ALGAE TO BIODIESEL

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Dear Mr. Fabiano, Dr. Churchill, and Dr. Seider,

Our team was assigned the task of evaluating the technical and economic potential of a number of recently proposed methodologies for the culture and processing of algae to produce biofuel. In the spring of 2010, a CBE 459 design team focused on cultivating algae with the Simgae™ Algal Biomass Production System, extracting algal lipids using OriginOil™ single-step extraction technology and converting the lipids into n-alkanes (green diesel) via catalytic hydrotreating. They concluded that the process might be profitable, but that an excessive capital investment on the order of 2.8 billion dollars would be required. To minimize the risk to investors and improve the likelihood of garnering initial funding, it was determined that maintaining profitability while simultaneously reducing capital investment was essential. To accomplish this task, we have carefully evaluated each of the three modules that comprise the algae to biodiesel venture and have worked to improve them from both a technical and an economical standpoint.

Based on the research conducted by Miao and Wu in 2006, a "heterotrophic" algal cultivation process involving two distinct photosynthetic and fermentation stages was introduced. By drastically reducing reliance on photosynthetic growth and shifting the majority of biomass production to a fermentation phase, this novel approach significantly reduces the required area of open, sunlit fields and therefore sharply decreases capital costs for cultivation. A better understanding of the Single-Step Extraction process developed by OriginOil™ was made possible by the recent publication of several patents that detail the various lipid-extraction technologies employed in the process. This led to a more accurate economic estimation of the variable and fixed costs associated with the second module. As an alternative to the catalytic hydrotreating process used in the prior analysis, we investigated a lipid-processing module based on catalytic transesterification. In this process, algal lipids are converted to biodiesel, which consists of fatty acid methyl esters, as opposed to the n-alkanes that comprise green diesel. A plan to carry out this process was designed and simulated using the Aspen PLUS software package. The results were compared directly with those published for a similarly sized catalytic hydrotreating plant.

The results from our endeavor are quite promising. Based on the current market price of $3.30 per gallon for pure biodiesel fuel, our economic analysis suggests that an overall algae-to-biodiesel venture comprised of the three modules described above might be profitable, with net earnings of almost $340 million annually. More significantly, the capital investment required for the process is $1.2 billion, which represents an almost 60% reduction from that arrived at by last year's group. These economics are favorable and translate to a Return on Investment (ROI) of 32% and an Investor's Rate of Return (IRR) of 35%. The details of the process design, as well as a series of recommendations regarding entry into the algae-to-biofuel market are presented in the attached report.

Sincerely,

Daniel Choi

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I. ABSTRACT

In the spring of 2010, a CBE 459 design team focused on cultivating algae with the Simgae™ Algal Biomass Production System, extracting algal lipids using OriginOil™ single-step extraction technology, and converting lipids into green diesel fuel. It was determined that the process was profitable, but required a staggering capital investment of 2.8 billion dollars. In the past year, both public and private institutions have joined the race to produce biofuels from an algal feedstock in an economically responsible manner that is by maintaining profitability while minimizing high capital costs. The intention of this report is to contribute to the global discourse on alternative-fuels and to reevaluate the promise of algae as a renewable resource for alternative fuels in light of the latest research and technological advances.

The algae-to-biofuel venture was segmented into three modules: algal cultivation, lipid extraction and lipid processing. Each module was studied thoroughly and several strategies were proposed for the reduction of its associated fixed, capital and variable costs. As contrasted with a previous study, it was concluded that heterotrophic algal cultivation and transesterification lipid-processing technologies would improve the efficiency and reduce the total capital investment. Once each module was designed in detail, the three segments were stitched together to perform an overall economic analysis. Based on the current market price of $3.30 per gallon for pure biodiesel, a project life of 15 years, and a 15% discount rate, the results indicate that an algae-to-biodiesel process may not only be profitable, but also a sound and reasonable investment. The project's projected Net Present Value (NPV) is $1.3 billion and the Return on Investment (ROI) was determined to be 32%.

Although these economic results are promising, they are based on an analysis that necessarily invoked highly uncertain postulates in the dearth of published data. For example, the kinetics used to model the lipid-processing module were based on data collected for palm oil at similar conditions, while the details of lipid-extraction energy usage for a high-density slurry were approximated on the basis of results for low-density slurry. Furthermore, it was concluded that the income from the sale of the algal biomass byproduct of lipid-extraction is a critical factor in the profitability. Based on its protein content, this report considered the use of algal biomass as

animal feed and determined its economic worth accordingly. However, to ensure the economic success of biodiesel production, an additional analysis should focus on algal usage of biomass as a feedstock and confirm the safety of its use. Further analyses could examine other potential applications for the byproduct, including opportunities within the pharmaceutical and power generation industries. Overall, in order to convince investors that the attractive economics published in this report may be translated into actual earnings, it is critical to move beyond modeling. Pilot studies must be conducted in order to bolster the proposed algae-to-biodiesel venture with experimental data and identify possible pitfalls.

II. INTRODUCTION AND PROJECT CHARTER

A. Motivation

In 2007, the United States Government passed the Energy Independence and Security Act (EISA), which mandated the production of 36 billion gallons of biofuel per year by 2022 - of which 21 must be advanced biofuels.[1] As various investors, public research organizations and private companies investigate biofuel production on an industrial scale, two renewable sources for advanced biofuels have consistently shown high promise: cellulose and microalgal biomass.[1] Cellulose, which is often harvested from corn but may also be isolated from other grasses and woody plants, can be used to produce ethanol, an additive to gasoline.[2] While ethanol is an essential biofuel that will likely play an increasingly important role in the alternative energy market, it has a low energy density that makes it unfit for many applications, such as use as a jet fuel.[2] The consumption of corn for the purposes of ethanol production also competes with the national supply of corn to the food industry. In fact, with food prices rising sharply in recent months, "many experts are calling on countries to scale back their headlong rush into green fuel development, arguing that the combination of ambitious biofuel targets and mediocre harvests of some crucial crops is contributing to high prices, hunger and political instability."[3] Therefore, significant attention has shifted to microalgal biomass as a key renewable resource.

In comparison with other crops, algae grow faster, which makes them particularly suitable for use in biofuel production. Some algal strains not only have lipid contents greater than 50%, but also grow at rates up to 100 times faster than terrestrial plants[2]. With regards to biofuel productivity, algae are also superior. While a mere 18 gallons of oil may be extracted from each acre of corn, algae farms boast of productivities approaching 600 gallons per acre.[2] Furthermore, unlike other crops that require high volumes of fresh water, a utility that is becoming increasingly scarce, algae may grow in saltwater and even in contaminated water[4]. Since algal cultures may thrive in polluted environments rich in agricultural runoff, and they can capture and recycle phosphorous, a key nutrient rapidly declining in global reserves.[5]

In the content of the passage of EISA and a simultaneous rise in social pressure to consider algae as a suitable feedstock for fuel production, Mr. John A. Wismer from Arkema Inc. proposed a CBE 459 design project in 2010 that involved the preliminary process design for an algae-to-biofuel production facility. This design was to be accompanied by a thorough economic analysis.

The process design and economic analysis performed in 2010 suggested that the process was profitable, but required an excessive capital investment of 2.8 billion dollars.[6] This conclusion provided motivation for further investigation of the proposed design with an aim to significantly reduce the initial capital investment.

B. Project Charter

In the past year, both public and private institutions, including the Department of Energy and ExxonMobil Corporation, have launched projects with the aim of minimizing the high capital costs and other barriers associated with industrial scale algae-to-biofuel production. In the same vein, Dr. Warren Seider, a faculty member of the University of Pennsylvania, proposed this design project, which considers existing algae-to-biofuel ventures in the context of the latest research and most recent technological advances in an effort to optimize their overall process design and profitability. The design team was asked to investigate heterotrophic algae cultivation, which is a novel technique that couples traditional photosynthetic algae growth with a fermentative lipid-accumulation phase. The project scope also included a comprehensive reanalysis of the OriginOil™ Single-Step Extraction technology in light of recent patent publications. The completed design and simulation of a transesterification lipid-processing module using Aspen PLUS software was also proposed. These results were to be joined together in a detailed economic analysis that optimizes profitability while making key recommendations regarding the potential promise for an algae-to-biodiesel venture.

C. Innovation Map

Essential to optimizing existing algae-to-biofuel ventures is a thorough understanding of the novel process and material technologies that promise to inspire improvements and progress. Shown in figure 1 is an innovation map that may be used to relate the voice of the consumer to the new material and process technologies that drive product development in a visual manner.

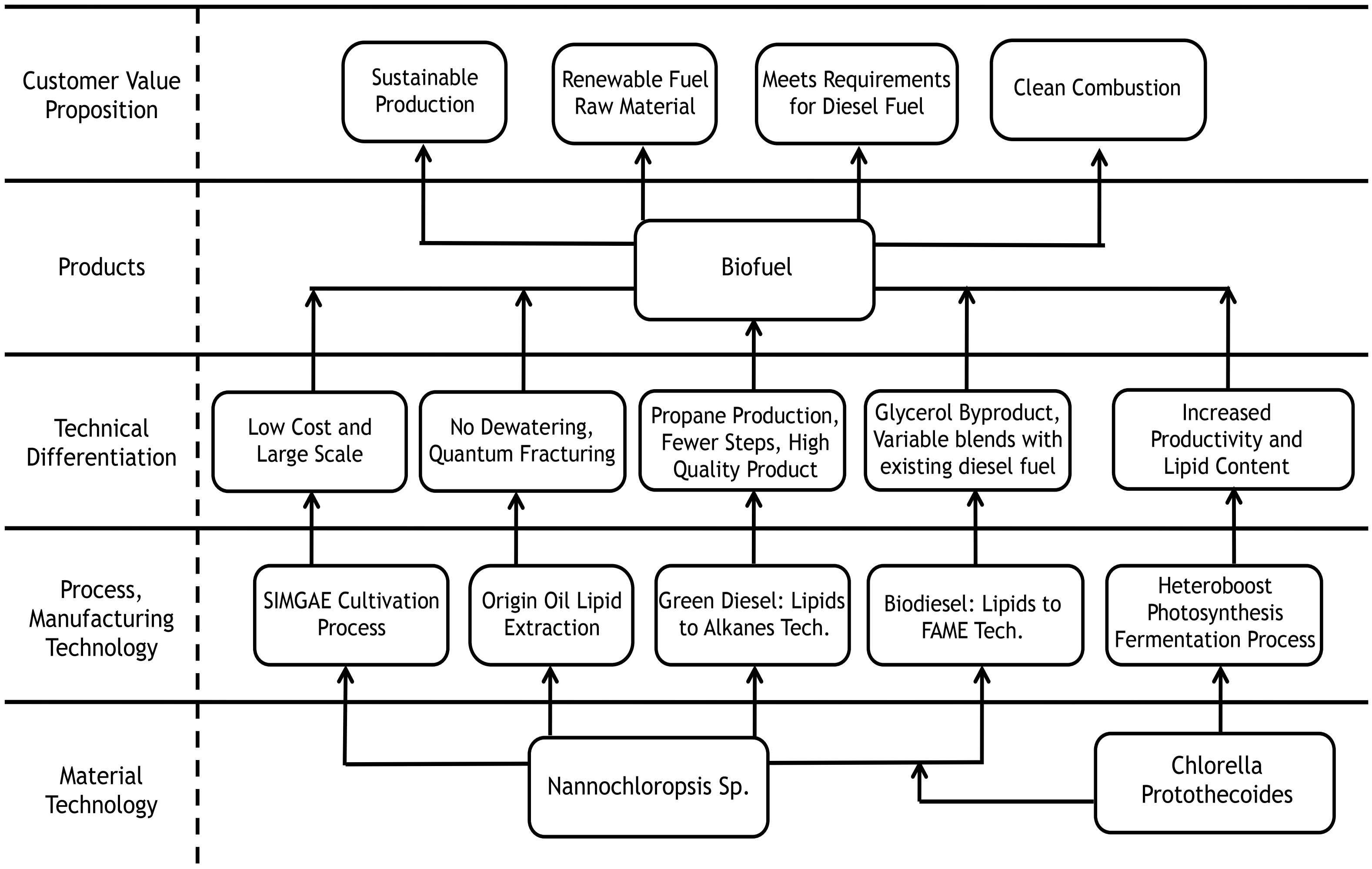


Figure 1. The innovation map relates the consumer to novel materials and process technologies.

The many dozens of algal strains in existence worldwide comprise the material technologies available for biodiesel production from an algal feedstock. In the past, *Nannochloropsis sp.* was identified as a key algal species for renewable biomass generation and was studied in terms of the Simgae™ Algal Produced Biomass System. However, research on the novel heteroboost process technology, which combines traditional cultivation with fermentation, identifies the species *Chlorella protothecoides* as the algal feedstock.[11] Although algae strains that are genetically engineered for superior biofuel productivity are only in the research and development phase, they may one day join natural algal species as available material technologies for biofuel production.

One processing technology associated with algal cultivation is the Simgae™ Algal Biomass System. This system involves the photosynthesis of algae suspended in long, clear polyethylene tubes that permit the transmission of sunlight, as well as temperature, pressure and compositional control. The Simgae™ model differentiates itself by being extremely low cost, yet amenable to industrial-scale implementation. Another cultivation technology, involving the use of the *Chlorella protothecoides* algal strain, is the heteroboost cultivation model. This cultivation model involves an autotrophic phase, during which algae fix CO_2 by photosynthesis, followed by a heterotrophic phase, during which algae metabolize an available energy source and significantly increase their lipid-content. Since autotrophic growth is very slow, coupling it with heterotrophic growth vastly accelerates both biomass and lipid generation. The heteroboost model does not specify a particular cultivation method, and therefore may be used in concert with the Simgae™ model; it is seen as a complementary process rather than an alternative.

OriginOil™ advertises a lipid extraction processing technology that is also key to the algae-to-biodiesel venture that eliminates the need for dewatering and other energy intensive steps associated with traditional solvent lipid-extraction techniques.[69] The OriginOil™ Single-Step extraction methodology involves the generation of shockwave-inducing micro and nano-bubbles in addition to the use of electromagnetic pulses as a means of breaking open the algal cell wall with high energy efficiency.

Both transesterification and catalytic hydrotreating are two key processing technologies that consume a lipid feedstock and produce a transportation fuel. Through catalytic hydrotreating, triglycerides are converted to n-alkanes, which may be a direct substitute for diesel fuel in the transportation fuels market and used in engines with little or no modification.[84] Propane and fuel gas, both of which are valuable, are generated as byproducts. Since catalytic hydrotreating has already been developed as a refining technique for other feedstocks, it is often cited as less costly to design and implement at existing oil refineries and fuel production plants. Transesterification also consumes triglycerides, but converts them to fatty acid methyl esters (FAME). Sold as biodiesel fuel, FAME is rarely delivered to an engine in its pure form and is more commonly mixed with petroleum diesel in specified blends, such as B5 and B20, which are 5% and 20% FAME respectively.[84] During transesterification, crude glycerol is generated as a byproduct. Although pure glycerol is a valuable chemical commodity, the supply of crude glycerol has risen steadily in the past decade as biodiesel production expands, leading to plummeting prices for this byproduct.

