlet is recycled back to the pond. The cake density is set to 20 wt. % necessary for the flash hydrolysis. The slurry is stored in vessel V-101 awaiting flash hydrolysis. The algae slurry is then pumped by PM-102 through series of heat exchangers HX-102 to be heated from 20^{o}C to 280^{o}C. The slurry is then subjected to flash hydrolysis within 10 seconds in a plug flow reactor PFR-101. The products are cooled to 30^{o}C, and are then sent to belt filter B-102 to remove the solid part of the algae (biofuel intermediate). The pregnant protein solution is then subjected series of ultrafiltration by UF-101 giving 40 wt.% concentrated proteins. The protein concentrate is subjected to spray drying where it allowed to move countercurrent to a hot stream of air (140^{o}C) in the dryer (SDR-101). The dry protein is collected at the bottom of the dryer while the exhaust air (70^{o}C) is sent to the cyclone (CY-101) to recover protein fines.

The filtrates from both BF-102 and UF-101 are sent to anaerobic digester AD-101 for methane production. Since the methane generated is not enough to produce power and carbon dioxide for the entire process, natural gas is employed to supplement the power production. Natural gas is combusted in a boiler SG-101 to produce steam at

6 MPa and 300°C. The steam is sent to a multi-staged power generator T-101 for power production. The flue gas from the boiler SG-10, which contains 18 % carbon dioxide is cooled to 30°C through heat exchanger HX-103 and sent to the algae pond. The steam generated from the power house is used to heat the algae slurry for the hydrolysis.

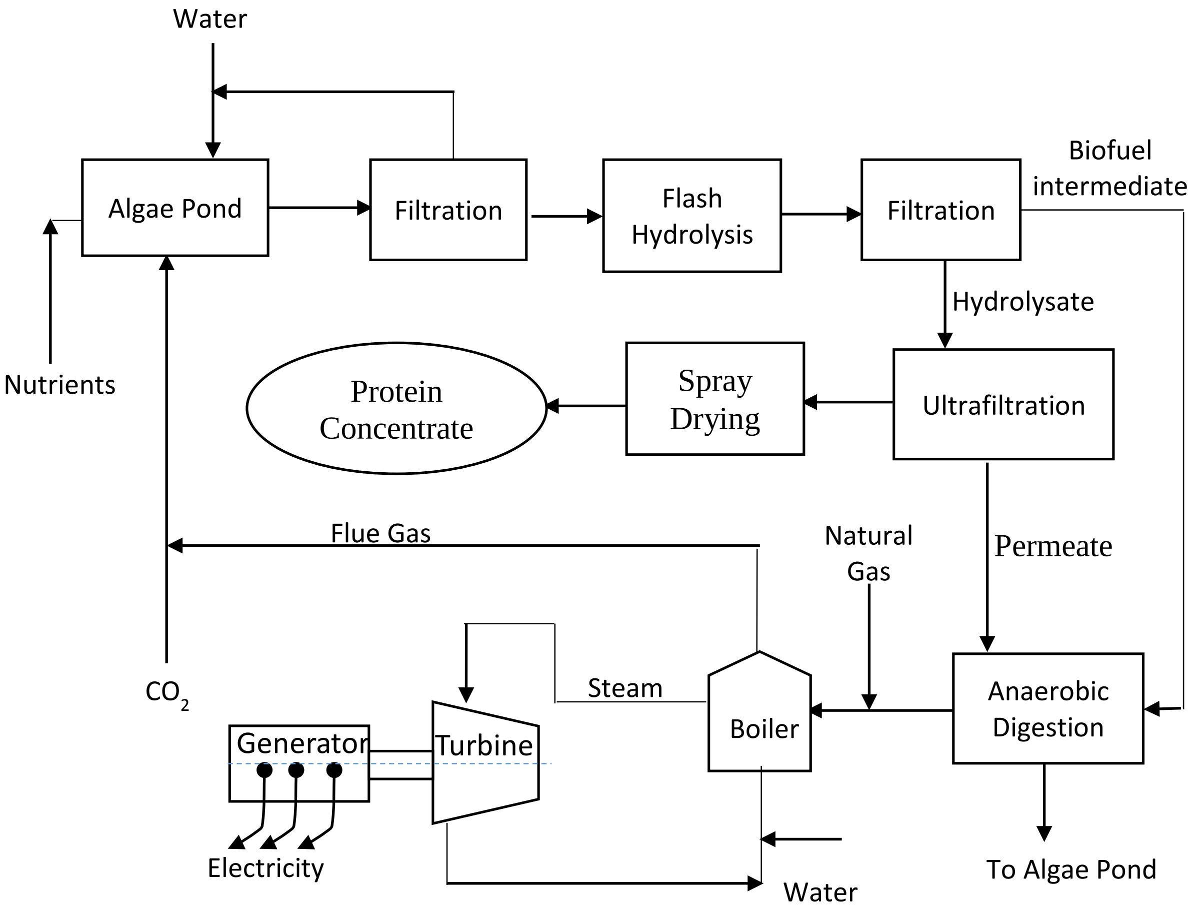


Figure 2. Simplified Process flowsheet for protein concentrate production.

4.4 Economic Analysis

Numerous studies have evaluated the economics of microalgae production, but most of them concentrate on biofuel production (Amer et al., 2011; Benemann, 2013; Davis et al., 2011; Draaisma et al., 2013; Lam and Lee, 2012; Rios et al., 2013; Taylor et al., 2013). Richardson et al. reported Farm-level Algae Risk Model (FARM) and used it to simulate the economic feasibility and probabilistic cost of biomass and bio-crude oil production for two projected algae farms. Rogers et al. also reported sustainability and economic requirements of a 160 MT/day algal biofuel facility based in New Mexico. Since the outcome their work cannot be substituted for the current work, it is necessary to assess the actual financial viability of protein concentrate as opposed to biofuel production.

In this work, it is assumed that the year of construction is 2015; construction period is three years; start-up period is one year; project life is 15 years; inflation (to update equipment cost) is 4%; interest rate is 10%. In financing the project, 30% of the fixed capital investment is provided in the first year, 40% in the second year, and 30% in the third year. Concerning depreciation, straight line method was employed with salvage value being 5% of the fixed capital. Moreover, the operation hours is assumed to be 7920 /year (330days/year). The cost of materials and equipment is obtained from the SuperPro Designer. Other assumed parameters include the following: Pond paddlewheel ($5000/unit); energy requirement for a paddlewheel (0.73 W/m^2); pond liner ($0.77/m^2); Landscaping ($0.16/m^2), raceway covering ($0.98/m^2) [Rogers et al., 2014].

Table 5. Capital Cost estimation for 160 MT/day protein concentrate

Name Description	Quantity	Unit Cost ($)	Total Cost ($M)
MX-101 Mixer	28	2000	0.15
Algae Pond	19138	8000	153.104
PM-101 Centrifugal Pump	3	206000	0.618
PM-103 Centrifugal Pump	1	55000	0.050
V-101 Receiver Tank	1	55000	0.055
PM-102 Centrifugal Pump	3	187000	0.561
PM-104 Centrifugal Pump	1	45000	0.045
M-101 Centrifugal Fan	1	16000	0.016
PFR-101 Plug Flow Reactor	1	75000	0.075
HX-101 Heat Exchanger	6	130000	0.780
UF-101 Ultrafilter	245	147000	36.015
SG-101 Steam Generator	2	975000	1.950
T-101 Multi-Stage Steam Turbine	1	4115000	4.115
MX-102 Mixer	28	2000	0.056
MX-103 Mixer	1	2000	0.002
MX-106 Mixer	1	2000	0.002
HX-102 Heat Exchanger	75	2000	0.15
AD-101 Anaerobic Digester	2	6136000	12.272
MX-106 Mixer	1	2000	0.002
HX-102 Heat Exchanger	8	126000	1.008
HX-103 Heat Exchanger	5	122000	0.610
HX-104 Heat Exchanger	1	54000	0.054
BF-101 Belt Filter	14	280000	3.920
BF-102 Belt Filter	5	280000	1.400
SDR-101 Spray Dryer	1	303000	0.303
CY-101 Cyclone	2	3000	0.006
Land Acquisition (Acres)*	7177	3000	21.530
Equipment installation			27.834
Startup Cost			12.244
Working Capital			6.402
Total			**263.827**

*This is excluded from the fixed capital investment.

Table 6 Annual Operating Cost for 160 MT/day Protein Concentrate

Bulk Material	Unit Cost ($)	Annual Amount (MT)	Units	Annual Cost ($M)
Labor	69.000	88829	hrs	6.129
Ammonium Nitrate*	0.150	46023	MT	6.903
Carbon dioxide*	40.000	3523	MT	0.141
Diammonium Phosphate*	0.500	4744	MT	2.372
Methane*	0.136	76074	MT	10.346
Water*	0.013	5135536	MT	0.668
Dft Membrane	400.000	27413	m^2	10.965
Power (kWh)	0.100	191883086	kWh	19.188
Steam (High P)	20.000	979713	MT	19.594
Cooling Water	0.050	107724417	MT	5.386
Facillity-dependent[a]				62.742
Laboratory/QC/QA[b]				0.919
Waste Treatment[c]				1.099
Total				**146.452**

a Estimate based on capital investment parameters (i.e., maintenance, depreciation and miscellaneous costs).
b This accounts for the cost of off-line analysis, quality control (QC) and quality assurance (QA).
c 1 barrel of peptide concentrate costs $3.33 wastewater treatment.
*These are the main inputs to the SuperPro Designer.

CHAPTER 5

RESULTS AND DISCUSSIONS

In this work, protein concentrate of 160 MT/day commercial facility has been analyzed. This throughput requires 336 MT/day dry algae. The amount of carbon dioxide required to grow these microalgae is estimated to be 648 MT/day, which is produced from an in situ 21-MW power plant run by approximately 12 MT/h natural gas (methane). This means the cost of supplying carbon dioxide to the pond is approximately 4 % of the operating cost. Ketheesan et al. reported that the cost of supply and transfer of CO_2 accounts for nearly one-third of the total algal cultivation cost. Li et al. also reported that the cost of the carbon source in the algal medium ranges from 8 to 27% of the daily production cost. The amount of water consumed in the entire process is estimated to be 15,576 MT/day. With the area of 0.3 acre per pond, the total area require for the facility is approximately 7177 acres inclusive of area required for downstream process equipment.

The fixed capital investment (FCI) and the annual operating cost (AOC) for the production of 160 MT/day of protein concentrate via flash hydrolysis are estimated to be 264 million and 145 million US Dollars respectively. It is worth noting that the value of FCI excludes the cost of land acquisition. Moreover, it is evident that the FCI is contingent on the algae cultivation stage, which is driven by the algae pond construction. The major drivers here are the pond liner, paddle wheel, and pond cover while landscaping plays the minor role (Fig 3 and 3.1). The other major FCI drivers are cost of equipment installation, ultrafiltration, filtration and anaerobic digestion. This result is in agreement with the work of Rogers et al.