III. OVERALL CONCEPT STAGE

A. Market and Competitive Analysis

As the 21st century advances, there is growing sociopolitical pressure to develop suitable renewable substitutes for the petroleum-based fuels currently consumed globally at a furious pace. Among these renewable substitutes, there is significant interest in the development of a biofuel that can compete successfully in the transportation fuel market. Such a departure from traditional petroleum-based fuel serves the interests of a variety of consumers and impacts a wide array of markets. Governmental organizations often see the development of a biofuel venture as a means of creating a significant number of jobs for the energy industry while simultaneously reducing dependence on foreign powers for oil supplies. Environmentally conscious organizations and businesses support biofuel production as a sustainable operation with potentially net neutral carbon dioxide generation. Additionally, the automotive industry has seized upon the notion that the consumption of biofuels dramatically reduces the emission of pollutants, enabling them to meet strict standards set by environmental protection agencies.

The possibility that biofuels may be adopted as valid alternatives to petroleum-based transportation fuels indicates that the potential exists for a vast consumer market. Although biofuels derived from algae are often positioned in opposition to biofuels generated from other oil sources, such as corn and soybeans, algae is a crop that occupies a unique niche in the alternative fuels market. Aside from far surpassing other crops in oil productivity and growth rate, algae have the potential to grow in salty and contaminated water, capturing phosphorous from runoff, thereby enabling their cultivation in regions that are unsuitable for traditional agriculture. Furthermore, algal farming on an industrial scale does not compete with global food production, a serious concern regarding biofuels derived from corn and other key farm products.[3] Algae are even used to produce a distinct high-energy density diesel fuel that lends itself to different applications than the low energy-density cellulosic ethanol derived from corn and other crops. Considering these distinguishing characteristics, the algae-to-biofuel venture faces few barriers other than the high costs of investment and lack of experimental data that makes scaling up algal biofuel production challenging.

Currently, fatty acid methyl ester (FAME) biodiesel, which may be produced from algal and other vegetable oils via transesterification, is the most widely consumed commercialized alternative fuel. In 2009, the US consumed 877,000 barrels of biodiesel per day, a significant volume, but one that is still dwarfed by the 18,771,000 barrels of petroleum oil consumed daily.[7] This process design considers a daily biodiesel production level of 17,500 barrels. Since this figure represents just 2% of the current market, sufficient demand is expected for the biodiesel product.[83] According to the most recent sales projections and analysis conducted by the Chemical Industry News and Intelligence Service (ICIS), pure FAME biodiesel fuel is currently priced at $3.30 per gallon.[93] However, this price may change drastically in the future since the economics of biodiesel are highly sensitive to the political environment. The adoption of legislation that provides various tax incentives to biodiesel producers and the development of carbon emission programs that create demand for carbon dioxide consumption technologies have the potential to impact the market for alternative fuels.

B. Customer Requirements

In order to function within the context of the current liquid transportation fuel market, biodiesel produced from algae must meet certain industrial standards of purity. Not only must the hydrocarbon chain length range from 12-22 carbon atoms, but the fuel must meet the official US ASTM 6751 Industrial Standards. Listed in table I, the US ASTM 6751 Industrial Standards cover a range of physical, chemical and thermodynamic properties, including kinematic viscosity, oxidation stability, flash point and alcohol content.[84] Achieving these standards is a key concern of biofuel production and requires a carefully operated and monitored separation train.

Customers rarely consume biodiesel in its pure form. Since few engines may run pure biodiesel, known as B100, without increased maintenance or modification, many consumers seek specified blends of biodiesel diluted in petrodiesel. B20, for example, is a mixture of 20% biodiesel in petrodiesel. Although the petrodiesel blends still require fossil-fuel derived petroleum diesel, blends that contain less than 20% biodiesel may be used in existing diesel equipment without any special modifications.[84]

Table I[82]. The US ASTM D6751 Industrial Standards for Biodiesel Purification

Biodiesel Standards ASTM D6751 (United States)			
Property	**Test Method**	**Limits**	**Units**
Flash point (closed up)			
Alcohol Control One of the following must be met: 1. Methanol Content 2. Flash point	EN 14110 D 93	0.2 max 130 min	mass % °C
Water and sediment	D 2709	0.050 max	% volume
Kinematic viscosity, 40°C	D 445	1.9-6.0	mm^2/s
Sulfated ash	D 874	0.020 max	% mass
Sulfur	D 5453	0.0015 max or 0.05 max[a]	% mass
Copper strip corrosion	D 130	No. 3 max	
Cetane number	D 613	47 min	
Cloud point	D 2500	Report	°C
Cold soak filterability	Annex A1	360 max	seconds
Carbon residue (100% sample)	D 4530	0.050 max	% mass
Acid number	D 664	0.50 max	mg KOH/g
Free glycerin	D 6584	0.020 max	% mass
Total glycerin	D 6584	0.240 max	% mass
Distillation temperature, atmosphereic equivalent temperature, 90% recovered	D 1160	360 max	°C
Oxidation stability	EN 14112	3 min	h
Calcium and Magnesium, combined	EN 14538	5 max	ppm (μm/g)
Sodium and Potassium, combined	EN14538	5 max	ppm (μm/g)
Phosphorus content	D 4951	0.001 max	% mass

C. Overall Preliminary Process Synthesis

The overall algae-to-biodiesel venture involves three separate modules. Several alternatives exist for each module, and will be fully considered and discussed in the sections that follow. The most promising overall flowsheet is presented below in figure 2. In the first module, algae are grown autotrophically via the Simgae™ cultivation process in the presence of carbon dioxide and nutrients. They are then sent to fermentation tanks, where they metabolize glycerol in the presence of oxygen, and increase their both their productivity and lipid content. From Module I, the algae are sent to Module II where the OriginOil™ Single-Step Extraction technology is used for lipid extraction. The lipids isolated in Module II are then transesterified in Module III to produce FAME, which is refined and sold as pure biodiesel.

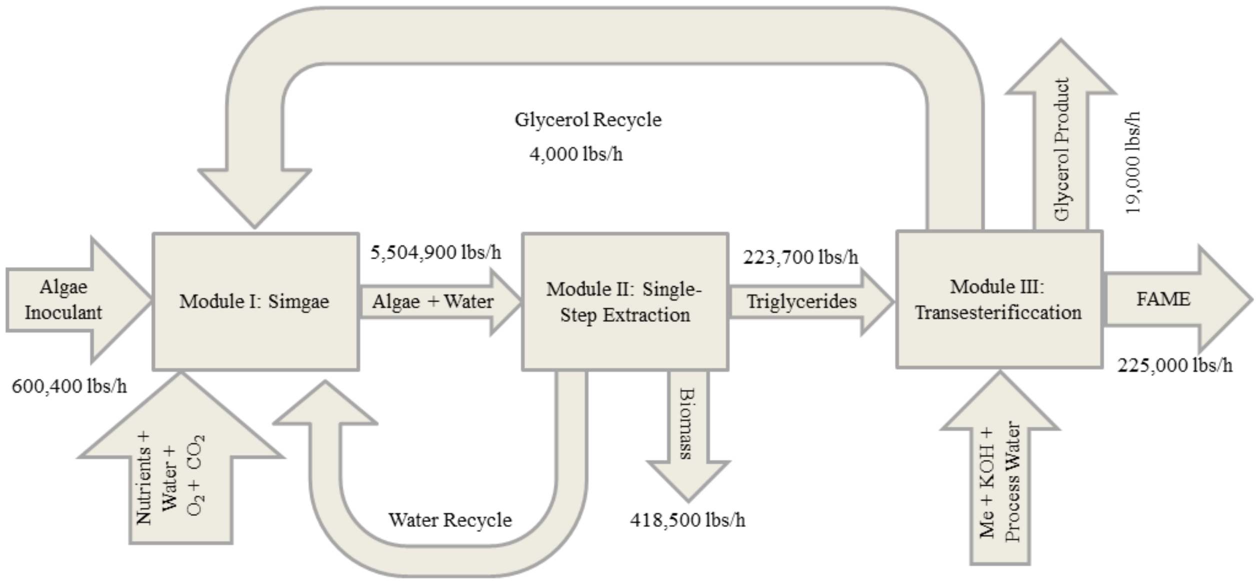


Figure 2. Illustrates the overall flowsheet of the algae-to-biodiesel process

What follows is a thorough investigation of each of the three modules. Each module was divided into the following three sections: Concept Stage, Manufacturing & Development, and Feasibility & Economic Analysis. The overall economics of the algae-to-biodiesel venture will be presented in the Overall Feasibility & Economic section, followed by the conclusion and a series of recommendations. The modules will be presented in the following order: algae cultivation, lipid extraction and tranesterification.

IV. MODULE I: ALGAE CULTIVATION

1. Algae Cultivation: Concept Stage

A. Bench-Scale Laboratory Work: An Algae Cultivation Model

Recent research in the area of algae cultivation has been driven by a growing interest in the algae-to-biofuel venture. Although biofuel derived from algal lipids is regarded as having the potential to fully replace fossil fuels, the realization of such potential is not yet in sight. This is primarily due to the absence of a cost-effective strategy for large-scale cultivation, which has thus far deemed the process economically infeasible on an industrial scale.[8] The process design and economic analysis performed by the CBE 459 design team in 2010 suggested that while the algae-to-biofuel process could be profitable, it required a staggering capital investment on the order of $2.8 billion.[6] Of the total $2.8 billion capital investment, $2.2 billion accounted for the cost of land required for algae cultivation. Given that large volumes of algae cultures are required to achieve high biodiesel production rates, recent studies have focused on optimizing the process of algae cultivation to maximize not only the algae growth rate, but also the algal cell lipid content. Algae cells are generally cultivated using one of the following growth models: autotrophic, heterotrophic or mixotrophic.[9] What follows is a discussion of the aforementioned cultivation models that highlights their key characteristics, advantages, and disadvantages. Because the process design proposed for algae cultivation is highly dependent on the cultivation model adopted, selecting a cultivation model early in the design process is critical.

Autotrophic Cultivation

The autotrophic growth model allows for carbon-dioxide fixation by photosynthesis, which is defined as the utilization of water, carbon dioxide, chlorophyll and light energy by plants.[10] Algae cells that are grown autotrophically typically require the following conditions: exposure to natural or artificial light, a supply of aseptic air, and a supply of carbon dioxide gas.[10] Autotrophic growth, also known as phototrophic growth, may occur in either open or closed architectures, such as ponds or photobioreactors respectively. One of the major challenges of phototrophic growth is that algal strains with rapid growth rates accumulate relatively small quantities of lipid in their biomass (<20% of dry weight), while those that accumulate relatively high amounts of lipids (40-50% of dry weight) exhibit very slow growth rates.[11] Moreover, biomass productivity is relatively low during autotrophic growth in comparision to other modes of cultivation.

Heterotrophic Cultivation

Non-obligate phototrophic algae cells may be grown heterotrophically when situated in a carbon-rich environment in the dark, or under conditions of very dim lighting [12] In *Chlorella protothecoides* algae, Xiong et al., (1992) report a very high productivity under such conditions and a lipid content up to 58%, roughly three times the lipid content reported for autotrophic cultures.[7] Rates of heterotrophic growth in the dark are reported to be comparable to those of autotrophic growth, although the rate of carbon dioxide fixation under heterotrophic conditions is minimal.[13]

Mixotrophic Cultivation

Mixotrophic cultivation combines the autotrophic and heterotrophic growth models. It is important to note that the maintenance of the photosynthetic cell machinery requires the generation of large quantities of algal pigments, which require a nitrogen source for proper generation. Yet, to achieve neutral lipid accumulation during heterotrophic growth, nitrogen-limited conditions are required.[14] Establishing nitrogen-limited conditions while maintaining an adequate level of photosynthetic pigments is challenging and a key drawback of mixotrophic cultivation. Another disadvantage involves glucose-bleaching, which occurs when algae are grown in a carbon-rich source and results in the biodegradation of chlorophyll and the disruption of photosynthesis.[11]

B. Picking a Microalgae Strain

Although algae are generally depicted as common plant-like, photosynthetic and aquatic organisms, not all species and strains of algae are suitable for cultivation and biofuel production. This is mainly a consequence of the fact that different species of algae grow at different rates, photosynthesize at different rates, and have different lipid contents at maturity, all under very different cultivation conditions[1]. While all algae strains are composed of proteins, carbohydrates, fats and nucleic acids, the percentages of each of those components vary significantly with algal species. For the purposes of biofuel production, it is not only the lipid content that is significant, but also the fraction of lipids that is triglycerides, since this is the lipid component that may be processed into biofuel.[9] One of the most basic distinctions between different algal strains is whether they are fresh-water or salt-water strains. Choosing a salt-water algae strain allows one to avoid the consumption of fresh water, which is considerably more expensive and less available than saline water. Based on the location chosen for cultivation, fresh water may not even be a viable option. The use of algae strains that are tolerant to produced water – an aspect of algae cultivation that has been thoroughly investigated by Eldorado Biofuels®, VM Technology – could be even more advantageous.[15] This technology is primarily concerned with oil deposits that are associated with significant bodies of underground water, known as "formation water." In the course of oil and gas production, this water is pumped out of the ground as "produced water," which contains salts and minerals, in addition to contaminants such as hydrogen sulfide, ethyl benzene and xylene. If these toxins are removed, then this produced water is the perfect substrate for algal growth.[15] Currently, most produced water is re-injected into commercial disposal wells, at a cost of \$0.30-\$10.00 per barrel. However, the ability to make use of this technology is dependent on location, as an existing infrastructure for oil drilling and processing is a must. Picking an algal strain is therefore heavily influenced by the following factors: cultivation model and location.

C. Process Synthesis: Algal Biomass Cultivation Systems

A critical decision to be made in algae cultivation is the choice between closed or open systems. Because the location proposed for algae cultivation is highly dependent on the cultivation model adopted, once again, it is clear that selecting a cultivation model at an early stage of the design process is critical. Shown and discussed below are some of the most common closed and open architectures used for algae cultivation, as well as some of the advantages and disadvantages of each system. The architectures presented below are only applicable to the photosynthetic growth model. Heterotrophic growth is essentially identical to fermentation, which is concerned with neither maximizing surface area, nor light attenuation in algal cultures.

Open Cultivation Systems

Open architecture approaches, such as traditional raceways and ponds, present the cheapest and simplest of all current options; expensive cultivation systems need not be installed, as large open bodies of water can be utilized for the cultivation of algae. A raceway pond for example, as described by the National Renewable Energy Laboratory (NREL), is an artificial and shallow pond, in which algae are kept suspended in water and are circulated around a racetrack.[16] The depth of the raceways is limited by the extent to which light can penetrate the water body. The ponds are supplied with the required nutrients and adequate CO_2 levels to maintain viability of the algae, while the algae slurry is removed at the end of the raceway. Flow of the algae and water body is maintained by paddlewheels. Shown in figure 3 is a picture of Earthrise Farms raceway ponds that are currently in use for microalgae production in California, USA. Although cheap and simple, open architecture approaches suffer major drawbacks with regards to contamination, maintenance of adequate CO_2 concentration levels, maintainability, evaporation and temperature control. Other disadvantages include the absence of an adequate stirring or agitation mechanism that would reduce mass transfer limitations and increase productivity; agitation would not only serve to facilitate fluctuating light regimes, but it would also increase the mass transfer rates between the surrounding growth medium and the cultured algae.[17]

Figure 3: Raceway Ponds for Microalgae Production. Earthrise Farms, California.