Operating cost, on the other hand, is controlled by facility-dependent cost, which comes from maintenance, depreciation and other miscellaneous cost. Utilities cost is due to huge energy consumption by pumping, heating and cooling, and pond agitation (Table 6 and Fig.4). The total power consumption (19.5 MW) is dictated by energy-intensive equipment summarized in Table 7. Algae growth consumed approximately 94% of the total power used in this work.

Apart from the main product (protein concentrate), power generation contributes moderately (3%) to the revenue of this project. Moreover, Low and high pressure steam, which is considered as additional credit contributes approximately 5% to the revenue. These percentages are based on the minimum product prices of $4.13 (see Table 8). The minimum value of the protein concentrate is calculated using excel solver. This is illustrated in Table A1 in the appendix.

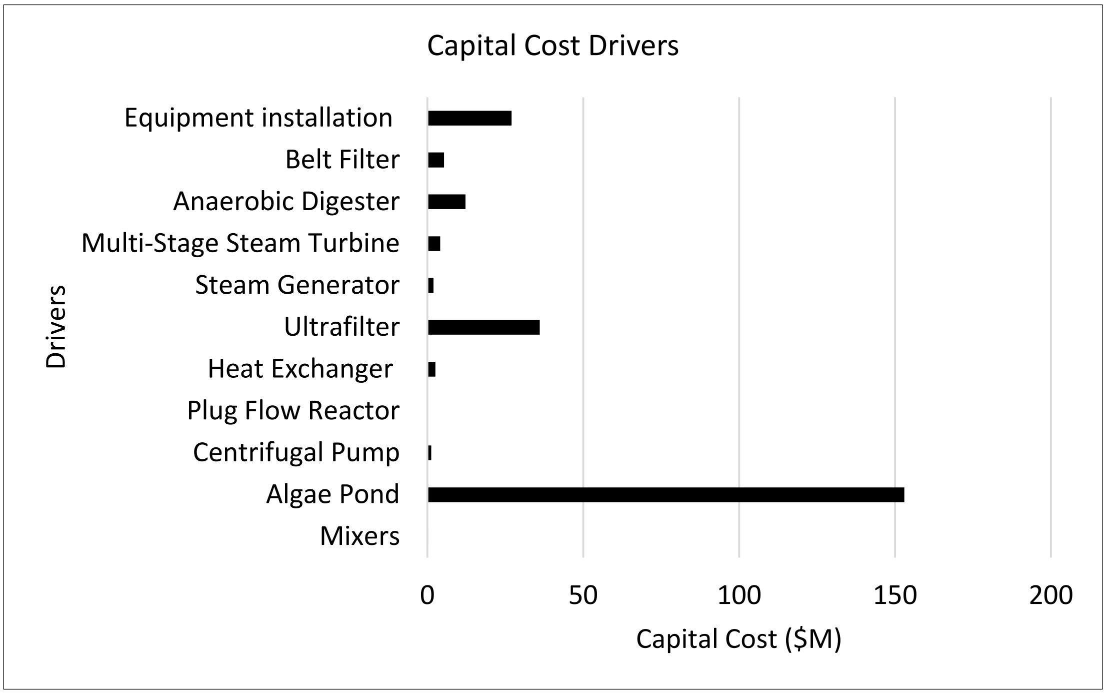


Figure 3 Capital Cost Drivers

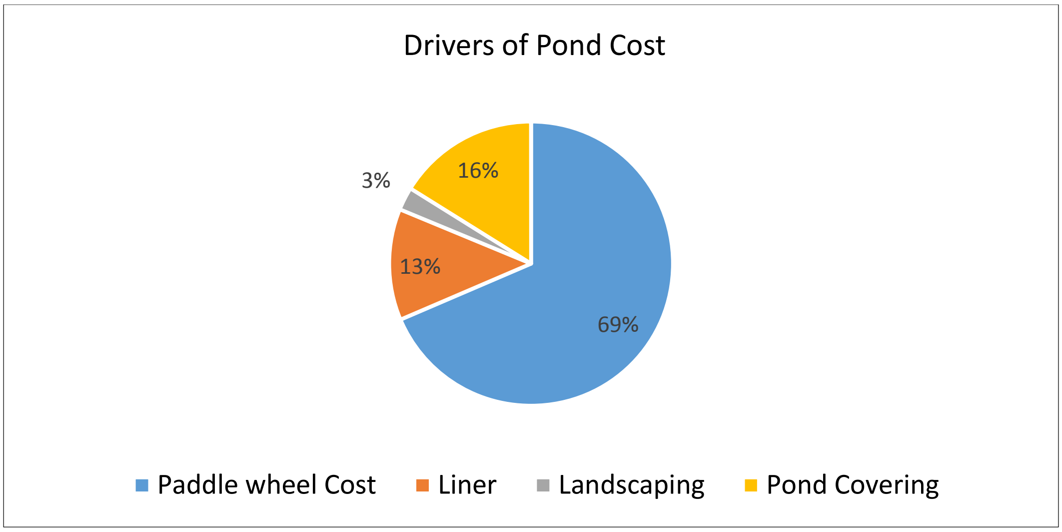


Figure 3.1 Drivers of Pond Cost

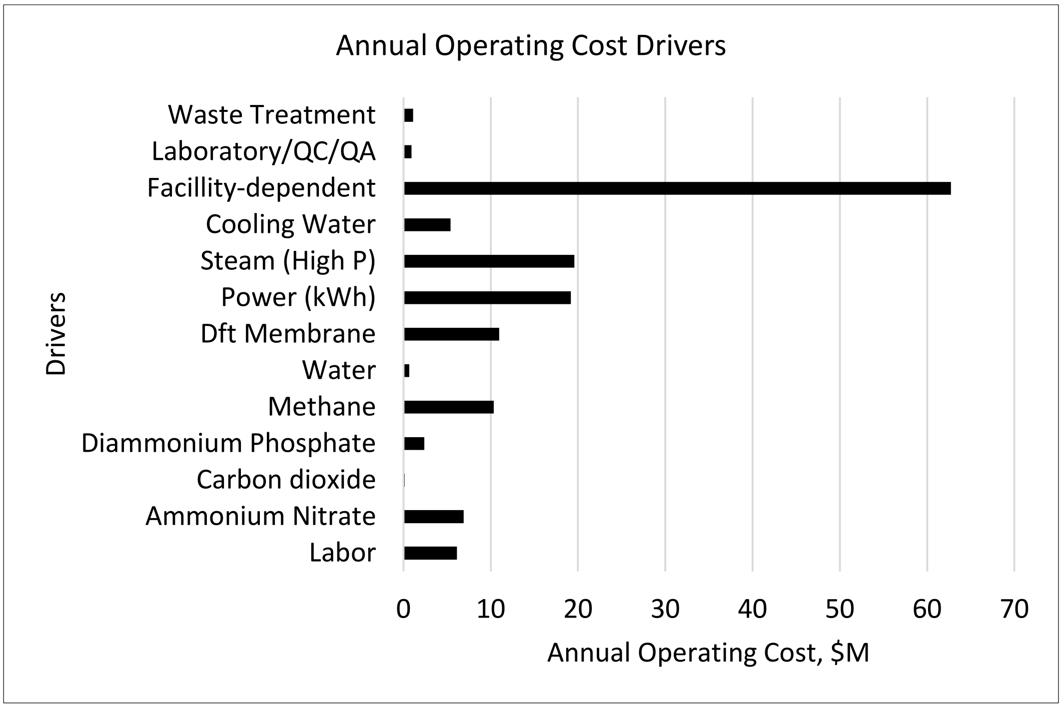


Figure 4 Annual Operating Cost Drivers

Table 7 Energy consumption at the various sections

Section	Power (kwh/h)
Algae Growth	16765
Biogas Production	166
Ultrafiltration	652
Pumping	1441
Drying	359

Table 8 Revenue/Credit Summary for 160 MT/day of Protein Concentrate

Description	Rate	Rate Unit	Price	Price Unit	Revenue ($M)
Power Generation	168412221	kwh/yr	0.08	$/kWh	13.472
Protein Concentrate	50688	MT/yr	4.13	$/kg	209.341
Total					**222.813**

5.1 Sensitivity Analysis

Results from this simulation gave the following baseline values: FCI of $264 million, AOC of $145 million, Annualized cost of $180 million, unit cost of $2.86/kg protein, and a minimum product price of $4.13/kg. These values can be compared to some of the protein prices on the market (Fig. A3). These values are controlled by the kind of microalgae employed, algae productivity, nature or kind of pond (depth, lined, open, mixing power, kind of nutrients etc.), algae slurry to the flash hydrolyzer, percent algae conversion in the hydrolysis, protein content in the algae, percent of total water recycle, project life, discount rate, tax rate, debt/equity ratio etc. This section is dedicated to analyzing how some of these factors affect the financial and technical aspect of this work.

5.2 Effect of Flash Hydrolysis Percent Conversion

The percent conversion in the flash hydrolysis was pegged at 85%, which affected the quantity of protein extracted. Changing the percent algae conversion from 85% to 95% produced a protein throughput from 6.4 to 7.1 MT/h. While the concentrated protein price from $4.13 /kg to $3.71/kg, the unit cost of protein changed from $2.86/kg to $2.56/kg (see Figures 5.1). However, changing the percent conversion did not affect the AOC and the FCI since none of the factors that affect AOC and FCI was affected by the change in percent conversion.

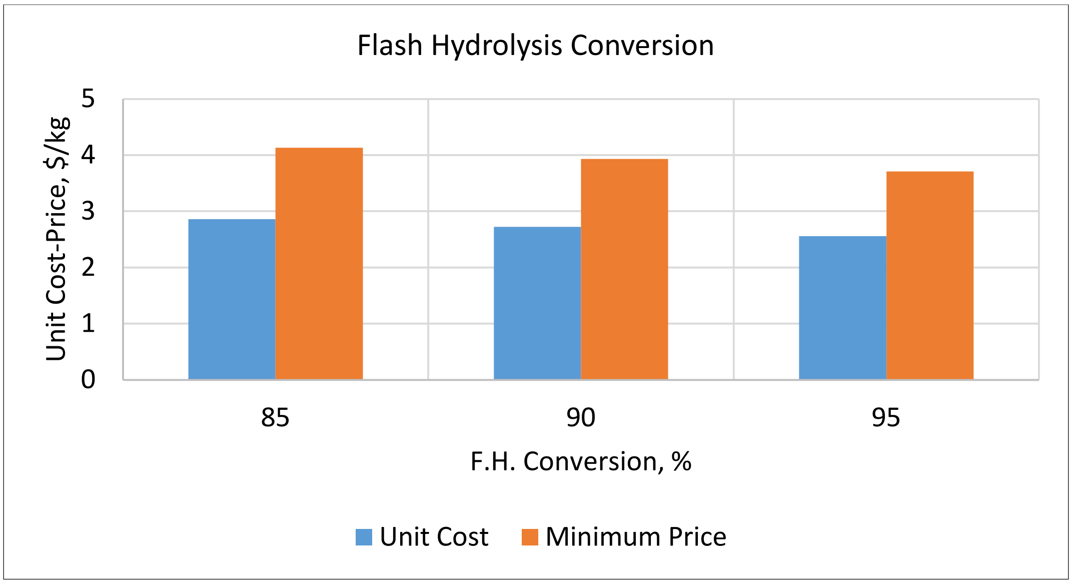


Figure.5.1 Flash Hydrolysis (F.H.) Conversion

5.3 Effect of Pond Depth

In this simulation, the baseline pond depth was is assumed to be 0.3 meters. This has tremendous effect on the amount of light used and consequent algal productivity. Changing the pond depth affected the FCI, AOC, acres of land use, protein price and unit cost. Changing the pond depth from 0.3 to 0.4 meters (Fig. 5.2) did prompt the FCI to dip greatly by 16% ($264M to $223M). This decrease stems from the fact that the total number of ponds and the land required decreases. Due to the decrease in the number of pond, the AOC also decreased from $146 M to $131 M (Fig.5.2). Decrease in the annual operating cost prompted the unit cost, which is the annual operating cost divided by the total annual protein, decrease from $2.86/kg to $2.58/kg (Fig. 3). Since the minimum protein price is contingent on both the FCI and AOC, their decrease consequently reduced the protein price from $4.13/kg to $3.61/kg (Fig. 5.3).

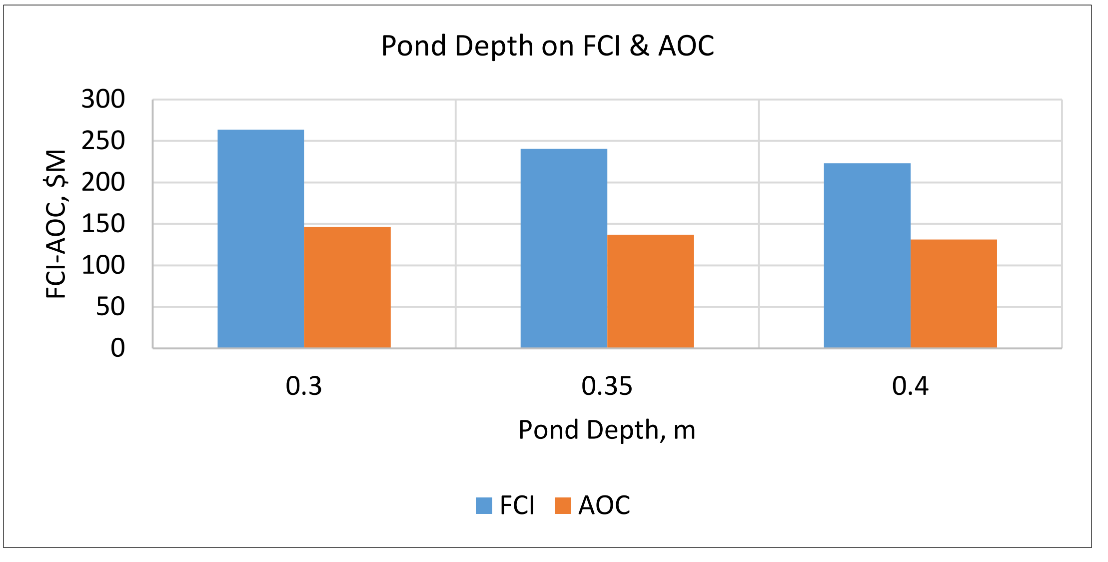


Figure 5.2 Pond Depth on FCI & AOC

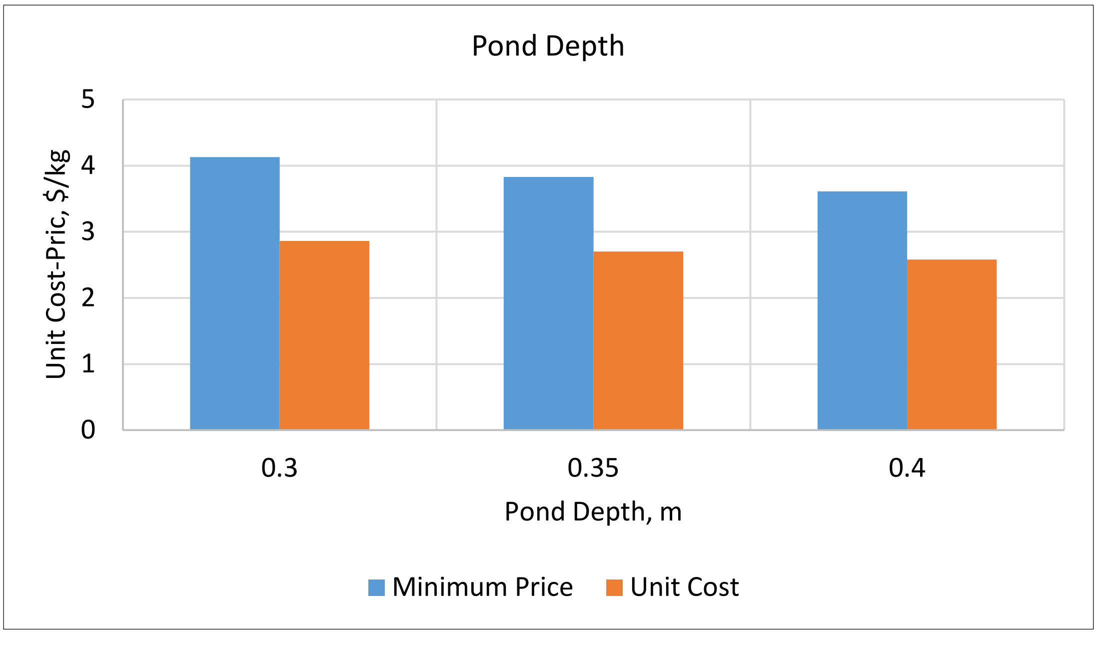


Figure 5.3 Pond Depth

5.4 Effect of Protein Content in the Algae

The percent protein content in the microalgae dictates the amount of annual protein produced. The baseline protein content in the microalgae is assumed to be 54%. Changing the protein from 54% to 70% increases the protein from 6.4 MT/h to 8.34 MT/h, which in turn decreases the protein price from $4.13/kg to $3.18/kg (Fig 5.4). Increase in the protein content decreases the total amount of solids generated during flash hydrolysis. This consequently reduces the number of filters required for filtration prior to ultrafiltration. Moreover, the number of ultrafilters also decreases due high protein concentration gradient across the membrane. These reductions in equipment slightly decrease the both FCI (from $264 M to $262 M) and AOC (from $146 M to $145 million) as the protein content in the microalgae increases. The unit cost eventually reduces from $2.86/kg to $2.20/kg (Fig 5.4). Moreover, the annualized cost decreased from $180M to $179M.

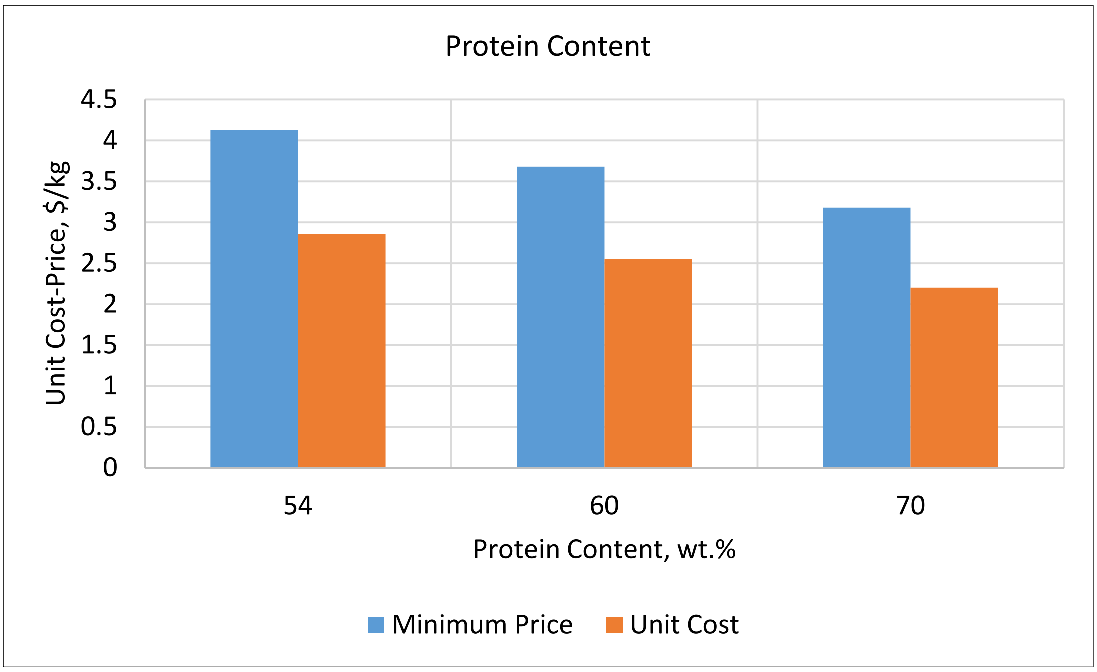


Figure 5.4 Protein Content

5.5 Effect of Algae Productivity

Davis et al. employed algae productivity 25 g/m^2/day as the baseline in their simulation of techno-economic analysis of autotrophic microalgae for fuel production. Rogers et al. also use a value of 15 g/m^2/day in simulating a critical analysis of paddlewheel-driven raceway ponds for algal biofuel production at commercial scales. In this work the baseline algae productivity was assumed to 15 g/m^2/day. This value affects the concentration of the biomass a given pond for a given area. Varying the algae productivity from 10 g/m^2/day to 25 g/m^2/day increases the algae biomass from 6.4 MT/h to 38.7 MT/h. This increase moved the minimum protein price from \$4.13/kg to \$1.50/kg while the unit cost of protein changed from \$2.86/kg to \$1.04/kg (Fig. 5.5). The increase in algae productivity does not affect FCI and AOC since this factor describes the microbial growth rate and not the need for additional area. Microbial growth rate can be improved by employing genetically modified culture. This also depends on the location of the pond where radiant energy is present 12 hours/day.