Figure 4: CalPoly® Controlled Environment Agriculture Group Tubular photobioreactors (PBRs)

Closed Cultivation Systems

Closed architecture approaches, such as the photobioreactors (PBRs) shown in figure 4, represent an extreme alternative to open architecture systems.[18]

PBRs are defined as closed culture systems for phototrophs, in which a great proportion of the light (>90%) does not impinge directly on the culture surface, but has to pass through transparent reactor walls to reach the cultivated cells."[19] The advantages of using a closed system for the cultivation of algae, as opposed to an open system are numerous. The use of a closed system protects the algae from invasions by competing microorganisms, allows for genetic stability, and provides an environment in which conditions may be better controlled.[20] These factors combined ensure the dominance of the desired species and allow for control of the production of desired products. Optimization of PBRs to ensure maximum utilization of incident solar radiation, effective mixing and reduced mass transfer limitations serve to increase overall productivity of the system. Moreover, agitation in the system is more easily controlled and will lead to a reduction in photo-inhibition and shading – two phenomena that are more likely to occur in high cell density cultivations receiving high concentrations of solar radiation.[21]

Nonetheless, the construction and installation of PBRs with such functionality for at an industrial scale would require a combination of sophisticated materials and design. Therefore, a large-scale cultivation system based on PBRs comes at a much more elevated cost than simple raceway ponds. Moreover, one of the main disadvantages of PBRs is the deposition and growth of algae on the reactor walls, which reduces the wall clarity and in turn limits the amount of sunlight that reaches the photosynthetic algae. Adequate agitation and mixing would also be necessary to prevent temperature and concentration gradients from developing throughout the reactor.

Valcent's Vertigro® Algae Biofuel Oil/CO_2 Sequestration System

Another cultivation system to consider is the Vertigro® system that consists of a series of vertical and closely packed bioreactors.[22] These reactors, shown in figure 5 below, are made of thin film membranes that allow for high levels of light penetration and are configured for optimum flow for the growth of algae.[23] The reactors shown are composed of plastic bags that hang from light aluminum scaffolding, with sprinkler systems such as those used in mechanized farms. According to Global Green Solutions Inc, the sponsor for Valcent's demonstration pilot plant, the algae consume up to 90% of their weight in CO_2, while utilizing sunlight as their energy source for photosynthesis. It is also reported that the algal species accumulate lipids up to 50% of their dry weight – lipids that will be oil-suitable for bio-fuel blending with diesel.

According to data obtained from a continuously operating test bed facility at Valcent, the apparatus could yield up to 4,000 barrels of oil per acre per year, at an estimated cost of $20 per barrel. The system may be built on non-crop lands that are close to major city markets and can be implemented in a variety of environments, two key advantages. The system is also described as modular and easy scalable.[24] Global Green Solutions Inc. has recently announced that it will sponsor the construction of a demonstration pilot plant of the Vertigro® biofuel Oil/CO_2 Sequestration System in El Paso, Texas, at an initial capital investment of $2,500,000. However, while a system such as Vertigro® seems to necesitates little to no heavy engineering, it nonetheless requires a very water intensive operation with a heavy pumping requirement.[25] Moreover, at the moment, Vertigro® is in the proof-of-concept phase and at scale of operation very far from becoming commercial.

Figure 5: Vertigro® vertical and closely packed algae bioreactors.

The Algae Tree

The Algal Tree for the production of biomass and biofuels is proposed as a component of an integrated sustainable communities proposal that combines biomass and biofuel production with solar and wind energy technologies.[26] The design of the cultivation system, an "Emerald Forest" composed of tree-like photobioreactors shown in figure 6, is chosen because of trees' inherent efficiency in capturing incident sunlight. The system would be run as a batch process whereby the photobioreactors are filled with seawater and algae, and are supplied with CO_2 and nutrients on a continuous basis. The algae would theoretically remain within the bark or shaft of the tree, while the sunlight is captured by the "branches" and "leaves" of the tree. These "leaves" and "branches" consist of optical materials that not only distribute the captured light, but

Figure 6: A possible design for the "Algae Tree" PBRs.

also direct sunlight towards the algae in the bark of the algae tree. Once lipids have been extracted from the algae, the remaining biomass is sent to a gasification plant. The fuel gas produced through the gasification of the biomass is then used by a nearby power generating plant. The authors state that cornerstone of sustainability in this design is the utilization of CO_2 produced in the gasification plant as a source of nutrition.[26] Such a facility would ideally be located on arid and desert lands that are close to large bodies of water. The authors state that several areas in the United States would be suitable for such a facility, such as the Southeast Cost, the Western States, parts of Nevada, and the Gulf of Mexico.[26] The authors state that the African Sahara desert alone could produce enough liquid fuels to meet a considerable fraction of the world's liquid fuels need. While the technology may seem innovative and promising, several barriers must be overcome before this design is efficiently implemented at an industrial large-scale. The biggest obstacle to this design is that its efficiency relies on the establishment of integrated and sustainable communities; the "Emerald Forest" is a part of a whole.[26] The complexity of the design, and its use of sophisticated materials would not only complicate the scale-up of such a system, but would also lead to very high operation and maintenance costs.

D. Location Screening

Any location chosen for the cultivation of algae must have the following two features: a nearby water source, and a means of transportation. The heterotrophic, autotrophic and mixotrophic cultivation models all require that algae cells be suspended in an aqueous medium. A means of transportation is also required for moving the algae from the site of cultivation to the site of lipid extraction. However, based on the choice of cultivation model and cultivation process, certain additional features will be of paramount importance. For autotrophic growth, a nearby CO_2 source, and a constant source of light are critical. In the case of heterotrophic growth, sunlight and carbon dioxide are not critical to algae growth, but oxygen and carbon-rich organic compounds are required. Because the choice of location is very specific to the cultivation model and cultivation process of choice, more detail with regards to location is presented after the proposed model has been identified in section E under "Cultivation Location: Thompsons, Texas, USA", on page 32.

E. Proposed Module I Parameters

In light of the above discussion, presented below is the set of four cultivation parameters chosen for this specific design project: cultivation model, algae strain, cultivation process and location. While the reasoning behind each decision is clearly outlined, one must bear in mind that each parameter is highly dependent on several factors, each of which must be evaluated separately at any given point in time. The cultivation process of choice will be examined more thoroughly under the section titled *Feasibility & Development Stages*, and the associated process diagrams, material and energy balances, and equipment list and unit descriptions are shown.

Cultivation Model: Photosynthetic-Fermentation Model

For this design project, a photosynthesis-fermentation model (PFM) was adopted for the cultivation of the algae. This model is based on a study performed by Miao and Wu (2004), and was primarily proposed to overcome the dualistic challenge of autotrophic cultivation of algae: algal strains with relatively rapid growth rates accumulate lower amount of lipids in their biomass (<20% of dry weight), whereas those that accumulate relatively higher amounts of lipids (40-50% of dry weight) exhibit very low growth rates. [27, 2] The PFM model capitalizes on a double CO_2-fixation process, in which a *Chlorella protothecoides* microalgal culture is first grown autotrophically to increase its biomass, and then metabolized heterotrophically to maximize its cell density and lipid content[5]. Although the mixotrophic model also combines autotrophic and heterotrophic cultivation, the main distinction between the mixotrophic model and PFM is the separation between the autotrophic and heterotrophic phases. Xiong et al. (1992) emphasize that whereas the mixotrophic model combines both phases, the PFM model separates the culture process into two independent but sequential steps: the nitrogen-sufficient photosynthetic cultivation, and the nitrogen-deficient heterotrophic growth. [9]

The study cites the conversion ratio of glucose to biomass and glucose to oil as important indicators for biofuel feedstock production, and reports a biomass yield of 0.617 $g/g_{glucose}$ and a lipid yield of 0.298 $g/g_{glucose}$ in PFM.[9] These values were contrasted to those reported for the heterotrophic or fermentation model (FM) alone, which are 0.33 $g/g_{glucose}$ and a lipid yield of

0.176 g/$g_{glucose}$. This suggests that the combined approach of autotrophic (PM) and heterotrophic (FM) growth allows for a 69% increase in the accumulated lipid content. The study also reports average cell growth rates of 0.286 g $L^{-1}d^{-1}$ and 23.9 g $L^{-1}d^{-1}$ in the PM and FM stages, respectively, with a maximum achievable cell density of 123 g L^{-1} in the fermentation model stage.[9] After optimization of the photobioreactor in the FM, a lipid content of 58.4 dry-wt% was reported.

The main obstacle to the broader commercialization of biodiesel production is the relatively high cost of the oil feedstock, just as when making bread, the cost is mainly due to flour and egg, and not the baking process.[28] Since the commercialization of any process is highly dependent on the feasibility of industrial production, a complementary study on the large-scale cultivation of *C. protothecoides* was performed by Xiufeng et al. (2007) to determine whether the same high lipid content and high biomass concentration could be achieved.[28] The cultivation of *C. protothecoides* was performed at four different scales: in an aeration flask, a lab-scale bioreactor, a pilot-scale bioreactor and a commercial scale bioreactor. Because this project is concerned with large-scale production at 17,560 bpd of biodiesel fuel, the experiment run in the commercial scale bioreactor will be most relevant to this discussion. The commercial scale cultivation was performed in an 11,000 L stirred tank bioreactor containing 8,000 L medium, with the same algal strain used in the original study by Miao and Wu et al. (2004). Organosilicon (0.1%) and chloromycetin (0.01 g L^{-1}) were reportedly added to the commercial-scale bioreactor to reduce bacterial contamination and the formation of foam.[28] pH was maintained at a value slightly over 6.0 by the addition of K_2HPO_4 and KH_2PO_4 at a ratio of 0.86, and aeration rate and agitation speed were set at 25 m^3 h^{-1} and 180 rpm, respectively. Xiufeng et al. (2007) assert that as cultivation was scaled up, the lipid content of *C. protothecoides* decreased to roughly 43 dry-wt%, whereas the cell density achieved the same level in both pilot and commercial-scale bioreactors.[28] The authors therefore conclude that their experiments suggest the feasibility of industrialization of the process, while asserting the need for further optimization of cultivation technique to avoid a significant decrease in lipid content as the process is scaled up. In light of the above discussion of enhanced productivity and lipid content, the photosynthetic-fermentation growth model, with its associated parameters, was proposed for this design project.

Algae Strain: Chlorella protothecoides

While the selection of saltwater algae strains allows one to avoid the competitive market for freshwater, the high biomass and lipid content achieved in the study performed by Xiufeng et al. (1992) are specific to *Chlorella protothecoides,* which is a freshwater algae strain.[28] For the purposes of this design project, *C. protothecoides* was assumed to have a lipid content of 46 dry-wt%, and to achieve a cell density of roughly 0.2 $g.L^{-1}$ in the PM stage, and a cell density of 120 $g.L^{-1}$ in the FM stage. While all algae strains are composed of proteins, carbohydrates, fats and nucleic acids, the percentages of each of those components vary significantly with the species of algae. For the purposes of biofuel production, it is not only the lipid content that is significant, but also the fraction of lipids that is triglycerides. Algal oil isolated from *Chlorella protothecoides* is not uniform in chemical species, but is rather comprised of a variety of triglycerides. Researchers regularly use gas chromatography to analyze biodiesel once is has been isolated in order to determine the compositions of triglycerides in a particular algal species. As shown in table II below, biodiesel fuel produced from *Chlorella protocoides* is mostly comprised of 9-octadecenoic acid methyl ester, 9, 12-octadeadienoic acid methyl ester, and hexadecanoic acid methyl ester.[28] The composition of triglycerides in *C. protothecoides* is not directly relevant to algae cultivation, but is critical for the discussion of the transesterification process, presented as Module III.

Table II. Composition of Triglcyerides extracted from *Chlorella protothecoides*

Component	Chemical Formula	Fatty Acid Chain Species	Relative Triglyceride Content (%)
Triglyceride 1	$C_{57}H_{104}O_6$	$C_{19}H_{36}O_2$	69.8
Triglyceride 2	$C_{57}H_{98}O_6$	$C_{19}H_{34}O_2$	19.5
Triglyceride 3	$C_{51}H_{98}O_6$	$C_{17}H_{34}O_2$	10.7

Cultivation Process: Simgae™ & Fermentation Tanks

The cultivation design proposed for this project must be compatible with the proposed cultivation model. This implies that the cultivation process must have two stages: a photosynthetic autotrophic phase (PM) and a fermentation heterotrophic phase (FM). The algal biomass production system proposed for the photosynthesis model (PM) stage of this project is the Simgae™ cultivation system that was developed by XL Renewables, Inc. and is currently licensed by the Diversified Energy ® Corporation (DEC)[29]. Simgae™, which stands for "simple" and "algae," is advertised as offering a low cost and simple approach to growing algae at large scales, thereby overcoming the significant capital and operations and maintenance costs required for the construction and maintenance of large-scale cultivation systems.[30] This system consists of clear and flexible polyethylene tubes through which the algae fluid (algae cells + nutrient medium) may pass while undergoing photosynthesis in the presence of sunlight. Because of the low cell density achievable in autotrophic growth, a significant amount of land will be required to lay down the tubes to ensure the desired algal mass production rate. For the fermentation model (FM) stage of this process, industrial-scale fermentation tanks will be utilized. The framework, layout and specifications of the cultivation fields, the reactor tubing and the fermentation tanks are discussed in great detail in the feasibility and development stage of Module I.

Proposed Location: Thompsons, Texas, USA

Because the PFM model combines both autotrophic and heterotrophic cultivation, the following features are critical for the sustainability of the cultivation process: a nearby water source, a means of transportation, a nearby CO_2 source, flat land, a constant source of light, oxygen and a source of carbon-rich nutrients. The location chosen for this project is the W.A. Parish Electric Power Generating Station, located in Thompsons, Texas. There are three main reasons why this location is suitable for this project. Firstly, the land surrounding the power plant is flat, as shown in the topographic map in figure 7.[104] For the purposes of this project, the algae tubes will lay at an incline of 3-5°; the presence of flat land will allow for installation of ramps, earthen beds and manipulation of the layout of the fields as desired. Secondly, this location has a nearby CO_2 source, which is the flue gas emitted by the W. A. Parish Electric power generating station. Thirdly, the town of Thompsons is located in Fort Bend County, within the Houston-Sugar-

Figure 7: A topographic map of the proposed location: Thompsons, Texas.

Land-Baytown metropolitan area, which is the largest economic center of the South Central United States.[31] The heavy concentration of petrochemical industries in the Houston area therefore requires the presence of a well-established infrastructure for freight rail transportation to facilitate the import and export of products.

The choice of location may depend on other factors and costs, such as property tax rates, as well as the interplay between the above factors; for example, the discouragingly high cost of fresh water required for this project is offset by the presence of a nearby electric power generating plant, and a relatively constant source of natural sunlight. Moreover, while the need to move a significant amount of soil may be prohibitive, the difficulty and cost of transportation are offset by the presence of a well-established infrastructure for freight rail transportation. Shown below is a more detailed discussion of how each of the afore-mentioned criteria will be met at this site.