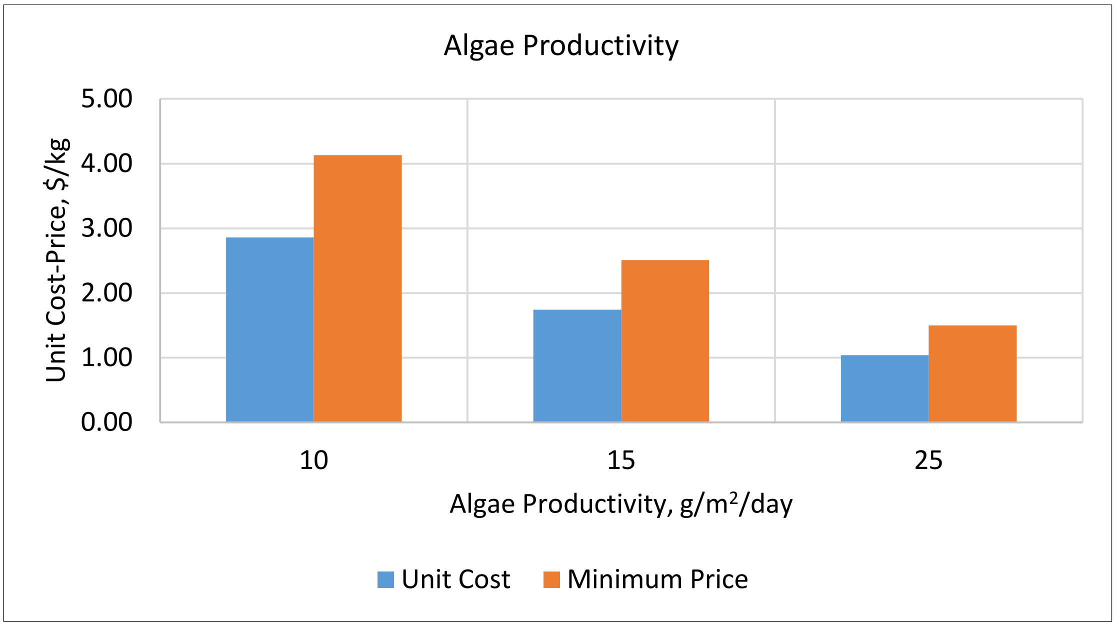


Figure 5.5 Algae Productivity

5.6 Effect of Water Recycle

Water recycle in the algae industry is highly indispensable practice. In this work, without water recycle, the annual amount of water required would be 127 million gallons per day (MGD) ($2.1 million per year), which is approximately 1.4% of the AOC. However, with incorporation of recycling strategy, the annual water utilized was approximately 4.2 MGD ($0.07 million per year), which represents 97% reduction in the cost of annual water use. In this work, it is assumed that 98 % of the water is recycled. Testing the sensitivity of percent water recycle on the AOC did not show any significant change. This is buttressed by the fact that the annual cost of water contribution to the AOC is merely 1%. Davis et al. reported that varying the percent water recycle from 80-100% did not change the unit cost and minimum price of the algae. Fig. 5.6 depicts the change in water use as the percent recycle changes.

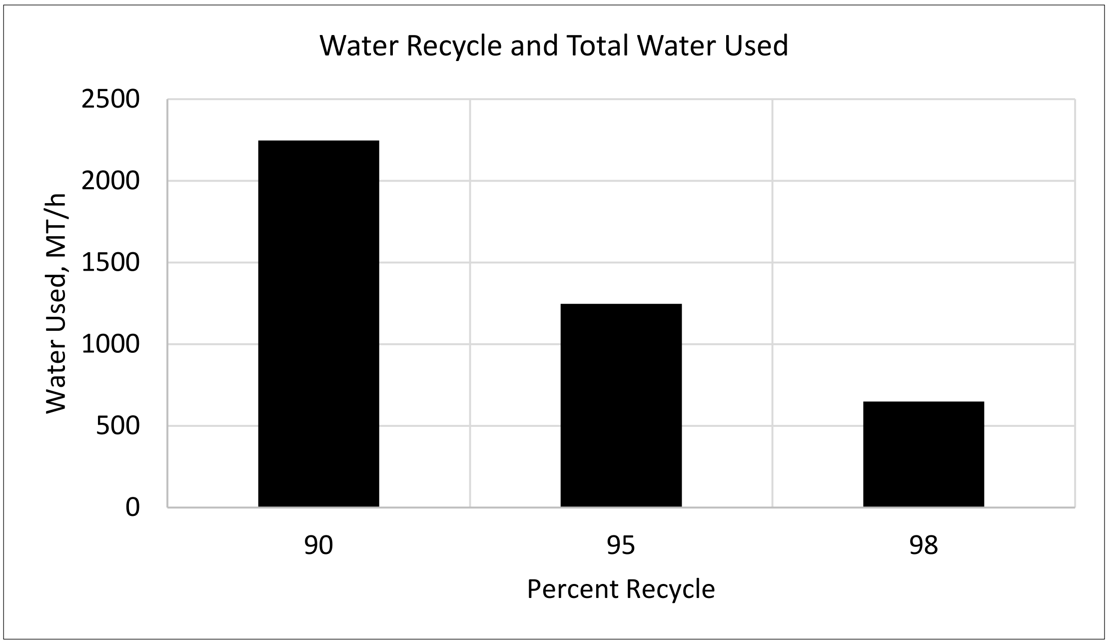


Figure 5.6 Water Recycle and Total Water Used

5.7 Effect of Algae Slurry

In flash hydrolysis, the more the water content of the slurry the more energy required to pump, and the more the annual operation cost for a specific production capacity. In this work, the simulation was done using 20 wt.% of algae slurry. Dote et al and Minowa et al. published the first reports of hydrothermal liquefaction of microalgae using a batch reactor with high feed concentration dry matter algae mass, 50 wt.% and 78.4 wt.%, respectively. However, pumping these slurries through a continuous reactor is highly impractical due to flowability issues.

In this study, decreasing the percent weight of the algae slurry increases the FCI and AOC due to the increase in the total volume of slurry with the amount of dry weight of algae being constant (14 MT/h i.e. the baseline). Moreover, the total mass of water increases from 649 MT/h to 866 MT/h. Furthermore, the number of filters, ultrafilter, anaerobic digesters, pumps, etc. increases when the percent algae in the slurry decreases from 20 to 5%. Increase in the number of equipment increases the amount of total power consumed from 19.5 MW to 22 MW. Furthermore, the total water consumption increases by 25% (i.e. 649 MT/h to 866 MT/h). As a result, the FCI increased from $264 million to $481 million while AOC increased from $146 million to $301 million (Fig.5.7). Consequently, the unit cost and minimum prices increased from $2.86-$5.94 and $4.13 to $8.34 respectively (Fig. 5.8).

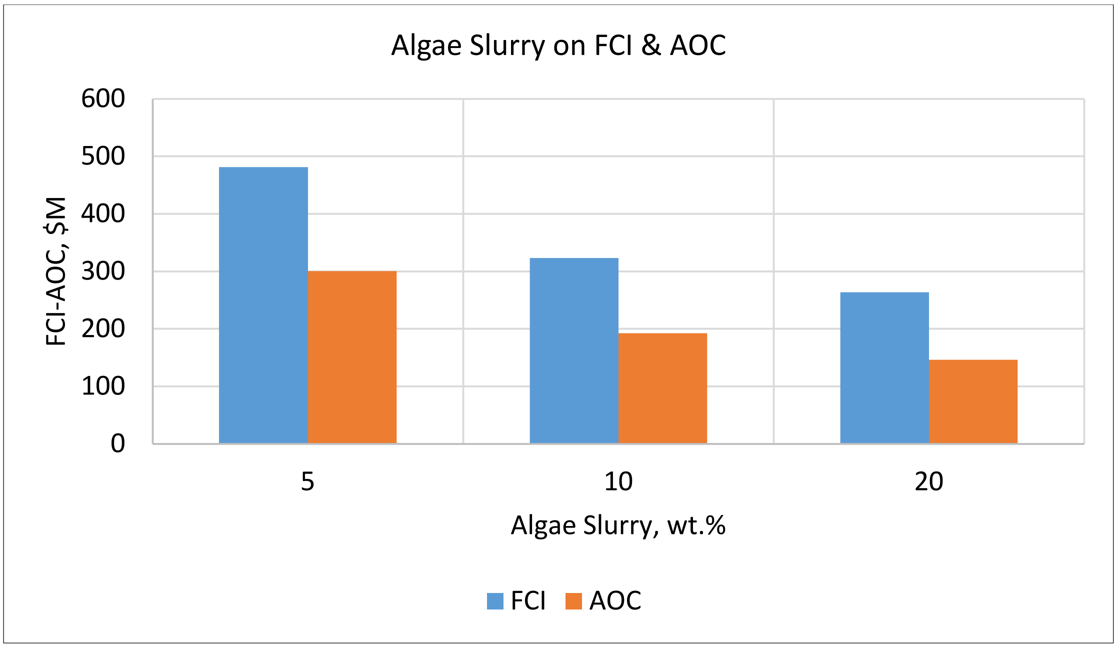


Figure 5.7 Algae Slurry on FCI &AOC

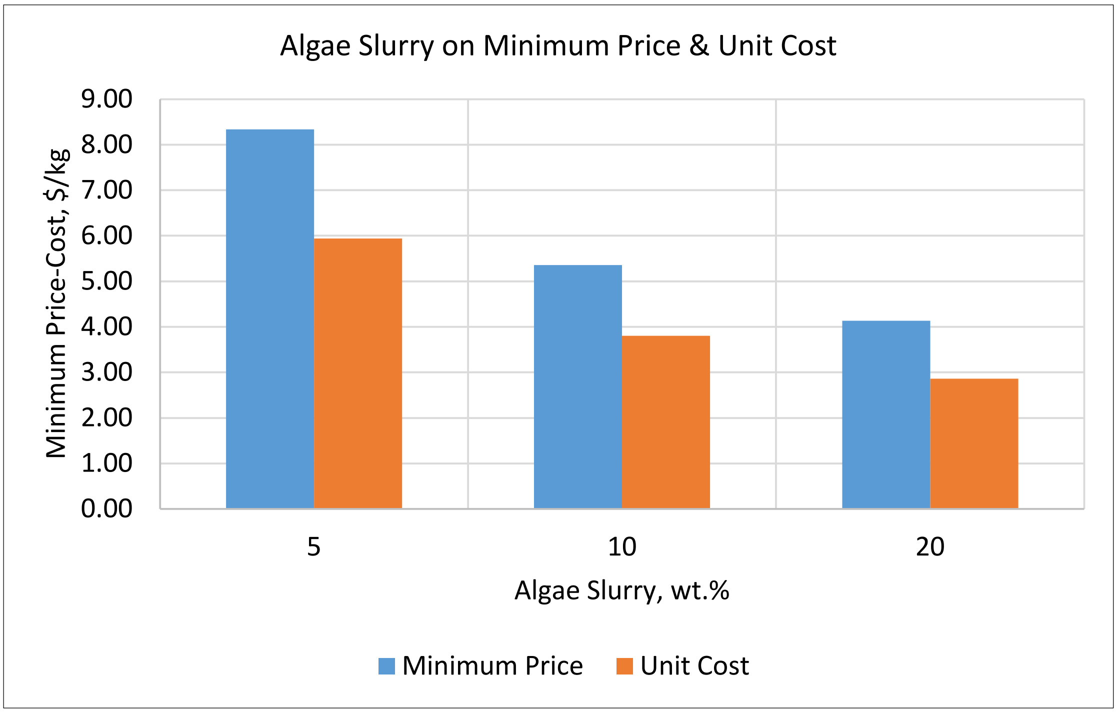


Figure 5.8 Algae Slurry on Minimum Price-Unit Cost

5.8 Effect of Project life

The life of a project is crucial as far as recouping of the FCI is concerned. Moreover, it is well-advised to reduce the project life in profitability analysis so as to offset any risk of inflation. Increasing or decreasing the project life affects the minimum product price and the annualized capital investment. The baseline project life in this work is assumed to be 15 years. Changing the project life from 10 to 20 years decreases the minimum product price from \$4.75 to \$3.90 while the annualized cost decreases from \$188 M to \$176 M (Fig.5. 9 and 5.10).

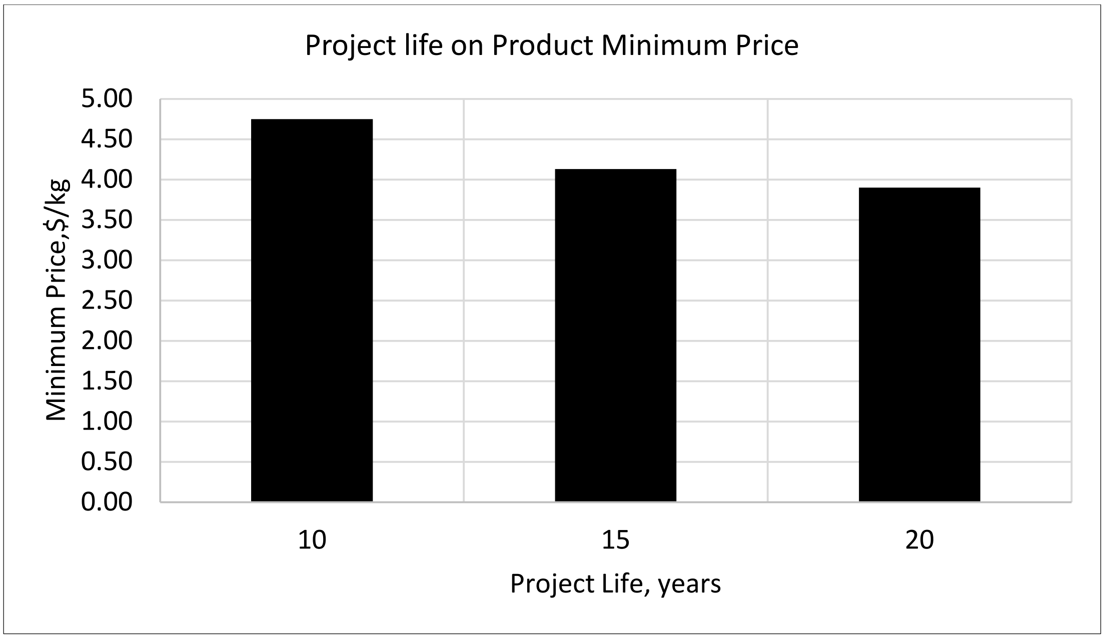


Figure 5.9 Project life Product Minimum Price

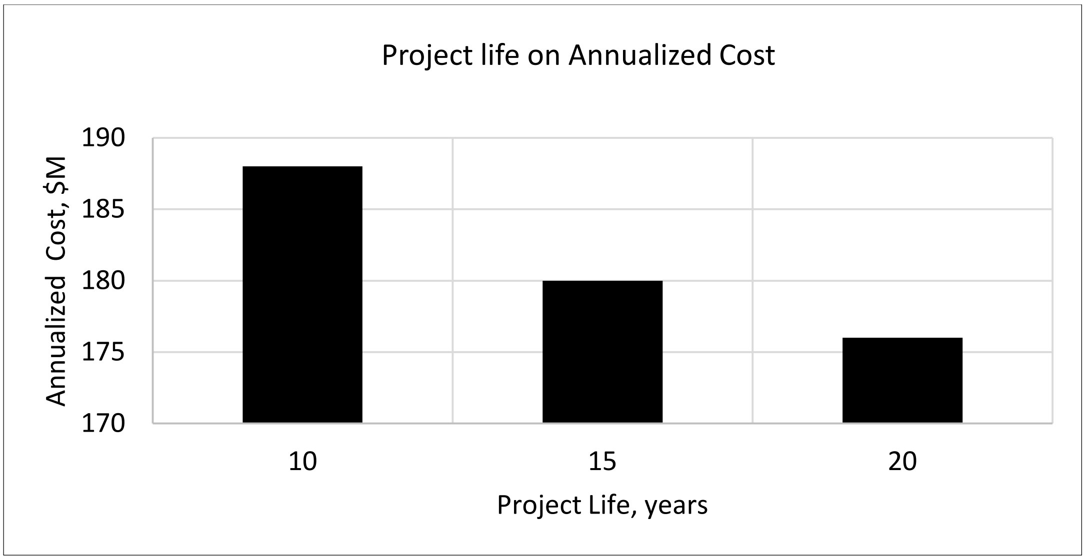


Figure 5.10 Project Life on Annualized Cost

5.9 Effect of Discount Rate

Another important parameter of interest in this work is the discount rate of the interest rate. The baseline interest rate in this work is assumed to be 10%. Varying the interest rate from 7-12% increases the minimum product price from \$3.87 to \$4.33 while the annualized capital investment jumps from \$174 M to \$183 M (fig. 5.11 and 5.12).

Figure 5.11 Discount rate on minimum Product price

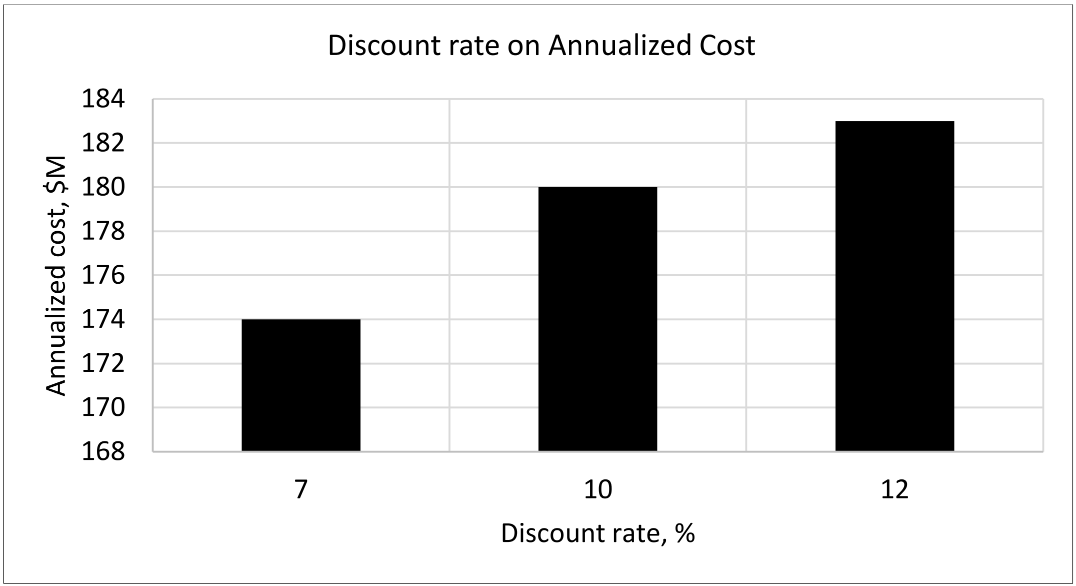


Figure 5.12 Discount rate on Annualized Cost

5.10 Effect of Product Price

The viability of protein concentrate production from microalgae by flash hydrolysis was assessed by looking at the net present value (NPV), internal rate of return (IRR) and the payback period (PBP) based on the product prices. It is obvious that product prices from $4.13/kg-$6.00/kg reduce the PBP from 10-3 years at an interest rate of 10% (Fig. 5.13). This is highly attractive provided there is a limited risk in the process technology, supply and cost of raw material and market for protein concentrate. Besides, many companies prefer PBP of 3 to 5 years [El-Halwagi, 2012]. The protein price in this work can be compare to that obtained by Navarro da Silva et al. who reported R$23.70/kg (6.25US/kg) of 80 wt.% whey protein concentrate.

In order to assess the viability of the project, the sensitivity of the NPV was tested via the minimum prices. It could therefore be inferred from Fig.5.14 that NPV becomes more positive with product prices climbing from $4.13 to $6.00 (see Table 7.1).

Though IRR is a subjective financial parameter, its knowledge will assist stakeholders in the algae industry to make easy and sound decision. It is truism that higher product price gives higher IRR. The higher the IRR the more attractive the enterprise.

Fig. 5.15 highlights the effect of internal rate of return as product prices increase. These evaluations are based on 10% interest rate, 40% tax rate and 100% debt financing.

Table 9 Effect on product price on NPV, PBP, and IRR

Protein Price $/kg	NPV Million Dollars	PBP Years	IRR %
4.13	0.0	9.8	25
4.50	57	7.1	30
5.00	134	5.1	36
5.50	212	4.1	42
6.00	290	3.3	48