- ***A Nearby Water Source***

Fresh water, which is water that contains no more than 250 - 300 mg of dissolved solids per liter of water is required for the cultivation of *C. protothecoides*.[32] Texas sits on top of the Ogallala Aquifer and the Edwards Aquifer, which are both fresh water aquifers.[33] However, given that more than 1 million people depend on the water supply from the Edwards Aquifer alone, the amount of water required for cultivation may not be met by these sources.[34] A fraction of the fresh water required for cultivation may therefore be obtained from local water desalination plants. Texas is home to one of the world's largest inland water desalination plants, The El Paso Public Water Service Utilities Board, which produces up to 27.5 million gallons of fresh water per day.[35] Water permits will be required to allow for use of this water.

- ***A Means of Transportation***

The heavy concentration of petrochemical industries in the Houston area within a 30-60 mile radius of the proposed plant location has resulted in the presence of a well-established infrastructure for freight rail transportation [36] According to the Texas Department of Transportation, in 2008, Texas had a total of 44 freight railroads, with a total rail mileage of 14,982, and a total rail capacity of roughly 400 million tons.[37] Transportation via freight rail will be key for the transportation of soil.

- ***A Nearby CO_2 Source***

The W.A. Parish Electric Generating Station, operated by NRG Texas LLC. in Thompsons, is a coal-fired and natural gas-fired power plant that comprises 8 generating units. Units 1 through 4 burn natural gas to produce a total of 1,190 MW of power, whereas units 5 through 8 burn coal to generate 2,475 MW of power.[38] The flue gas emitted from this station will be used as a source of CO_2 for photosynthesis, as the amount of CO_2 in air (0.04 vol.-%) is not sufficient to allow for significant CO_2 fixation into organic hydrocarbons. Although flue gas contains SO_x and NO_x, these gases do not have an adverse effect on algae growth, but are instead consumed as nutrients by the algae cells.[39]

- ***A Constant Source of Sunlight***

According to the National Renewable Energy Laboratory (NREL), Thompsons receives as low as 112 Wm^{-2} during the month of January and as high as 250 Wm^{-2} during the month of June.[40] While Xiong et al. report high growth rates under an illumination of 29.5 Wm^{-2}, Tilzer et al. (1983) report that the fractional absorption of light by photosynthetic pigments shows wide seasonal fluctuations and ranges between 4 - 70%.[41] This implies that a minimum absorption of 26% is required under conditions of 112 Wm^{-2} to ensure the reported high growth rates. However, when exposed to prolonged periods of high light intensities, *Chlorella*'s lack of tolerance to such conditions will cause it to bleach. Algal bleaching is a process by which carotene, chlorophyll (a) and chlorophyll (9b) and the carotenols pigments degrade.[42] This can cause a rapid decline in the algae's productivity. Under conditions of 250 Wm^{-2}, absorption of a mere 12% of the incident sunlight would be sufficient to ensure optimum growth rates. Measures will therefore be taken to minimize the bleaching effect, such as reducing the autotrophic residence time of the algae.

- ***The Climate***

Temperatures of roughly 25 and 28°C were reported as ideal for autotrophic and heterotrophic growth respectively.[9] According to the NREL, temperatures in Thompson range between 10°C to 16°C between November and March, and from 16°C to 27°C between the months of April and October.[40] To maintain the algal cultures at their optimum temperature, auxiliary heating will be required to maintain the temperature of the PM medium at the desired temperature before it is pumped into the reactor tubes.

- ***A Source of Glycerol***

Glycerol is a byproduct of the transesterification process that converts extracted algal triglycerides into biodiesel. Transportation of glycerol is facilitated by the fact that the plant for processing the extracted lipids will be located nearby the algal cultivation site.

- ***A Source of Soil***

A reactor area with an incline of roughly 2° will have an elevation of 10 feet. For a reactor area with the proposed dimensions, a soil volume of roughly 93,000 yd^3 will be required to be amassed per field. To ensure the integrity of the structural support, soils that qualify must have the following qualities: strength, high resistance to erosion and chemical attack, low moisture absorption, and limited shrink/swell reactions. Because the Houston-Sugar-Land-Baytown metropolitan area is the largest economic center of the South Central United States, there is an abundance of geotechnical and civil construction contractors in the area. Transportation via freight rail will be utilized for the transportation of the soil.[31]

2. PFM Cultivation: Manufacturing & Development Stages

The algal biomass cultivation system proposed for the photosynthesis model (PM) stage is the Simgae™ algal biomass production system that is currently licensed by the Diversified Energy® Corporation (DEC). For the fermentation model (FM) stage of this process, industrial-scale fermentation tanks will be utilized. What follows is a thorough examination of the design of the proposed algae production system, as well as its utilities and equipment requirements. Of particular interest is the design of the reactor tubes that will be used for the autotrophic cultivation stage, which, unless otherwise stated, is based on Patent #US2008/0311649, "Pressurized Flexible Tubing System for Producing Algae." [44]

A. Proposed PM Cultivation Process: Simgae™ Algal Biomass Production System

Simgae™, which stands for "simple" and "algae", is advertised as offering a low cost and simple approach to growing algae at large scales, thereby overcoming the significant capital costs required for the construction and maintenance of any large-scale cultivation system. Shown below in figure 8 is a conceptual layout of the Simgae™ system.[43] While the general layout of the Simgae™ system will be adopted for this project, key modifications will be introduced to the process that is schematically presented in figure 8 on the following page.[43] As shown, the algae cells are suspended in tubes in a mixture of fresh water, nutrients and dissolved carbon dioxide. The algae fluid (algae cells plus surrounding PM medium) travels through the reactor tubing system and is collected at the end of the tubes. Once at a higher algae cell concentration, the algae fluid enters a common outlet line where it is transported to a secondary location, such as a harvest sump for collection or storage. Key to the setup of this system is the presence of a PM media storage tank on each field. CO_2-rich flue gas (32wt-%) and algae nutrients will be pumped and mixed with the fresh water in the PM medium storage tank before the resulting media is pumped into the tubes. This eliminates the need for an expensive CO_2-injection system and the need to dilute the flue gas with dry air before it is injected into the tubes. The direct injection of air into the tubes could displace the algae fluid and causing cells to settle to the tube bottom, so bubbling CO_2-rich flue gas through the PM medium is a key design element.

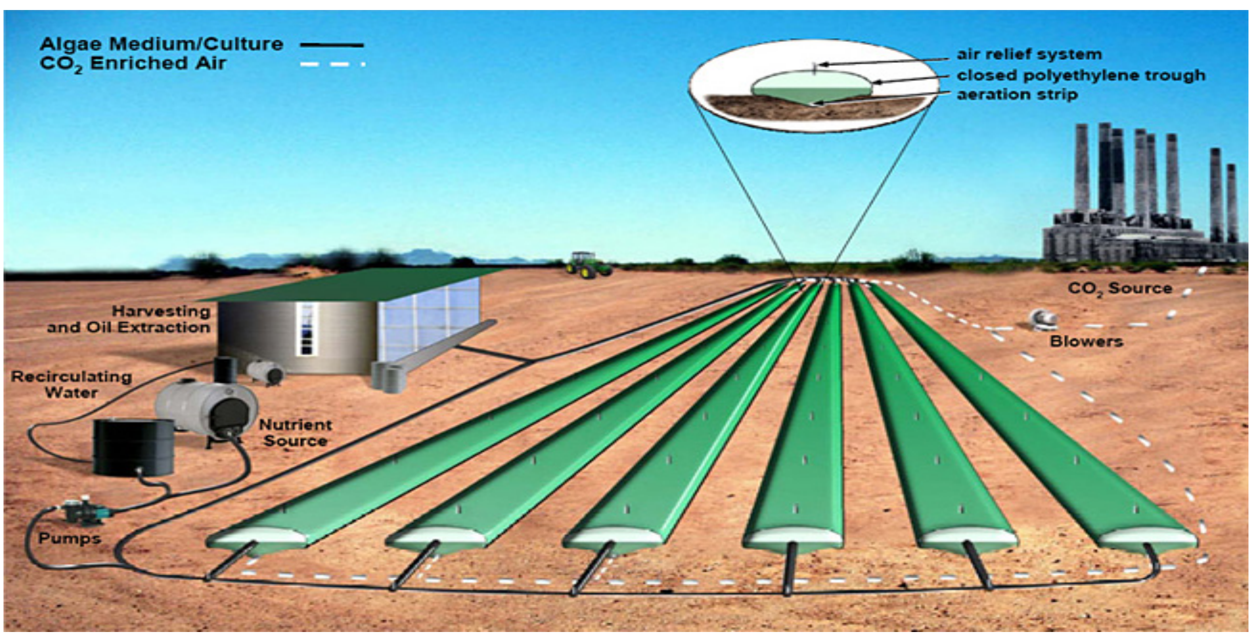


Figure 8: A conceptual layout of the DEC Simgae™ Algal Biomass

The Simgae™ cultivation system is only relevant to the photosynthesis model stage (PM) of the cultivation process. In the cultivation model proposed for this project, algae fluid collected will be transported to clarifiers, where the PM medium is separated from the algae cells. A significant fraction (90%) of the separated PM medium will be recycled and returned to the PM medium storage tank, while 5% of the recovered medium will be discharged. A periodic injection of fresh PM medium is necessary to maintain the concentration of solid particulates in the PM medium at a minimum, and to replace the purged fluid. The separated algae cells are sent to fermentation tanks where they are re-suspended in FM medium to complete the second phase of the cultivation process, the fermentation model stage (FM), which is discussed in more detail in section E, "Proposed FM Cultivation Process: Fermentation Tanks".

Advantages of Simgae™

While the Simgae™ system is neither an open nor a closed system, it has the simplicity and scalability of an open system, and permits the same level of control and accelerated growth as a closed system. The system's primary attraction is its simplicity; DEC states it is a "farmer's solution" to the problem of large-scale algae production. Conveniently, the reactor tubing used is similar to that used in conventional drip irrigation setups, and can be laid out in field troughs created by tractors or other traditional farming equipment. Most of the system's other components, such as pumps and plastic mulch are based on already-existing commercial or traditional agricultural products. DEC emphasizes that this particular aspect of Simgae™ allows for widespread application and adoption of the system, while allowing for ease of installation, operation and maintenance.[43] Scale-up of the system is achieved by simply laying out more reactor tubes, and supplying any additional necessary components.

B. Pressurized Flexible Tubing System for Producing Algae[44]

The framework, layout and specifications for the cultivation fields and reactor tubes are based on the XL Renewable Patent for the Simgae™ technology that is discussed in great detail below. The motivation for adopting this system, as well as its challenges and limitations are clearly outlined. While the cultivation field design is based on that presented in the patent, the framework and layout were not adopted in their entirety; several modifications were introduced to the original design. These modifications were introduced to optimize the productivity of the system, while minimizing the capital costs associated with cultivation.

Motivation

The patent cites the motivation for the invention of a flexible tubing system as the need for a large-scale production system that would allow for a cost effective means of installation, operation and maintenance, relative to production yields. The patent states that while microscopic algae are rich in oil, with a yield per acre that is considerably higher than other oilseed crops such as corn and rapeseed, large-scale operation has yet to materialize.[44] The patent indicates that the following challenges have yet to be properly addressed.

Challenges

The patent sets an upper cap on the algal yield due to the limited wavelength range (400-700 nm) of light energy that is capable of driving photosynthesis.[44] Photosynthetic efficiencies in algal bioreactors are also affected by factors such as respiration requirements during dark periods, the efficiency of solar energy absorption, and other growth conditions, which have resulted in an overall photosynthetic efficiency that is too low to drive economical large-scale production. The patent also states that in order to produce optimal yields, algae cultivation bioreactors need to be coupled with economical sources of CO_2; algae must have easy access to large quantities of CO_2 from flue gas, from nearby thermal power centers.[44] An alternative source of CO_2 found in organic wastes, such as livestock manure and food processing wastes. The patents therefore states that to ensure high yields, a cultivation center must be located near a dairy farm, thermal power plant, or another CO_2 source.

Competitive Advantage

- **Flexibility & Ease of Maintenance**

To highlight its design's competitive advantage, the patent describes two alternative proposals that could compete with the patent's proposed approach. The first is a proposal for a large-scale system that employs a series of pipes that would be elevated over an earthen bed.[44] This proposal however has one major shortcoming, which is its utilization of rigid pipes that are not only expensive to transport, but are also difficult to install and maintain. Another proposal suggests the use of polyethylene tubes made of two layers of 10 mil thick polyethylene coupled to a rigid roller structure.[44] The tubes would lie between two sets of rigid guardrails, and rollers would traverse the tubes to peristaltically move the algae through the tubes. The patent however states that both proposals suffer from the same shortcoming: they both require an external rigid structure of framework to operate the system, which would not only increase capital costs, but would also increase the difficulty in maintaining and erecting the algae cultivation system.

- **Scalability**

The algae-cultivation apparatus proposed by the patent aims to provide a large-scale algae production system for algae-based biofuels that is simple to maintain, and has a lower capital cost than elevated rigid piping or other existing systems. Strong emphasis is placed on the *scalability* of the system: to achieve large-scale algae production, cultivation fields may be simply aligned in parallel, each with its own storage tanks for the supply of FM and PM media, clarifiers for the separation of the media, and fermentation tanks for the heterotrophic growth of the algae cells.[44]

C. Layout & Design of the Cultivation Fields

Reactor Bed Tubes

The algae-cultivation apparatus described would be comprised of reactor tubes made from flexible and water-impermeable material with UV inhibitors. The reactor tubes have a thickness of 6-15 mil, and are arranged in parallel. Examples of tubing material include but are not limited to phthalated polyvinyl chloride, polypropylene and polystyrene, although clear low-density polyethylene is preferred. Polyethylene is not only economical, but it is a very commonly used material. It also allows for efficient heat transfer when the exterior temperature is significantly different from that in the tubing; polyethyelene has a heat transfer coefficient of 0.42 – 0.51 W $m^{-1}K^{-1}$, whereas polyester and polypropylene have a heat transfer coefficient of 0.03 W $m^{-1}K^{-1}$ and 0.1 - 0.22 $Wm^{-1}K^{-1}$, respectively.[45] The tubes are described as lying flat unless pressurized, as shown in figure 9 below, under an operating pressure of 5-20 psi.[45] These figures are from the patent cited above.[44] A circulation pump will be used to propel the algae through the reactor tubing, thereby keeping the reactor tubing pressurized and stationary. This implies that no rigid stationary support structure is required. Although the reactor tubing may be any shape that allows for photosynthesis, a circular geometry is preferred, as it will allow the reactor tubing to be laid without regard for its orientation. A cross sectional diameter no greater than 12 inches is preferred, and a wall thickness of 10 mil is sufficient to make the reactor tubing durable and long lasting. Shown in figure 10 is a cross-sectional schematic of a reactor bed.[45]

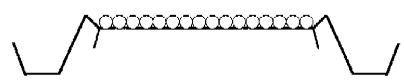


Figure 9: A perspective view of the length of a deflated (L) and a pressurized reactor tube (R).