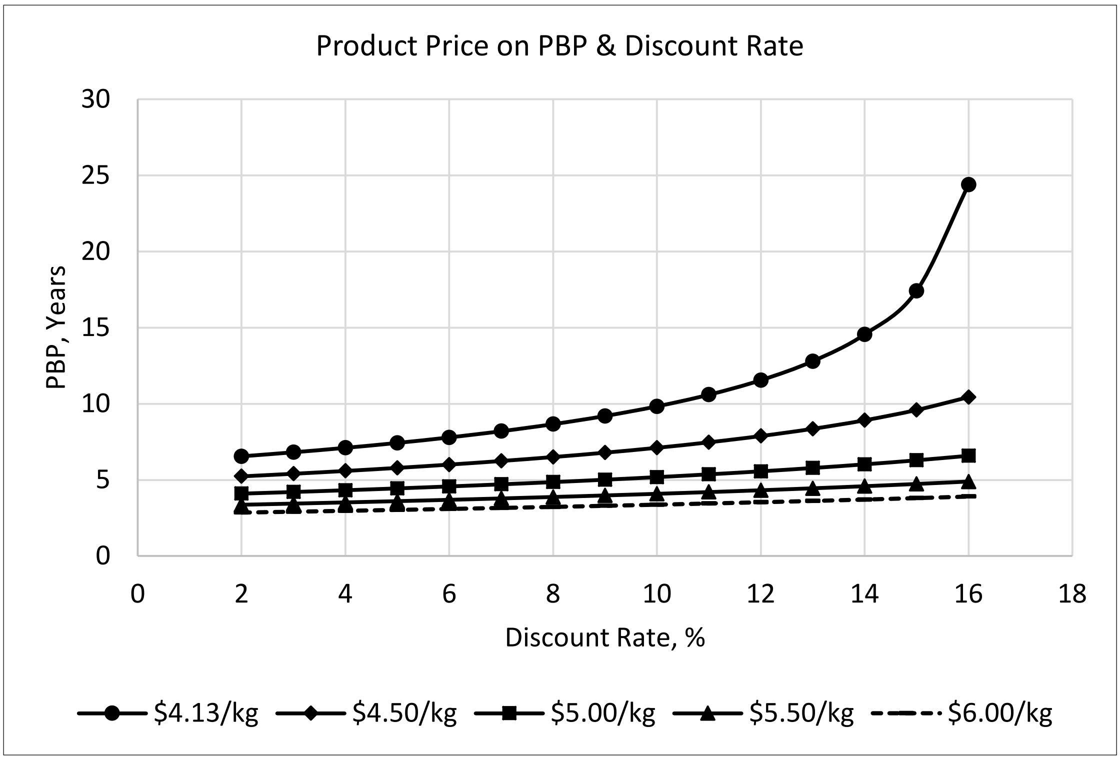


Figure 5.13 Product Price on PBP & Discount Rate

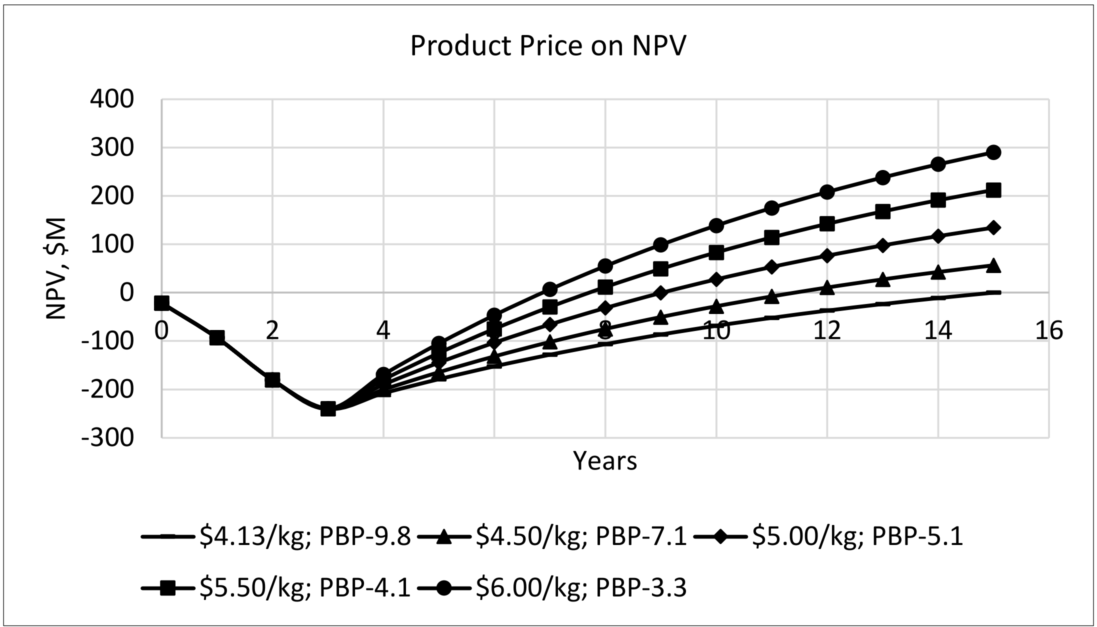


Figure 7.14 Product Price on NPV

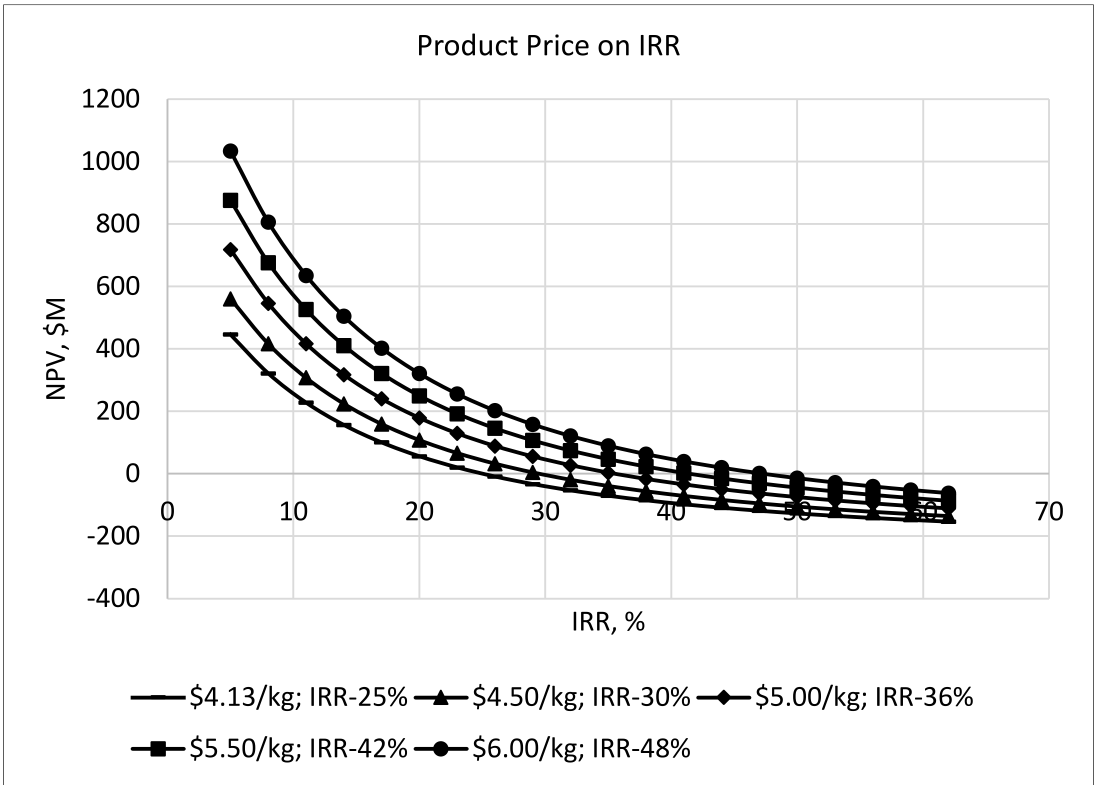


Figure 5.15 Product Price on IRR

5.11 Major Drivers of Unit Cost and Price

Figures 5.16 and 5.17 depict the summary of sensitivity of unit cost and price of protein concentrate to the aforementioned factors in the work. The only factors not discussed in the preceding section are the effect of pond covering, pond liner, paddle wheel energy consumption, and CO_2 and flue gas.

It is assumed that high density polyethylene (HDPE) can be used a pond cover to reduce evaporation of water from the raceway pond. However, foregoing pond covering reduces the FCI, AOC, and annualized cost (AC) to $233 million, $138 million and $168 million respectively, which resulted in unit cost and unit price reduction of ¢15 and ¢30 respectively.

To enhance photosynthetic efficiency of a raceway pond, there is the need to incorporate paddlewheel to agitate the algae culture, and expose them to the necessary radiant energy. Lundquist et al. reported that absence of light for the algae culture over the night can result in biomass loss of 25%. Mixing and agitation the pond demand huge amount of energy. In this work, 0.73 W/m^2 was used as published by Rogers et al. However, reducing this value to 0.22 W/m^2 reduced the FCI and AOC to $ 262.5 million and $134 million respectively, which consequently reduced the unit cost and unit price equally by ¢22.

Pond liners play crucial role in raceway pond by preventing pond contamination or leakage. These liners could be clay, concrete, HDPE, etc. However, due to cracks and seismic activities, HDPE is preferred to clay and concrete [Roger et al., 2013]. In this work the cost of lining all the ponds (19138) amounts to $5.3 million (2% of FCI). Without lining the pond, the FCI dropped to $243.4 million thereby reducing the unit price and cost by ¢20 and ¢10 respectively.

Supplying CO_2 to the microalgae via the flue gas has the advantage reducing global warming from power plants. However, in this work it is evident that using pure CO_2 is more economically friendly than employing flue gas. With the current CO_2 price of $40/ton, the FCI and AOC of this work reduced to $253 million and $139 million respectively. It could be seen that the unit cost and prices decreased by ¢12 and ¢16 respectively. However, forgoing flue as a source of CO_2 supply also offsets in situ major power supply to the plant. Nevertheless, minor power (3.2 MW) supply can be produced from the approximately 2 MT/h methane from the anaerobic reactor. This minor power supply represents 16% of the required total power for the whole process. This means the rest of the power should be purchased from outside the plant.

Furthermore, it is evident that algae productivity and slurry to the FH have the most remarkable effect on both unit cost and prices of the protein concentrate production.