Figure 10: a cross-section schematic of a reactor bed

For the purposes of this project, the tubing will have a 12-inch diameter, a length of 313 feet, and a 10 mil wall thickness. The patent recommends keeping the tubes partially submerged in the earth, thereby allowing for further insulation from temperature and weather fluctuations. While a 6-inch diameter, as recommended by the patent, is small enough to keep the tubing flexible when pressurized, to maintain the same suggested volume of 245 ft^3, an excessive length of 1,250 ft. would be required. Maintaining control of the algal culture inside a tube this long, with such a small diameter is very difficult, not to mention the fact that with the algae moving at very small flow velocities, plugging of the tubes becomes a concern. Using tubes with a large diameter also allows for sufficient volume in a single line of reactor tubing so that turbulence induced by the flow of fresh water circulates the algae from the bottom to the top of the tubing, and back. The increase in diameter from 6 to 12 inches also results in a 42% decrease in the cost per field. Based on this significant reduction in capital costs, and the aforementioned benefits of utilizing tubes of larger diameters, a decision was made to design the fields based on 12-inch diameter cultivation tubes. However, it is important to note that while the land area required per field decreases as the diameter increases, the pumping power requirement increases as both the flow velocity and volumetric flow rates increase. Please refer to Section B of the Appendix on page 274 for more a more detailed analysis of the reactor tube sizing calculations.

Based on the results discussed above, and the calculations shown in the appendix, the proposed cultivation system will be comprised of a series of fields, each with the following parameters: a gross area of 23.2 acres (383 ft. x 2640 ft.), a bed area of 19.3 acres (100 x 383 ft. x 2200 ft.), a net reactor area of 11.5 acres (100 x16 ft. x 313 ft.), and a total volume capacity of 393,000 ft^3. Each field will contain 100 reactor beds, each of which will have 16 tubes, for a total of 1600 tubes per field.

Modifications to the Patent's Original Design

For the purposes of this project, most of the patent's design specifications remained unchanged, with the following important exceptions: the reactor area relative to the total field area was increased by doubling the number of reactor tubes in each reactor bed from 8 to 16. The tube diameter was changed to 12 inches, and the length reduced to 313 feet. This results in a gross area of 24 acres, a bed area of 20 acres, and a net reactor area of 12 acres.

Another modification made to the patent's specifications involves the algae *residence time*, which is the time required for the algae to make one pass through the tubes. While the patent recommends a residence time of 3.5 hours, due to the cultivation model proposed for the project, the photosynthesis-fermentation model (PFM), the algae must have a longer dwell time. The PFM cultivation model proposed for this project recommends a photosynthesis model period of 120 hours. However, when exposed to prolonged periods of high light intensities, *Chlorella*'s lack of tolerance to such conditions may result in bleaching as discussed previously. Given that the algal cultivation site is located in Texas where temperatures in the summer may reach 100°F, a decision was made to decrease the algae fluid's residence time by 50%, to 60 hours.[40] Nonetheless, it is believed that the induction of bleaching is dark-reversible. [46] Therefore, even if some algae do undergo bleaching during the day, this damage can be reversed during the nighttime hours. Moreover, the residence time of the algae is not a fixed value, and can be modified according to the season. It may be decreased during the summer when there are periods of high light intensity, and increased during the winter. The residence time is controlled by the volumetric flow rate of the algae fluid and can be modified by increasing or decreasing the pump discharge rate.

Shown below in figure 11 is the top view schematic illustration of a single field, with arrows that show algae fluid flow direction as obtained from the patent, “Pressurized Flexible Tubing System for Producing Algae.” [44] As shown and described in the patent, the input line, 40, is flooded with new fertility water in which the algae cells are suspended from the fertility water source, unit 51. This medium is then distributed among the tubes in each reactor bed. For a field comprised of 100 reactor beds, each with 1-ft diameter 16 tubes, pumping a fluid across 1,600 feet will result in an appreciable pressure drop. The development of a pressure gradient will result in an unequal distribution of fluid down the tubes. Fluid will flow down the first tube at a volumetric flow rate greater than that for the 1,600th tube, a condition that is undesirable as a consistent fluid flow rate is specified for each tube.

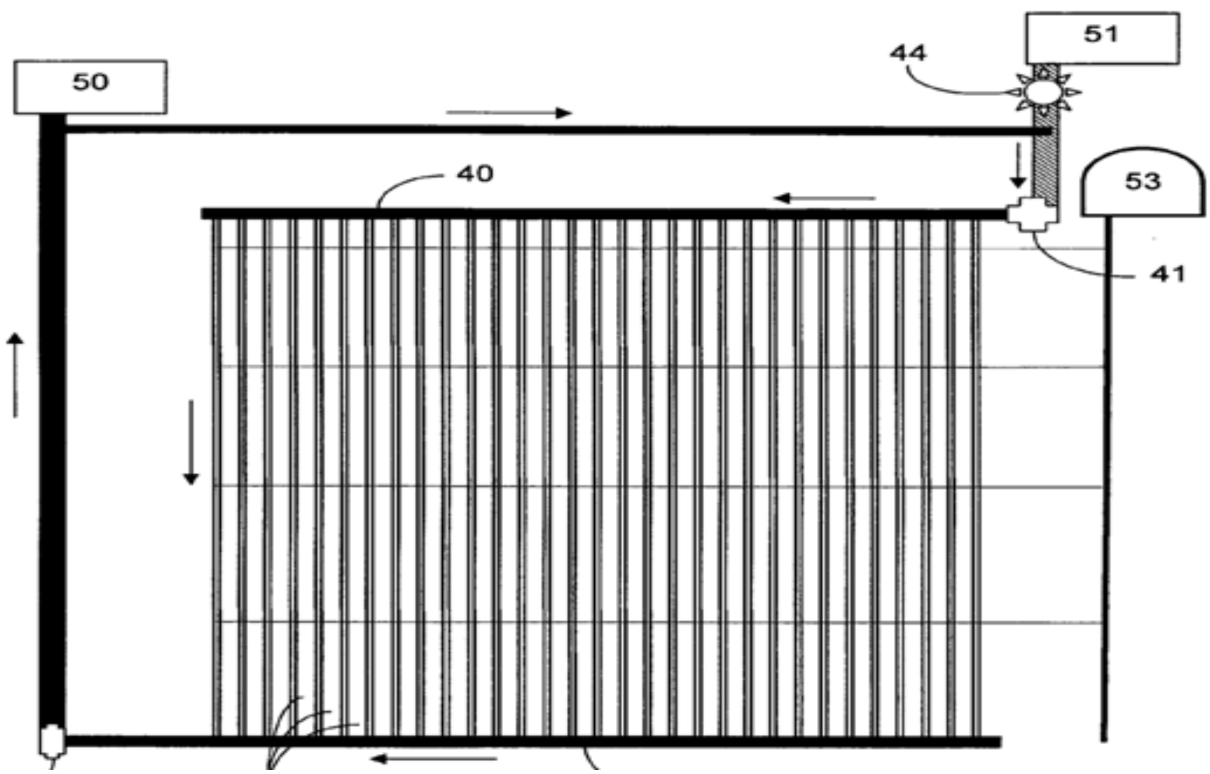


Figure 11: top view schematic illustration of a single field with arrows showing algae fluid flow direction as obtained from the patent, “Pressurized Flexible Tubing System for Producing Algae”

A more suitable and realistic arrangement is therefore shown in figure 12. In a crucial departure from the patent, unit 51, or the PM medium storage tank, is arranged such that a common inlet line is branched at three different levels. While this does not entirely eliminate the formation of a pressure gradient, the parallel arrangement of the branches does help direct the fluid towards the tubes with a minimal pressure difference between points located on the same horizontal line; the pressure at point L2 is roughly the same across the horizontal, and this should in turn ensure that the PM medium flows down all the tubes at very similar flow rates. The branching structure will require some rigidity: 1.5-ft rigid PVC pipes that can tolerate an operating pressure up to 50 psi will be used instead of the flexible clear polyethylene tubes, which by comparison can only tolerate an operating pressure of 14.7-20 psi.[44] The PVC pipes used need not be clear, as the dwell time required for photosynthesis has already been accounted for in the length of the clear polyethylene tubes.

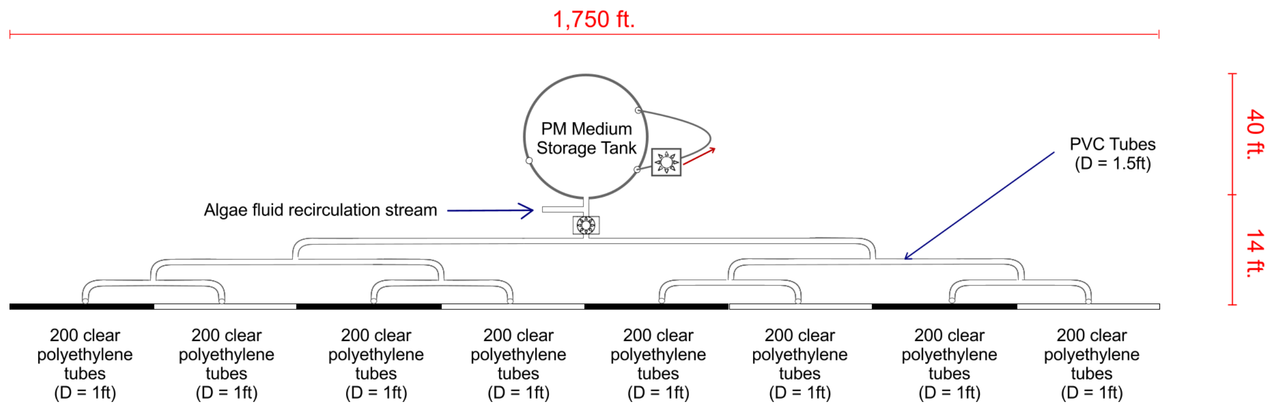


Figure 12:A top view schematic illustration of the preferred layout and arrangement of the PM medium storage tank and the common inlet line. Each of the 1.5-ft diameter final branch feeds into 200 tubes, each of which has a 1-ft diameter. The parallel arrangement of the branches directs the fluid towards the tubes with a minimal pressure difference between points located on the same horizontal line.

Finally, the layout of the field will be modified such that the algae fluid will flow through the tubes at an uphill incline at a 4º angle. This implies that for 313-ft long tubes, the end point of the tubes will be an elevation of 20 feet above ground. Allowing the water to be pumped into a system where the tubes are at an upward incline will ensure that the tubes remain full of algae fluid and without voids, which could develop should the algae be allowed to flow downhill. It will also keep the algae fluid more agitated than the current flat setting; media from the water storage tank will be pumped against the fluid's preferred direction of motion, which would be downhill due to gravity. This will result in intermixing between the rising and down-coming zones[47]. This setup is obtained by stacking more earth on one end of the field than the other. A ramp could be placed underneath the earth to ensure structural stability. Shown in figure 13 is a side view of a single field at an incline of 3 º - 5º angle.

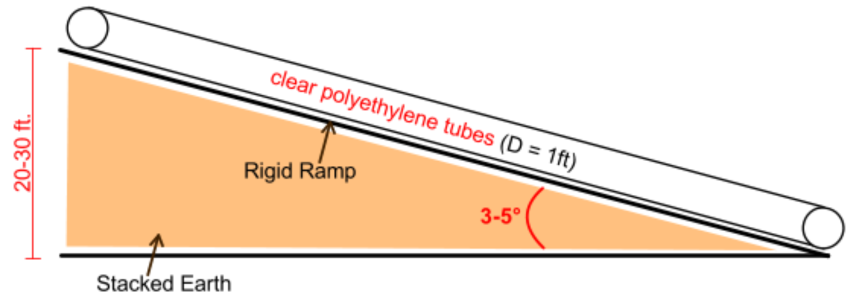


Figure 13: A side view schematic of a single field at an incline of 3 - 5º angle. For tubes that are 313 feet long, this will result in an elevation of 16-30 feet at the opposite end of the tubes.

Recirculation and Harvesting

The patent lays out the possibility of passing the algae fluid through a re-circulation line back to the reactor tubing, such that only a portion of the algae fluid volume is harvested in order to lower the algae content to a pre-determined level. During the harvest cycle, neither the circulation pump, nor the re-circulation line will be used, as the concentrated algae fluid flows out of the output valve, and the input valve is instead flooded with new PM media (fresh water and nutrients). The purpose of re-circulating the algae is to ensure that the algae is only harvested once it has reached a pre-determined concentration, however, no recommendation is given in the patent as to the fraction of algae fluid that is re-circulated. It is implied that this parameter varies from one cultivation system to the other. Shown in figure 14 is a top view schematic illustration of the reactor tubing in a field in circulation mode and another in harvest mode, with arrows showing the direction of flow. For the purposes of this project, re-circulation of the algae fluid was introduced in order to increase the flow velocity of the algae in the tubes to maintain transitional flow conditions. Given the desired flow velocity and a residence time of 60 hours, the recirculation rate was effectively determined.

From a fluid dynamic perspective, algae cells suspended in water can be viewed as discrete, non-flocculating particles. Therefore, to keep the algae fluid under transitional conditions, the

Reynolds number, N_{RE}, must be greater than 1 (N_{RE}> 1). The corresponding flow velocity was determined as follows:

$$v = \frac{(N_{RE} \times \mu)}{(d \times e)} \qquad EQN.1$$

Where,

μ = dynamic viscosity (1.E-3 $N.s.m^{-2}$)

d = diameter of the algae cells ~ 500µm

e = density of the surrounding fluid ~ 1,000 kg m^{-3}

Using these values, the minimum velocity required to achieve a N_{RE} of 1 was calculated to be 0.002 ms^{-1}, or 0.00656 ft s^{-1}. However, assuming the algae fluid makes one pass through the tubes, then for a residence time of 60 hours, and a tube length of 313 feet, the flow velocity could only be a maximum of 0.0014 ft s^{-1}. In order to ensure that the minimum fluid velocity of 0.00656 $ft.s^{-1}$ is achieved, the algae would must be re-circulated approximately 5 times per residence time, or every 12 hours. A recycle rate of 5 times per residence time will allow for a fluid velocity of 0.0072 ft s^{-1}.

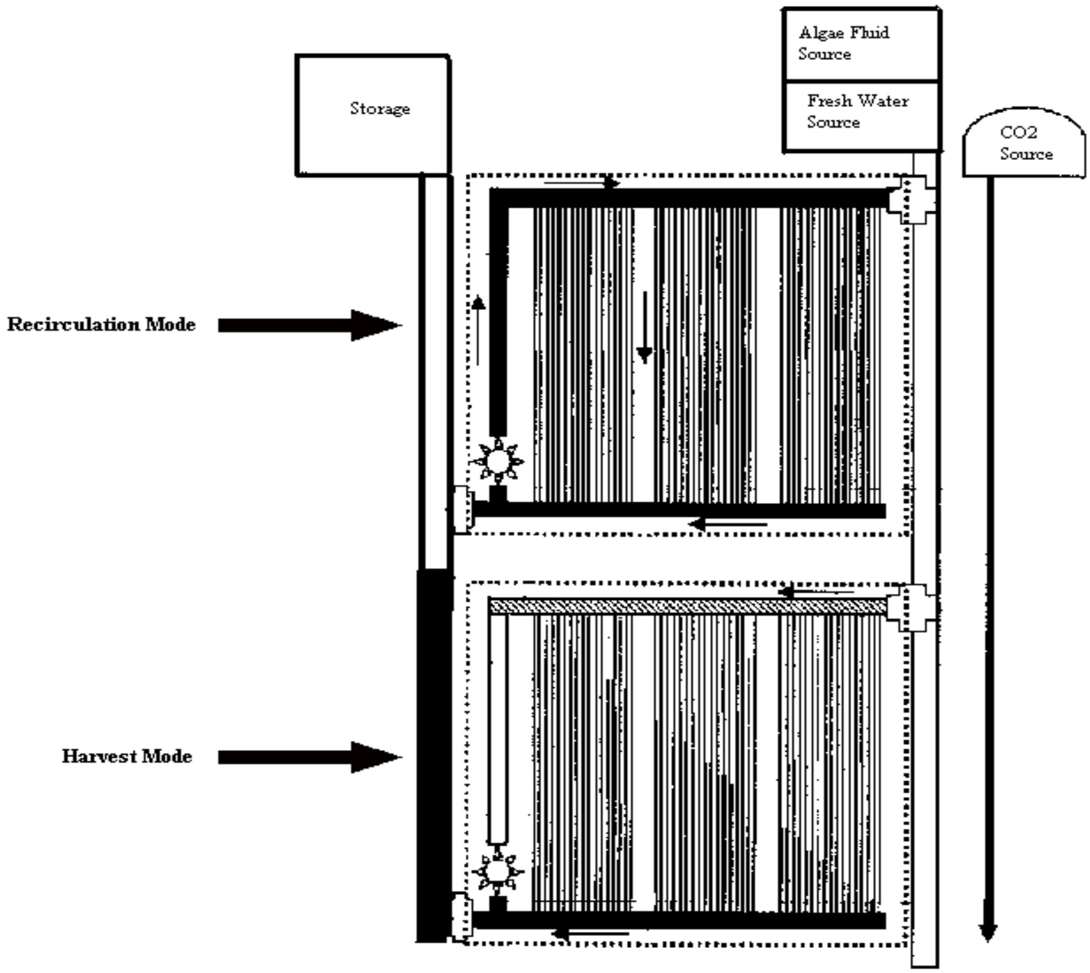


Figure 14: A top view schematic illustration of the reactor tubing in a field in circulation mode and another in harvest mode, with arrows showing the direction of algae fluid flow.

Pumps & Control Valves

Pumps will be needed to initiate or propel the movement of either PM and FM media or algae fluid through the system from sources to sinks. Centrifugal pumps will be used to circulate any fluid that is either free of algae cells or has a very low concentration of algae cells, such as the algae fluid leaving the Simgae^{TM} tubes. Using centrifugal pumps to pump algae fluids that have a high cell density will most likely cause the pumps to clog. Moreover, the passage of high cell density algae cells through a centrifugal pump may result in shearing of the cells. Centrifugal pumps will also be used to circulate the PM and FM media in their respective storage tanks; the use of stirrers in tanks that are roughly 45,000 ft^3 is inadequate to ensure proper mixing of the dissolved gases and nutrients.

A total of 8 pumps will be installed per field, however, only 4 of those are directly relevant to the PM stage. In figure 15 below, a top-view schematic of a single field is shown with all the pump locations clearly marked. In addition to the location of the pumps, the volumetric flow rates, as well as the required pump heads are shown. Two centrifugal pumps will be used to circulate the PM and FM media within the PM and FM storage tanks, respectively. The volumetric flow rate of media circulating in and out of these tanks must be such that the entire contents of the storage tank are turned over at least once a day. The residence time of the PM and FM in each of those tanks is roughly 6 hours.

A centrifugal pump will be placed at the entrance of the common inlet tube to initiate the movement of the PM medium from the PM medium storage tank into the tubes. A pump will not be required to re-circulate the algae fluid in the reverse direction, as the algae fluid will flow downwards by gravity. The pressure supplied by the pump however has to be sufficient to allow the algae fluid to re-circulate 5 times through the tubes. A pump will not be required to pump the algae fluid to the PM clarifier. Since end point of the tubes lie at an elevation of roughly 20 feet, whereas the PM clarifier has a height of roughly 13 feet; the algae fluid will flow down by gravity. A centrifugal pump will however be required to recycle and pump the recovered PM medium from the clarifier back to the top of the PM medium storage tank. A centrifugal pump will be also required to pump the separated algae slurry from the PM clarifier to the top of the FM fermentation tanks.

Detailed sample calculations regarding the pump power requirement of each of the pumps can be found in the Sample Centrifugal Pump Calculation in Section E of the Appendix on page 366. Control valves will be placed after each pump, except those used for mixing the PM and FM media within their respective storage tanks.

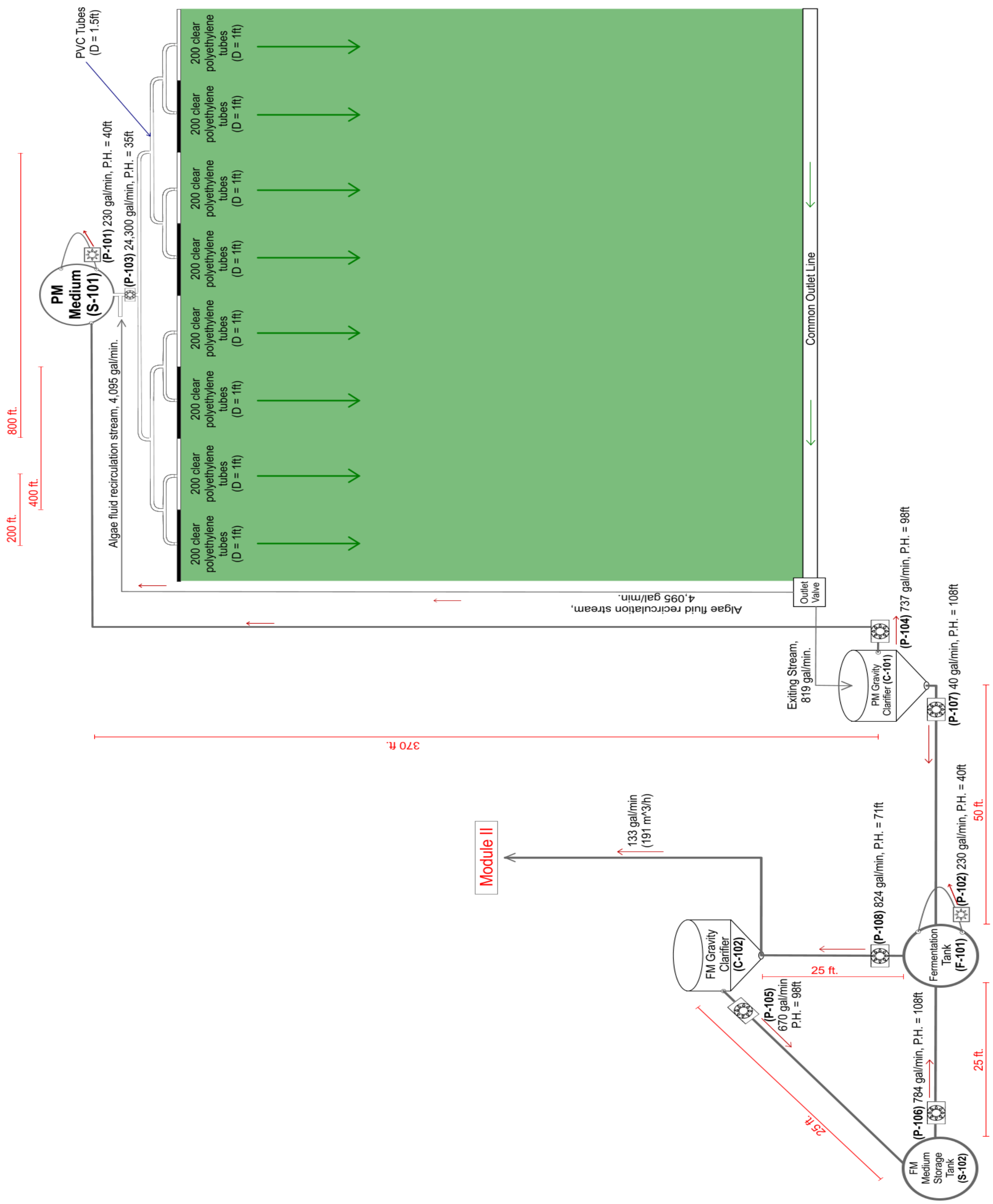


Figure 15: A top-view schematic of a single field with all the pump locations clearly marked. In addition to the location of the pumps, the volumetric flow rates, as well as the required pump heads are shown. 8 pumps in total will be required per field, two of which will be used to circulate the PM and FM medium in and out of their respective tanks for mixing.

Maintenance

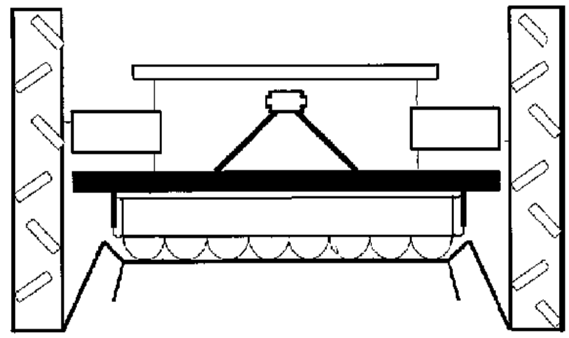


Figure 16: A rear view of a tape roller tractor aligned over a reactor bed.

The tubing system may be maintained by tractors that can straddle each bioreactor and travel up and down the rows of tubing for repair of leaks, replacement of reactor tubes, and overall periodic maintenance. In order to control the deposition of biofilm on the walls of the reactor tubes, mechanical agitation may be utilized. Rollers, mounted on a toolbar, are pulled through the bioreactor field by a tractor as shown in figure 16, adapted from the patent discussed above.[44] It is stated within the patent materials that the rollers create a venturi effect by periodically applying pressure to and partially collapsing the tube.[44] Despite the potential for frequent tractor maintenance, it is estimated the the cultivation system will be shut down for approximately one month each year for general maintenance.

Shown in figure 17 is a top view schematic of a single cultivation field with the relevant parameters and dimensions indicated. This schematic is relevant only to the PM stage. The arrows show the algae fluid flow direction. Figure 18 is a top view schematic of a single reactor bed, with all the relevant parameters indicated and the 16 tubes clearly shown. Table III is a summary of the important parameters relevant to a tube, reactor bed, or field.

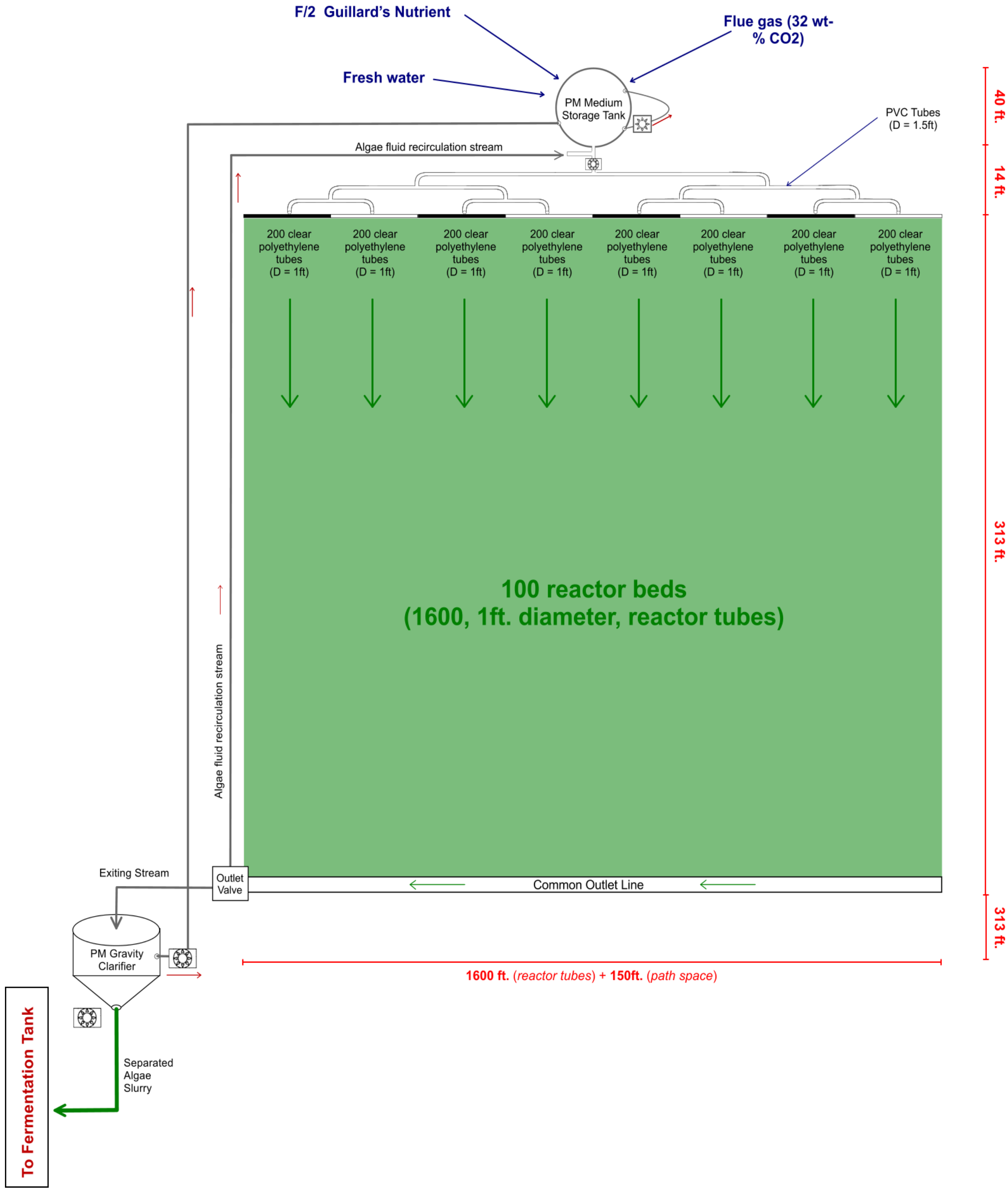


Figure 17: A top view schematic of one of the fields with the parameters relevant to the PM cultivation stage indicated. A single field has a total area 24 acres with 20 acres of reactor bed and an effective reactor area of 12 acres. Black arrows represent the flow of the salt water source. Red arrows model the flow path of algae. Each field will contain 100 reactor beds, each of which will have 16 tubes, for a total of 1600 tubes per field.