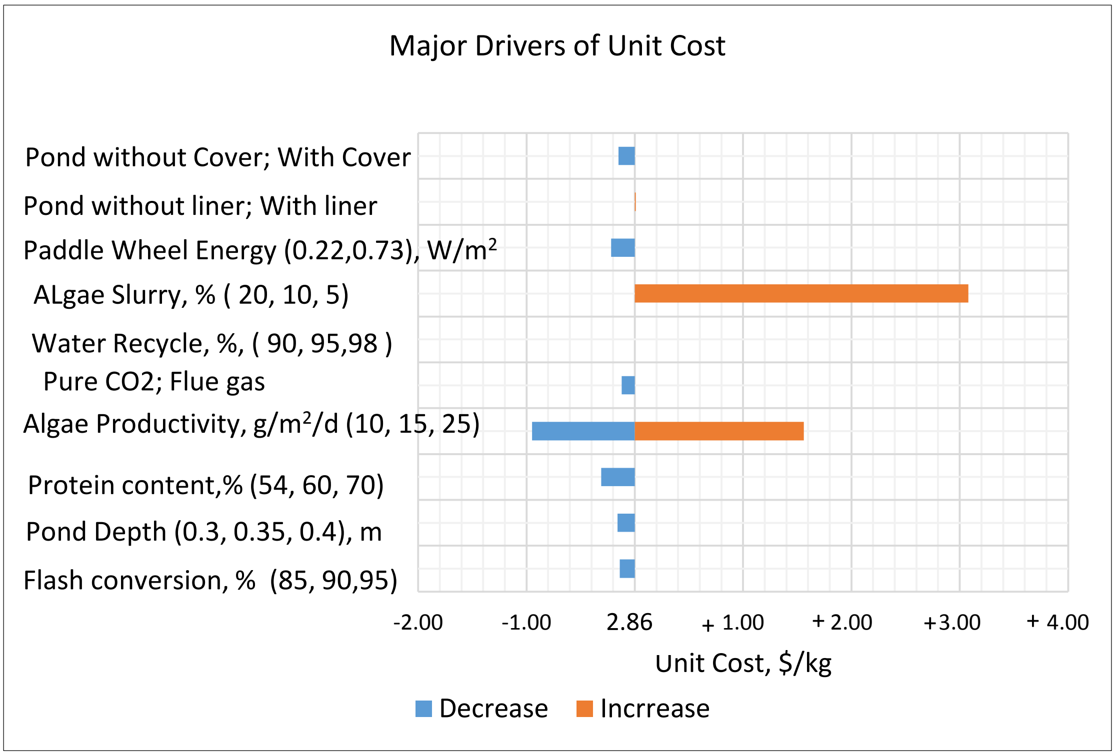


Figure 5.16 Major Drivers of Unit Cost

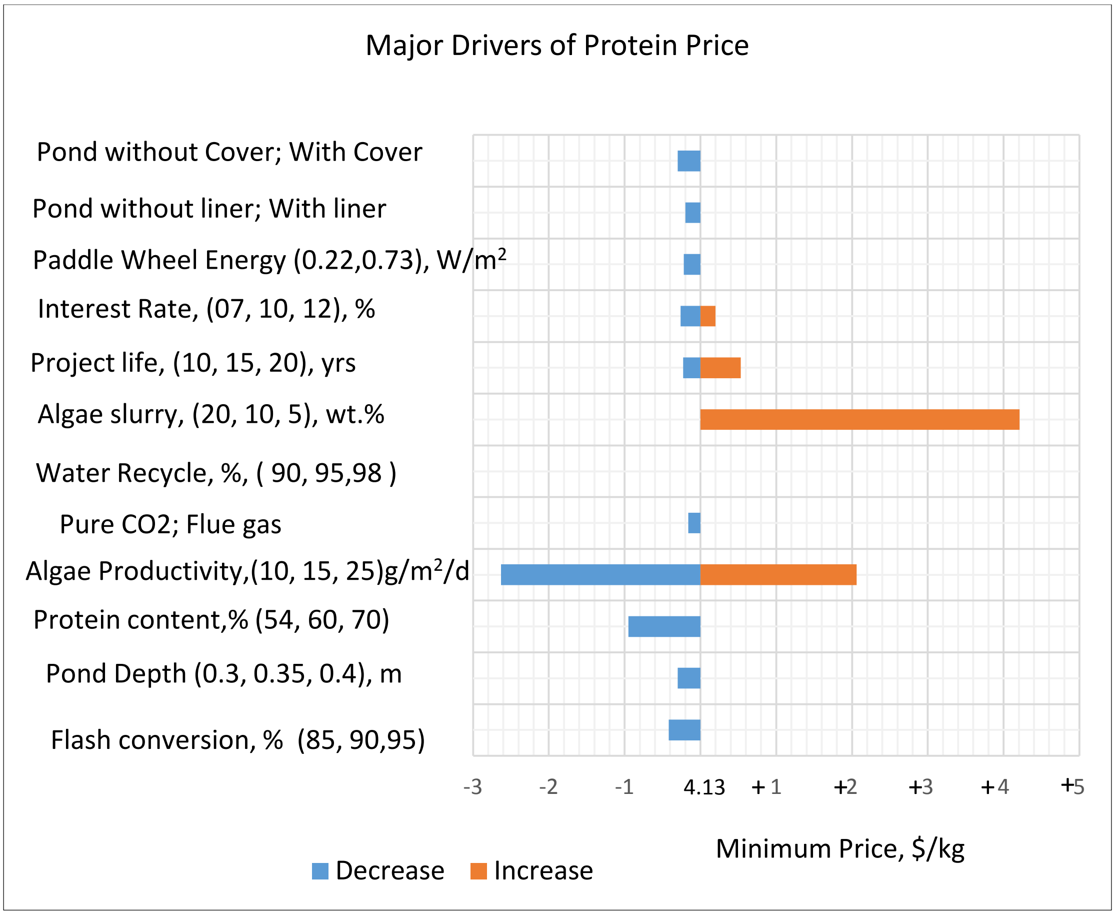


Figure 5.17 Major Drivers of the Protein Price

5.12 Effect of Plant Capacity

Most industrial enterprises capitalize on the economy of scale i.e. the bigger the better. Conversely, it does not work well in the present work. Under normal circumstances, increasing the number of barrels of protein concentrate produced decreases the selling price. However, the change in prices of protein is not remarkable above capacity of 160 MT/day as compared to below 160 MT/day (Fig. 5.18). Moreover, production capacities above 160 MT/day, require huge capital input that could scare investors (Fig. 5.19). Table A2 in the appendix delineates results of capacity impact on the salient economic inputs and output from this work.

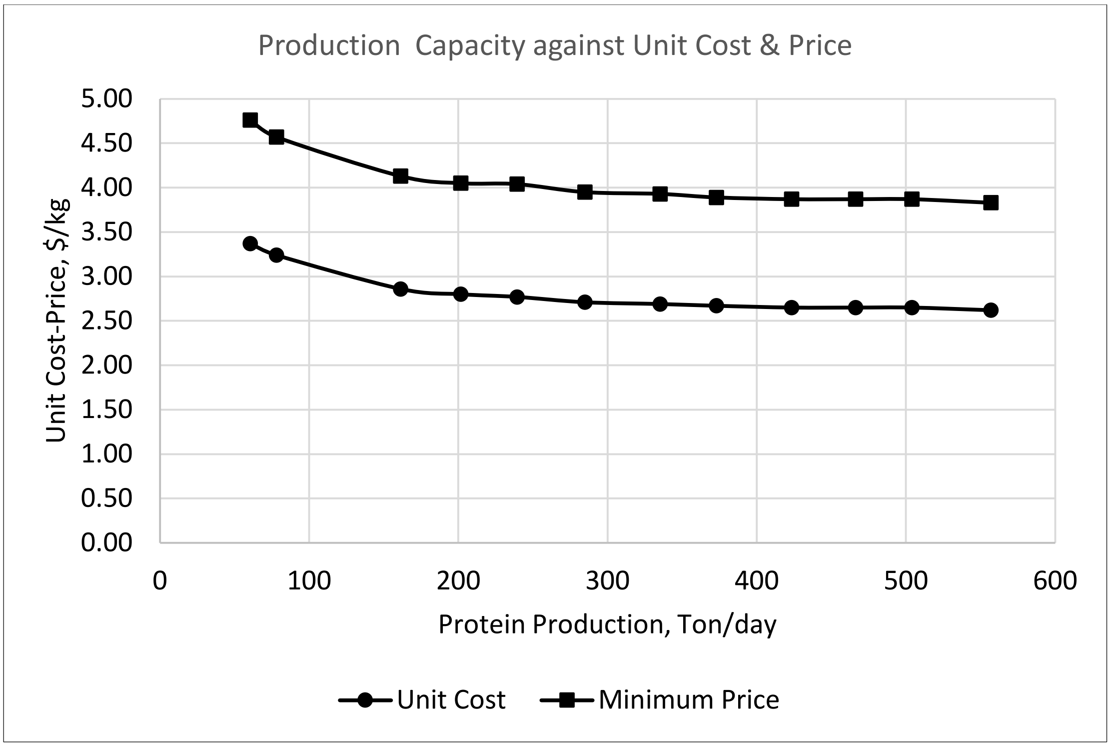


Figure 5.18 Production Capacity against Unit Cost &Price

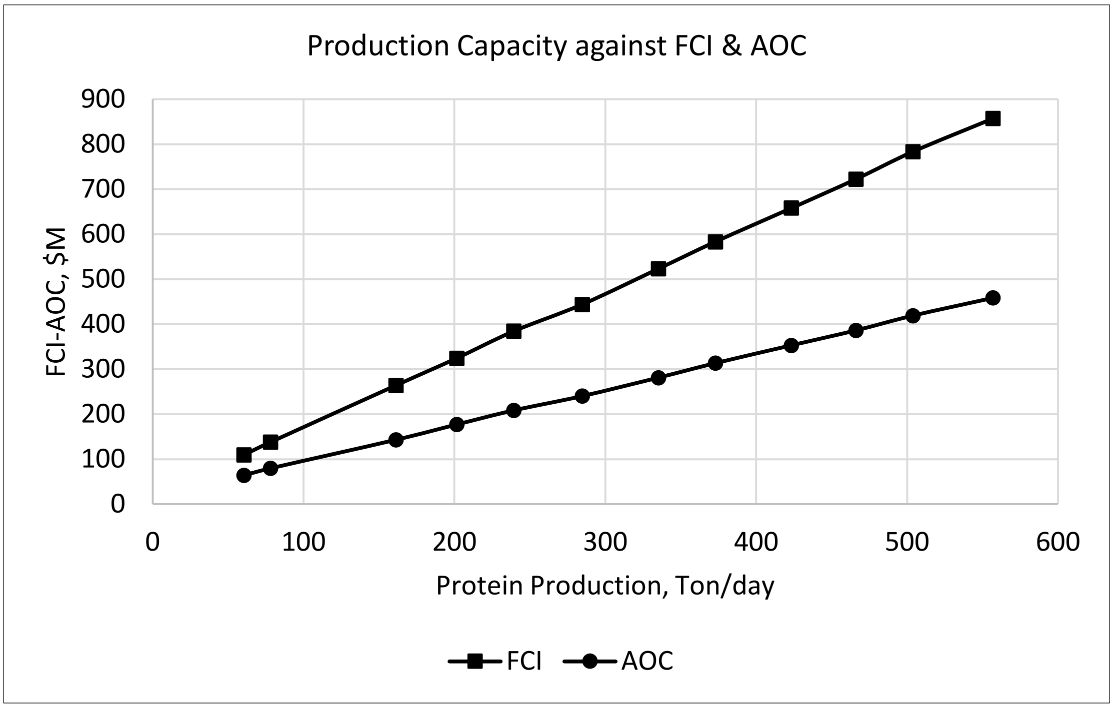


Figure 5.19 Production Capacity against FCI & AOC

CONCLUSIONS

This work has focused on the technical and economic assessment of 95 wt.% protein concentrate (food and pharmaceutical grade) production from microalgae. Data used is based on the previous work done at the Old Dominion Algae Laboratory. Additional data concerning protein production from microalgae has been generated to assist stakeholders in the algae industry to make meaningful technical and economic decisions. A baseline capacity of 160 MT/day of protein concentrate was employed in developing this model.

However, there are hurdles in the scalability of this model to a full-scale operation looking at the huge FCI and AOC. One of the obstacles in this work is the huge amount of freshwater required to grow the algae. To embark upon this enterprise, there is the need to locate the plant near places with abundance of water. The use of brackish or saline water, which is more abundant, has been successful in microalgae cultivation [Lee, 2001]. Wastewater has also become one the promising candidates that can be employed to grow microalgae in this kind of project. Using non-freshwater will eventually reduce the cost of water treatment and cost nutrient employed.