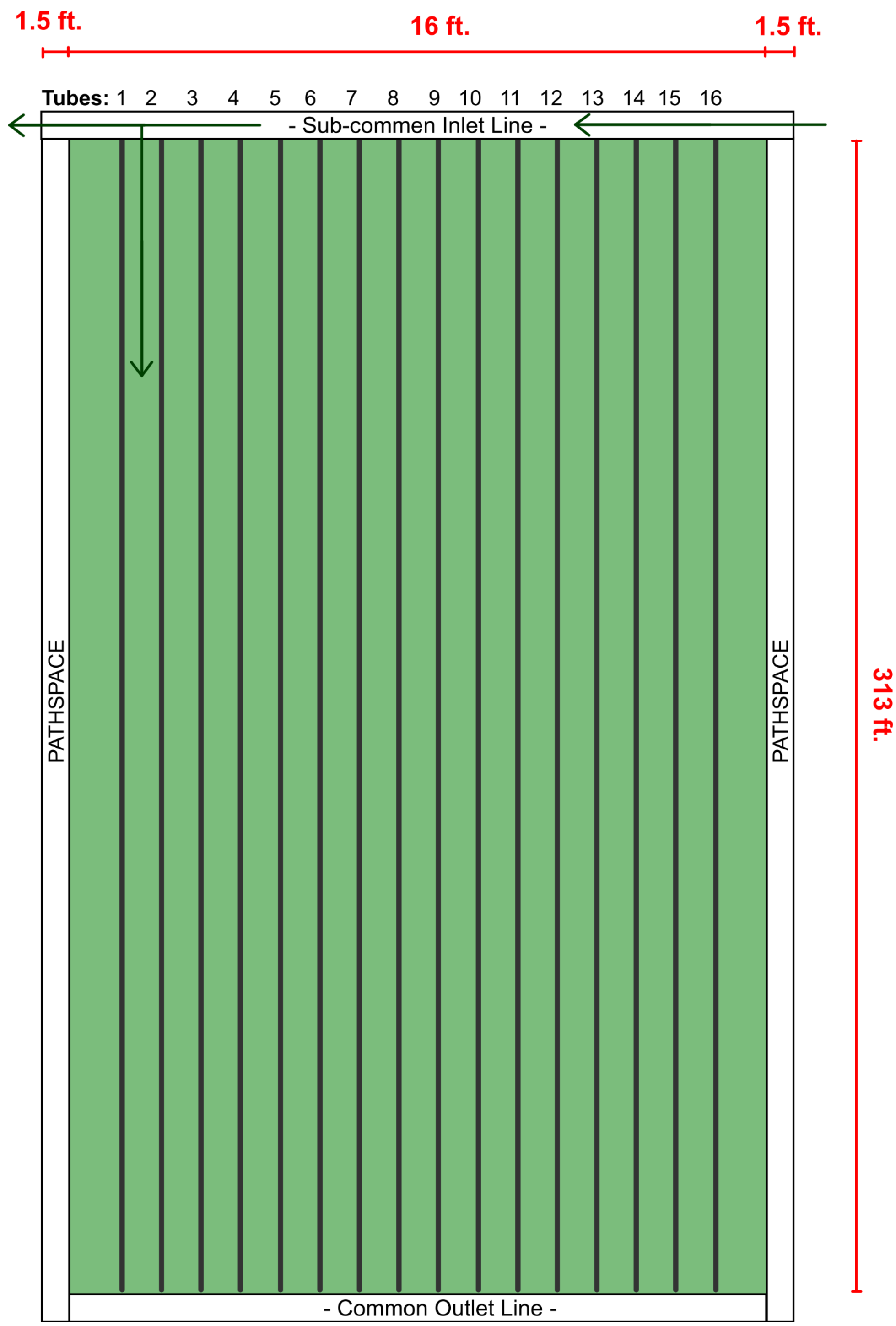


Figure 18: A top view schematic of a single reactor bed, with all the relevant parameters indicated and the 16 polyethylene reactor tubes clearly shown. The reactor tubes will be 313 ft long, with a 12 inch-diameter. The arrows show the algae fluid flow direction during the production stage.

Table III: Summarizes the important parameters relevant to a tube, reactor bed, or field.

SINGLE FIELD		
Gross	24 acres	383 ft X 2640 ft
Reactor Beds	20 acres	383 ft X 2200 ft
Net Reactor Area	12 acres	313 ft X 1600 ft
Space for Paths	152 ft	1.5 feet per pathway
Reactor Beds per field	100	--
Total Volume	393,000 ft^3	11100 m^3
Total Volumetric Flow Rate	1.8 ft^3 s^{-1}	52 L s^{-1}
Number of Pumps	8	--
Cost per Field	$ 1.16E+06	--
NET REACTOR AREA		
Number of Tubes	16	--
Length	313 ft	95 m
Width	16 ft	4.9 m
Area	0.12 acres	466 m^2
Total Volume	3930 ft^3	111000 L
REACTOR TUBING		
Tube Diameter	1 ft	0.308 m
Tube Length	313 ft	95 m
Dwell Time	60 h	2.5 days
Total Volume	245 ft^3	6.95 m^3
Re-circulation Rate	1 pass per 12 hrs	--
Flow Velocity	26 ft h^{-1}	7.9 m h^{-1}

D. Concerns Regarding the Simgae™ System

Biofilm Deposition

Concerns about the accumulation of biofilm on the walls of the reactor tubing have also been addressed in the patent. The deposition of algae on the reactor tubing not only reduces its clarity, thereby reducing light penetration, but also constricts the flow of algae through the system. The patent however states that a loss in system efficiency may be anticipated and overcome by the use of longer lengths of reactor tubing. A reactor tubing length of 300 feet is sufficient to ensure optimum system performance over a 5-year operating period according to the patent materials.[44] As mentioned above in the maintenance section, mechanical agitation could be utilized in order to reduce the growth of biofilm on the reactor tubing. This is cited as the most effective and economical method of controlling the deposition of biofilm. Carlson et al. (2010) recommend removing the biofilm by chemical means on a periodic basis, but complete biofilm removal is not desired; some biofilm accumulation protects polyethylene tubing from light of high intensity.[6]

Polyethylene Tubing Degradation

Concerns regarding the degradation of the polyethylene tubing were also addressed in the patent; polyethylene may degrade relatively rapidly under conditions of high temperatures and sunlight intensity. The reactor tubing will therefore not only made of UV light absorbent molecules, but will also be kept partially buried in the earth to limit degradation.

Temperature Control

Maintaining the temperature of the algae fluid within the optimum range for growth is crucial during the autotrophic phase. The algae strain proposed for this project, *Chlorella protothecoides,* grows well at 28±1°C.[9] As previously discussed, the reactor tubing may partially covered in earth as a key form of insulation from temperature and weather fluctuations. Furthermore, plastic mulch, a thermal barrier common in the agriculture industry, is distributed around the reactor tubing to further protect against temperature fluctuations. Further temperature control involves the injection of PM media at elevated temperatures during the wintertime and the injection of cooled PM media during the summer time. This strategy for temperature control requires auxiliary heating and cooling units.

Very Dilute Exiting Algae Fluid Stream

Despite achieving a residence time of 60 hours, the algae stream exiting the PM cultivation stage is very dilute, with an algae concentration no greater than 2.89 g L^{-1}. This is an inherent inefficiency of the system that results from the need for large quantities of fluid (fresh water and nutrients) for the cultivation of algae at very low growth rates (0.286 g $L^{-1}d^{-1}$) and very low concentrations. Increasing the residence time, or using algae strains with greater growth rates would lead to greater exiting algae fluid concentrations. However, for the purposes of this project, neither option was implemented. Instead, according to the cultivation model proposed, the photosynthesis-fermentation model (PFM), once the algae makes one pass through the Simgae™ System, it is transported to fermentation tanks, where it resides for another 60 hours. There, the algae will undergo heterotrophic growth in the presence of organic carbon, such as glucose or corn powder hydrolysate (CPH), which significant increases the algae fluid's exiting concentration. Please refer to section E for more details regarding heterotrophic growth.

Photoinhibition

Scale-up of the proposed cultivation system can be accomplished either by increasing the length or diameter of the tubes. Although increasing the tube diameter allows for better control of the cultivation conditions inside the tubes, it also increases the potential for photoinhibition. Ugwa et al. (2007) argue that increasing the tube diameter decreases the surface area-to-volume ratio and suggests that algae in the bottom portion of the tube may not receive enough light for cell growth.[47] To overcome the disadvantages of photoinhibition, an adequate mixing system that allows for efficient light distribution among the algae cells is required. This is particularly important in closed cultivation reactors where neither the supply of nutrients nor carbon dioxide are limiting for growth and biomass yield, but light energy is.[48] Alterations in the exposure of algae to areas of light and darkness may significantly increase photosynthetic efficiency.[49]

E. Proposed FM Cultivation Process: Fermentation Tanks

As mentioned previously, the overall configuration of the algae cultivation process can be divided into two stages: an autotrophic photosynthetic model stage (PM) and a heterotrophic fermentation stage model (FM). Once the algae cells have been separated from the surrounding PM media in the PM gravity clarifier, the separated algae slurry is transported to fermentation tanks, whereas the recovered PM medium is recycled back to the PM medium storage tank. This marks the beginning of the FM stage. In the fermentation tank, the algae will be re-suspended in glycerol-rich FM medium where it will undergo heterotrophic growth to maximize cell density and allow for enhanced lipid accumulation.[27] The algae are then sent to the FM gravity clarifier, where the FM medium is recovered and recycled to the FM medium storage tank. The separated algae slurry is sent to a storage tank where it is then transported to a lipid extraction unit in Module II.

Competitive Advantage

The use of fermenters for heterotrophic growth of algae has several advantages. Harvesting of algae cells from high cell density-cultures is much more easily achieved in fermenters than it is in other closed configurations. Operating conditions in the fermenter, such as temperature, pressure and pH can be controlled and set as desired. More importantly, fermenters are common pieces of equipment in the chemical industry and do not require sophisticated design or materials; maintenance and operation of fermentation tanks are well understood. Moreover, maintenance and construction costs are relatively low in comparison to more closed sophisticated reactors, such as photobioreactors. Although fermentation of algae is not proposed here as a substitution for autotrophic growth, but rather as a subsequent and complementary stage in the process of algae cultivation, shown in table IV is a comparison of the features of photobioreactors and fermenters.[49]

Table IV[49]: A comparison of the relevant features of photobioreactors and fermenters

Feature	Photobioreactor (PBR)	Fermenter
Energy Source	Light	Organic carbon
Cell density/dry weight	Low	High
Limiting factor for growth	Light	Oxygen
Harvestability	Dilute, more difficult	Denser, less difficult
Control of parameters	High	High
Sterility	Usually sanitized	Can be completely sterilized
Availability of vessels	Often made in-house	Commercially available
Technology base	Relatively new	Centuries old
Construction costs	High per-unit volume	Low per-unit volume
Operating costs	High per-kg biomass	Low per-unit biomass
Applicability to algae	Photosynthetic algae	Heterotrophic algae

F. Design of the Fermentation Tanks

Organic Carbon Source

The most widely used source of organic carbon in fermentation is glucose, although other sources such as acetate or corn powder hydrolysate (CPH) have also been used (Miao and Wu, 2004). Interestingly, Grady et al. (2010) examined the use of glycerol alone and in combination with glucose for the cultivation of heterotrophic *C. protothecoides* cultures.[105] In a complementary study, Heredia-Arroyo et al. (2010) recommend using a glycerol concentration of 10g L^{-1}, which was shown to result in similarly high biomass productivity and lipid content as for algae grown in a glucose-rich media.[50] Therefore, given that crude glycerol is a by-product of transesterification, a candidate for the Module III lipid-processing design, and has shown to yield the same favorable results as glucose, crude glycerol is proposed as the substrate for heterotrophic growth. Although using glycerol as a feed in the fermentation tanks will cut down on the profits made via byproduct sales, this loss is balanced by a significant reduction in the costs of raw materials for algae cultivation. In fact, Xiong et al. (1992) emphasize that the use of glucose in the fermentation media contributes to approximately 80% of the total medium cost.[9]

Stirring Speed

The stirring speed and circulation in the fermenter must be maintained at adequate levels to allow for efficient heat and mass transfer, and must be adjusted as the concentration or volume of the contents of the fermenter change. Although adequate mixing is generally achieved using impellers and baffles, the choice of mixing apparatus is often highly dependent on the sensitivity of the organisms to being sheared.[51] Bader (1986) points out that since shear scales with the mixer tip speed, in the case of shear sensitive algae, a reduction in tip speed is required. In order to compensate for the consequential reduction in mixing, Bader and Clark (1986, 1997) recommend using several impellers. Mixing can also be achieved by the "airlift" principle, although this approach requires a significant air flow rate sufficient to ensure adequate mass and heat transfer and may not suitable for achieving mixing in the case of algae with low specific gravities as they will float.[52] Regardless of which mixing approach is used, maintaining a well – mixed solution may be complicated by the change in the specific gravity of the algae cells during fermentation.[48]

Oxygen Supply

During fermentation, supplying a sufficient amount of oxygen is the single most critical factor to achieving a rapid growth rate and a high cell density.[53] Oxygen is usually supplied in the form of compressed air that has an oxygen concentration of 21% by volume. The concentration of dissolved oxygen in air-saturated water is referred to as 100% dissolved oxygen, and is usually on the order of 250 μM.[54] Since oxygen has a low solubility in aqueous media, 0.038 $g_{gas}kg_{water}^{-1}$

at 28ºC, care must be taken to not only supply sufficient amounts of the gas, but also to ensure adequate mixing, as recommended by Xiong et al. (1992), who suggest that the stirring speed in the fermenter be adjusted to control the O_2 concentration.[55] Matters are made more complicated by the fact that during fermentation and cell growth, the dissolve ocygen levels are expected to change and must therefore be empirically determined as a function of airflow rate, bubbles residence time, bubble size and temperature.[56]

Sterilization of the Fermenters and Culture Medium

Given that carbon is an attractive energy source for fungi and bacteria in addition to algae, sterilization of the fermenter and culture medium before fermentation is critical. Although the fermentation tank may be sterilized using a variety of methods, which include heat, irradiation or chemicals, sterilization by steam is the most common, non-toxic and cost effective method.[53] Quesnel (1981) states that the culture medium is commonly sterilized using steam, either in the fermentation vessel, or at a secondary location.[53] If sterilization of the medium is carried out in the vessel, longer periods of sterilization are required due to slower heat transfer, and care must be taken not to expose the culture nutrients to high temperatures for prolonged periods of time. However, this method has the advantage of eliminating the need to transport large volumes of culture medium.[53] Alternatively, the culture medium can be sterilized at a secondary location. This option has the advantage of being rapid, and preventing the culture medium from being exposed to high temperatures for prolonged periods of time. However, whether sterilization takes place *in situ*, or even at a secondary location, the effect of high temperatures on the culture medium is still a concern as it may affect the composition and viscosity of the culture medium. The level of sterility achieved is also a concern, and it is recommended that sterility tests be performed in which a sterilized medium is left un-inoculated for a certain period of time to determine whether bacteria or fungi grow.[53] Failure of microorganisms to grow would indicate that the fermenter and the culture medium have been successfully sterilized.

Mode of Operation for Optimum Growth Conditions

Steel tanks will be used for the FM stage of the cultivation process. The system will operate under steady-state conditions, such that the volumetric flow rate in and out of the fermentation tanks will remain unchanged during normal operation; the change in density as a result of increase in the algae concentration is assumed to be negligible. Xiufeng et al., (2007) recommend the following operating conditions: a temperature of 28±1°C, and an initial pH value greater than 6.0.[28] The pH value is maintained through the use of K_2HPO_4 to KH_2PO_4 of 0.86. The flow rate of aseptic air will be set at roughly 52m^3 h^{-1}, and the stirring rate at180 rpm. This will allow the dissolved oxygen (DO) value to remain over 20% air saturation. For the purposes of this project, fermentation tanks with volume capacity of 1,909 m^3 or 67,380 ft^3 will be used. Based on the total volumetric flow rate of the algae, and the volume capacity of the fermentation tanks, the residence time of the algae fluid can be calculated according to:

$$Volume = VolumetricFlowRate \times ResidenceTime$$

For a volumetric flow rate of 191 m^{3}hr^{-1}or 6,745 ft^{3}hr^{-1}per tank, the residence time is calculated to be 10 hours.