Not only the above-mentioned impediments are inherent in this work, but also the cost of mixing or agitating the pond for effective algal growth is real. To reduce this cost, there is the need to slope raceway ponds so the entire algae culture can flow in a fashion that will enhance utilization of radiant energy. This kind of pond design has been reported by Craags et al.1997.

Furthermore, algae productivity has most pronounced effect on the protein price, there is the need to galvanize the research and development of different strain of algae whose growth rates for protein concentrate production.

Apart from using microalgae to produce protein concentrate, it is evident that the biofuel intermediate has the potential of supplying energy through anaerobic digestion, which supplies almost 20% of the total power requirement for the whole process.

Moreover, the percent weight of protein in microalgae play important role in economic the feasibility in this present work as it controls the yield and consequently affects the product price of protein concentrate. To enhance attractiveness of this work, more research and development should be geared towards the cultivation of high proteinaceous microalgae.

Though ultrafiltration, a membrane separation process, contributes almost 40% of the total energy demand in this work, it is more promising in producing protein concentrate.

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APPENDIX

A1. Calculation of Direct Cost (DC)

Piping (A) = 0.35 x PC

Instrumentation (B) = 0.4 x PC

Insulation (C) = 0.03 x PC

Electricals Facilities (D) = 0.1 x PC

Buildings (E) = 0.45 x PC

Yard Improvement (F) = 0.15 x PC

Auxiliary (G) = 0.4 x PC

Installation = Installation of listed equipment + Installation of Unlisted (overlooked) equipment.

Unlisted equipment installation cost = 0.5 x unlisted equipment cost

$\Longrightarrow$DC = PC + Installation + A + B + C + D + E + F + G

A2. Indirect Cost (IC) is calculated as follows:

Engineering (H) = 0.25 x DC

Construction (I) = 0.35 x DC

A3. Other Cost (OC) is calculated as follows:

Contractor’s fee = 0.05 (DC + IC)

Contingency = 0.1 (DC + IC)

A4. Financial Calculations

Inputs

Tax rate	40%
Interest rate	10%
Project life	15yr

Cash flow

Land in zero years	-21.5
30% of the FCI in 1^{st} yr	-79.05
40% of the FCI in 2nd yr	-105.4
30% of the FCI In the 3rd yr	-79.05
Annual COP, $M	145

Revenue

Price of protein, $/ton	Unknown (y)
Total protein produced, ton/yr	50688
Energy, $M	13.31
Annual Revenue	$=50688y+13.31x10^{6} - 145*10^{6}$
Annual Income	$= (50688y-131.69*10^{6})*(1-\text{Tax rate})$
	$=30412.8y -79.014x10^{6}$
	$= (0.030413y-79.014)*10^{6}$

Table A1 Calculation of minimum protein concentrate

End of year n	Annual (nondiscounted Cash Flow)/$ Milllion	Discount factor $\frac{1}{(1+i)^n}$	Discounted cash flow in $ MM	Cumulated discounted Cash flow ($MM)
0	-21.5	1.000	-21.5	-21.5
1	-79.05	0.909	-71.8644	-93.4
2	-105.4	0.826	-87.1026	-180.5
3	-79.05	0.751	-59.3903	-239.9
4	(0.0304y-79.0)	0.683	0.0208y-53.96	0.0208y-293.81
5	(0.0304y-79.0)	0.621	0.0189y-49.05	0.0396y-342.87
6	(0.0304y-79.0)	0.564	0.0172y-44.60	0.0568y-387.46
7	(0.0304y-79.0)	0.513	0.0156y-40.54	0.0724y-428.00
8	(0.0304y-79.0)	0.467	0.0142y-36.85	0.0866y-464.86
9	(0.0304y-79.0)	0.424	0.0129y-33.50	0.0995y-498.36
10	(0.0304y-79.0)	0.386	0.0117y-30.45	0.1112y-528.82
11	(0.0304y-79.0)	0.350	0.0107y-27.69	0.1218y-556.50
12	(0.0304y-79.0)	0.319	0.0097y-25.17	0.1315y-581.67
13	(0.0304y-79.0)	0.290	0.0088y-22.89	0.1403y-604.56
14	(0.0304y-79.0)	0.263	0.0080y-20.80	0.1483y-625.36
15	(0.0304y-79.0)	0.239	0.0073y-18.91	0.1556y-644.27
		Sum =	**(0.1556y-644.27)**	

The value of y can be solved by equating the sum of discounted cash flow, NPV, to zero gives y to be $4.14/kg. i.e. the minimum price of protein concentrate.

Table A2 Simulation based on production of different capacities

Algae MT/day	Protein MT/day	FCI $M	AOC, $M	AC $M	Power (gen) MW	Power(used) MW	Water MT/h	Unit Cost $/kg	Price, $/kg	CO_2 Captured MT/h	Land acres	Natural Gas MT/h
105	60	110	64	78	9	7	250	3.37	4.76	10	2692	5
168	78	138	80	98	13	10	344	3.24	4.57	13	3544	7
336	161	264	146	180	21	19	649	2.86	4.13	27	7177	12
420	202	325	177	220	26	24	802	2.8	4.05	33	8971	14
504	239	385	209	259	30	29	952	2.77	4.04	40	10765	17
588	285	444	240	299	35	34	1105	2.71	3.95	46	12559	19
694	335	523	281	350	41	40	1294	2.69	3.93	55	14802	22
778	373	583	313	390	46	45	1454	2.67	3.89	61	16596	25
883	423	658	353	440	52	51	1642	2.65	3.87	70	18839	28
970	466	723	386	481	57	55	1800	2.65	3.87	76	20633	31
1051	504	784	419	522	63	60	1960	2.65	3.87	83	22427	34
1157	557	857	458	571	69	66	2148	2.62	3.83	91	24669	37

Table A3. Prices of Protein Concentrates on the market

Protein Type	Quantity	Price, $	Price $/kg	% protein	Market
Whey Protein Isolate powder	5 lb	57	25.11	90	Amazon
Soy Protein Isolate powder	2 lb	23	25.33	>90	Amazon
Fish Protein Powder	1kg	35	35.00	>90	Alibaba
Rice Protein Powder	1 ton	3800	3.80	>80	Alibaba
Spirulina powder (algae)	1kg	10	10.00	60	Alibaba
Sacha Inchi Powder (organic)	1kg	8.5	8.50	60	Alibaba

Source: Alibaba Group Holding Limited; Amazon.com, Inc.

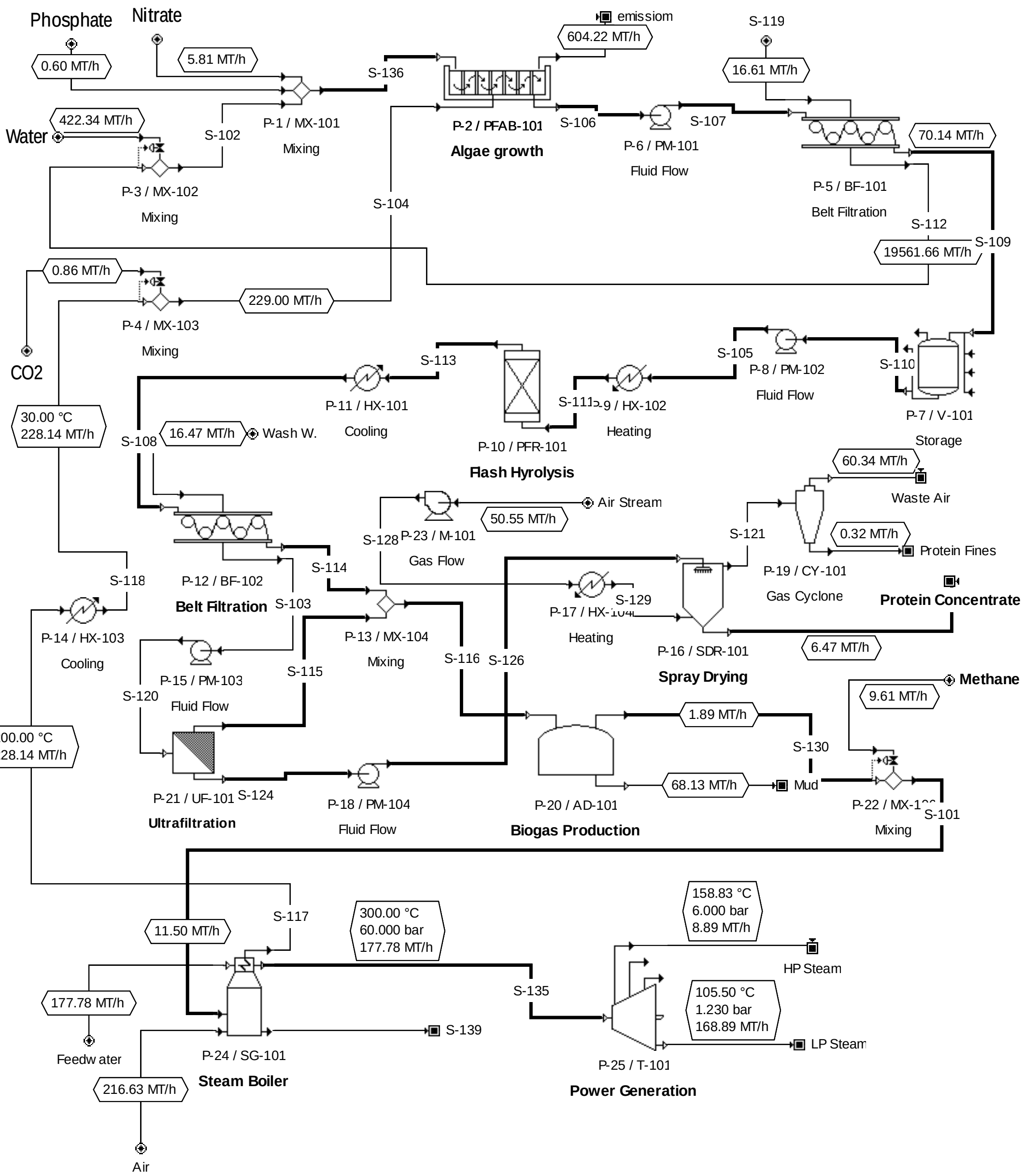

Figure A1. Production of 95 wt.% protein concentrate via flash hydrolysis modelled by SuperPro Designer v.9.

VITA

Alexander Asiedu is a graduate student from the Department of Civil and Environmental Engineering at Old Dominion University, Virginia.

He holds BSc. Metallurgical Engineering and MSc. Chemical Engineering both from Kwame Nkrumah University of Science and Technology, Ghana, in 2002 and 2008 respectively. He is currently pursuing Ph.D. Environmental Engineering at Old Dominion University.

He has eight-year professional experience as a plant metallurgist at Sian Goldfields in Ghana. His area of specialization is process design, economic analysis, and life cycle analysis.

The word processor for this thesis was Microsoft Word.