Pumps & Control Valves

Pumps will be needed to initiate or propel the movement FM media or algae fluid through the system from sources to sinks. As mentioned above, 4 pumps are required for the FM stage: A centrifugal pump will be required to pump the separated algae slurry from the PM clarifier to the top of the FM fermentation tanks. A centrifugal pump will be required to pump the FM medium from the FM medium storage tank to the top of the fermentation tank. A centrifugal pump will be required to pump the algae fluid from the FM fermentation tank to the top of the FM clarifier and a centrifugal pump will be required to pump the recovered FM medium from the FM clarifier back to the top of the FM storage tank. A displacement pump will finally be required to pump the separated algae cells from the FM clarifier to a storage tank that will then transport the algae cells to Module II for lipid extraction. The location of each of those pumps was shown clearly in figure 15 in section C under “Pumps and Control Valves, on page 46.

G. Additional Concerns Regarding the Fermentation Process

Heat Transfer

The process of metabolizing glycerol in the presence of oxygen releases heat as a byproduct. In order to maintain optimum conditions in the fermentation tanks, adequate heat transfer must occur. Heat may either need to be supplied or removed from the fermentation tanks. Since cooling water, supplied at a temperature of roughly 32ºC (90ºF) is too hot to cool the media to 28ºC, chilled water may be used to remove the heat generated through fermentation.[67] Although this operation may result in a significant addition to the variable costs of cultivation, it is nonetheless necessary to maintain optimum conditions in the fermentation tanks. To prevent heat loss from the fermentation tanks through the reactor walls by conduction, the fermentation tanks can be covered with insulating glass foam, which is not only very cheap, but also very light and add negligible cost to the existing design. Please see Section C of the Appendix on page 274 for additional heat transfer and calculations and the energy requirements for auxiliary cooling.

Alcohol Formation

As noted above, a biomass yield of 0.617 $g/g_{glucose}$ and a lipid yield of 0.298 $g/g_{glucose}$ were reported in the PFM model by Xiong et al., (1992).[9] This suggests that 91.5% of the glucose fed to the system is converted to biomass or lipid. The difference, 8.5% of the glucose fed to the system, is either unutilized or converted to ethanol among other alcohols. Although this was not listed as a concern in Xiong et al. (1992), the formation of alcohol in the fermentation tanks could adversely affect the optimum growth conditions by changing the pH to a less desirable value, or producing byproducts detrimental to algae growth.[9] The adverse effects of ethanol on bacterial growth, for example, have been well studied and documented; the presence of ethanol is known to induce leakage of the plasma membrane.[57] If it is determined that ethanol can adversely affect the growth of algae, measures will have to be taken to separate it from the FM medium, as recycling of the medium will cause accumulation of alcohol in the system.

H. PFM Cultivation: Continuous Requirements

Optimal Conditions for Chlorella protothecoides Cultivation

The optimal conditions for the growth of *C. protothecoides* are based on the experimental conditions under which data supporting the PFM model were collected, as well as other referenced studies which investigate the effects of varying the fermentation substrate on the heterotrophic growth of algae. The calculations shown below are primarily concerned with determining the raw materials, utilities and pieces of equipment required to sustain a PFM algae cultivation system, and will be essential for conducting a thorough economic analysis of the proposed cultivation process.

A General Material Balance (PM, FM)

In order to determine the algae production rate, it is critical to work backwards from the desired biodiesel production rate, which was set at 17,560 bpd, or 223,000 lb hr^{-1}. From this figure, a triglyceride production rate of 223618 $lb_{TG}.hr^{-1}$(101,460 kg_{TG} hr^{-1}), or 5,366,828 $lb_{TG}.day^{-1}$(2,435,040 kg_{TG} day^{-1}) was calcaulted. *Chlorella protothecoides* are considered to have a lipid content of 46 dry wt.%, with 90% of the total lipids in the form of triglycerides (TG). Assuming a lipid extraction efficiency of 90% in Module II, an algae production rate of 272,303 kg hr^{-1} on a dry basis was calculated, as shown below:

$$\frac{101{,}460\,\frac{kgTG}{hr}}{0.9\times0.9\times0.46} = 272{,}303\,\frac{kg}{hr} = 600{,}156\,\frac{lb}{hr}$$

Assuming algae cells have a water content of 90 wt.%, a total algae production rate of 2,450,718 kg_{Algae} hr^{-1}, or 5,403,853 lb_{Algae} hr^{-1} is required. The specific growth rate of *Chlorella protothecoides* growth on pure glycerol was measured by Grady et al. (2010) to be 0.1 hr^{-1}. From this figure, the mass of algae in the fermentation tanks required to ensure the production rate of 2,450,718 kg_{Algae} hr^{-1} mentioned above, the following calculation was performed:

$$\frac{0.1\,g_{algae}}{g_{algae}.hr} = \frac{2{,}450{,}718\,\frac{\mathrm{kg}}{hr}\times1000\,\frac{\mathrm{g}}{\mathrm{kg}}}{(M_{FM})\,kgalgae}$$

This equation is solved for the value of the mass of fermenting algae, M_{FM}, which gives a result of 24,507,180 kg of algae. Xiong et al. (1992) report a maximum achievable cell density of 123 g L^{-1} in the FM stage [2]. For the purposes of this project, the achievable cell density is also assumed to be 123 g L^{-1}. Therefore, from the value of M_{FM} the volume of algae fluid required in the fermentation tanks is calculated to be 204227 m^3, or 7,212,208 ft^3. Similarly, for a cell density of 123 gL^{-1}, the required volumetric production rate is calculated to be 20,423 m^3hr^{-1}, or 721,231 ft^3hr^{-1}.

From the FM stage results above, the PM stage material balance may be written. Based on a recommendation from Dr. Wen K. Shieh at the *University of Pennsylvania*, the algae fluid leaving the Simgae™ tubing is assumed to have a cell density no greater than 0.2 g L^{-1}. C. *protothecoides* cells growing in autotrophic cultures are reported to have a specific growth rate of 0.005 hr^{-1}, which is dramatically lower than that reported for heterotrophic or mixotrophic cultures.[58] The algae production rate of the PM stage required to ensure a subsequent FM algae production rate of 2,450,718 kg_{Algae} hr^{-1} was determined as follows:

$$2{,}450{,}718\,\frac{kg}{hr} \times \left(\frac{\frac{0.2g}{L}}{\frac{123\ g}{L}} \right) = 3{,}985\ \frac{kg}{hr} = 8{,}783\,\frac{lb}{hr}$$

This calculation shown above is a simple dilution calculation. Because the specific growth rate of the algae is very low in the autotrophic phase, the cell density of the algae is assumed to be constant at 0.2 gL^{-1}. Assuming an algae fluid density of 0.2 gL^{-1}, the exiting volumetric flow rate is calculated to be 19,925 m^3hr^{-1}, or 703,104 ft^3. For a residence time of 60 hours, the mass of algae fluid in the tubes is calculated as follows:

$$60\ hr = \frac{M_{PM}}{3{,}985\,\frac{kg}{hr}}$$

This equation is solved for the value of M_{PM}, which is 239,094 kg of algae. For an algae fluid density of 0.2g L^{-1}, the volume of algae required in the Simgae™ tubing is calculated to be 1,195,470 m^3, or 42,217,624 ft^3. Given that each Simgae™ field has a volume capacity of 11,130 m^3, or 393,000 ft^3, the number of fields required to host this volume is calculated to be roughly 107 fields.

The system is assumed to operate under steady state conditions. Although the mass flow rate of the algae entering and leaving the fermentation tank does change, it is assumed to have a negligible effect on the overall flow rate of the streams involved. The volumetric flow rates into and exiting each of the Simgae^{TM} fields is calculated to be 186 m^3hr^{-1} (52 ms^{-1}), or 6,568 ft^3hr^{-1}. As field will be associated with its own fermentation tank, 107 fermentation tanks are required. Based on the total fermentation volume calculated above to be 204227 m^3, or 7,212,208 ft^3, each fermentation tank will have a volume capacity of 1,909 m^3, or 67,380 ft^3. Given a total volumetric flow rate of 20,423 m^3 hr^{-1}, the volumetric flow rates into and exiting each of the fermentation tanks is calculated to be 191 m^3hr^{-1}(53 Ls^{-1}) or 6,745 ft^3hr^{-1}. Shown below in table V is a summary of the values calculated above for ease of inspection.

Table V: A Summary of the relevant design parameters of the PM and FM cultivation stages.

Parameter	**PM Stage**	**FM Stage**
Total Mass Production Rate (kg hr^{-1})	3,985	2,450,718
Total Specific Growth Rate (hr^{-1})	0.005	0.1
Cell Density (gL^{-1})	0.2	123
Total Volumetric Production Rate (m^3hr^{-1})	19,925	20,423
Residence Time (hr)	60	10
Total Volume Capacity per Field/Tank (m^3)	11,130	1,909
Number of Fields / Tanks	107	107
Mass Production Rate per Field/ Tank (kg hr^{-1})	37	22,904
Entrance/Exit Volumetric Flow Rate per Field/ Tank (m^3 hr^{-1})	186	191
Volume Capacity per PM/FM Storage Tank (m^3)	1,253	1,253

Gravity Clarifiers Requirement (PM, FM)

The outlet stream of each field and fermentation tank will be directed towards a clarifier where the PM or FM media will be separated from the algae cells, and recycled to the PM and FM media storage tanks respectively. The design of the clarification system is based on a consideration of the surface overflow rate and the solids loading rate. The overflow rate is assumed to be influenced only by the effluent requirements and the need to provide steady-state and consistent performance. Since the solids separated from the incoming stream are drawn off from the bottom of the tank, they are not accounted for in the calculation of the upward flow velocity.[59] The average overflow rate range recommended for the settling of air-activated sludge is 16-28 $m^3m^2day^{-1}$. The value used in the following calculation is the average of the two extremes: 22 $m^3m^{-2}day^{-1}$. The corresponding recommended depth is 4 meters.[59] Since the exiting volumetric flow rates in both cases are very close 186 m^3hr^{-1} and 191m^3hr^1, one calculation will be performed for both the PM and FM clarifiers, as the difference is assumed to have a negligible effect on the overall design of the clarifier. The exiting flow rate used in the following calculation is 4,500 m^3day^{-1}.

$$\left(\frac{4500\,\frac{m^3}{day}}{20\,\frac{m^3}{m^2.day}}\right) = 225\ m^2 = 2420\ ft^2$$

For a circular clarifier, this settling area corresponds to a diameter of 17 meters, or 56 ft. Therefore, for a height of 4 meters, the volume of the clarifier is calculated to be 900 m^3, or 31,783ft^3. For an exiting volumetric flow rate of 4,500 m^3day^{-1}, the residence time, or settling time in this case is calculated to be 6 hours. Each field will have 2 gravity clarifiers: a PM gravity clarifier and an FM gravity clarifier. As the gravity clarifiers have diameters of 56 ft, transportation of these units is not feasible and they must be fabricated onsite. The extra cost associated with onsite fabrication was taken into account for these items.

Carbon Dioxide (CO_2) Consumption (PM)

CO_2 gas is required for photosynthesis; microalgae consume CO_2 during their growth process, which along with other nutrients, is typically dissolved in the surrounding medium. In order to determine the consumption rate of CO_2 during photosynthesis, the following calculation was carried out, under the assumption that algae are 50% carbon by dry weight[60]. Since a mole of CO_2 has a molecular mass of 44 grams, 12 of which are carbon, the rate of consumption of CO_2 *per field* is calculated as follows:

$$\frac{44\,\frac{gCO_2}{mol}}{12\,\frac{gC}{mol}}\times 0.5\,\frac{gC}{gAlgae} = 1.83\,\frac{gCO_2}{gAlgae}\,perfield$$

Therefore, for an algae production rate of 3,985 $kg.hr^{-1}$, a consumption rate of CO_2*per field* is calculated as shown:

$$\frac{1.83 \frac{gCO_2}{gAlgae} \times 3{,}985 \frac{kgAlgae}{hr} \times \frac{hr}{3600\ s} \times 1000 \frac{g}{kg}}{107\ fields} = 19 \frac{gCO_2}{s} = 0.042\ \frac{lbCO_2}{s} perfield$$

For the purposes of this project, algae are assumed to obtain all their required carbon from CO_2.

Flue Gas Volumetric Flow Rate Requirement (PM)

In order to calculate the required volumetric flow rate of flue gas, it is necessary to know the required CO_2 concentration for autotrophic growth, and to supply a 50% excess of that amount. Given the carbon dioxide consumption rate of 19 g s^{-1}, calculated above, and the volume *per field*, the amount of CO_2 required per 60 hours, and the required CO_2 concentration in the PM medium can be calculated as follows:

$$\frac{19\ \frac{gCO_2}{s} \times 3{,}600 \frac{s}{hr} \times 60\ hr \times \frac{molCO_2}{44\ gCO_2}}{11{,}130\ m^3 \times 1{,}000 \frac{L}{m^3}} = \ 0.0083 \frac{molCO_2}{L} perfield$$

According to Henry's law, the amount of gas dissolved in a given liquid of a certain volume at a constant temperature is directly proportional to the partial pressure of the gas in equilibrium with the liquid.[61] Henry's constant for CO_2 in water at 298°K is given as 29.41 L.atm.mol^{-1}. Therefore, the partial pressure of CO_2 in the PM storage tank is calculated as follows:

$$P_{CO2} = \ K_{CO2} \times x_{co2} \qquad EQN.1$$

Where,

K_{CO2} = Henry's constant for CO_2 in water at 298°K

x_{CO2} = concentration of CO_2 in water in equilibrium with the gas at 298°K; 0.0083 mol.L^{-1}.

This calculation results in a CO_2 partial pressure of 0.25 atm. Therefore, for a CO_2 concentration of 32wt-% in flue gas, the required total pressure is calculated to be 0.77 atm, which for a 50% excess, amounts to 1.15 atm. Using the ideal gas law, at 298°K, the steady state required concentration of the gas in the PM media storage tank is calculated to be 0.047 mol.L^{-1}. In order to maintain this steady state concentration of CO_2 in the PM media storage tank, the required flue gas volumetric flow rate through the media is calculated to be 16 m^3 per hour *per field* as shown below:

$$\frac{\left(\frac{1{,}253\ m^3}{S.tank}\times\frac{1000L}{1\ m^3}\times\frac{0.0083\ moleCO_2}{L}\times\frac{44\ gCO_2}{1\ moleCO_2}\times\frac{1\ gf.g.}{0.32\ gCO_2}\times\frac{molef.g.}{33\ g}\times\frac{22.4Lf.g}{molef.g.}\right)}{60\ hr}=16\frac{m^3}{hr}$$

This is equivalent to 565 ft^3 per hour per field.

Because each reactor bed tube can be modeled as a plug flow reactor (PFR), it was interesting to determine the CO_2 concentration profile along the tube, which is shown in figure 19. Mapping out the concentration profile of CO_2 is useful in determining the concentration of CO_2 at the end of the tubes so as to ensure that CO_2 is supplied in sufficient amounts, and is not a limiting factor in algae growth. Please refer to Section B of the Appendix on page 274 for the solution of a PFR differential material balance equation, subject to the appropriate boundary conditions.

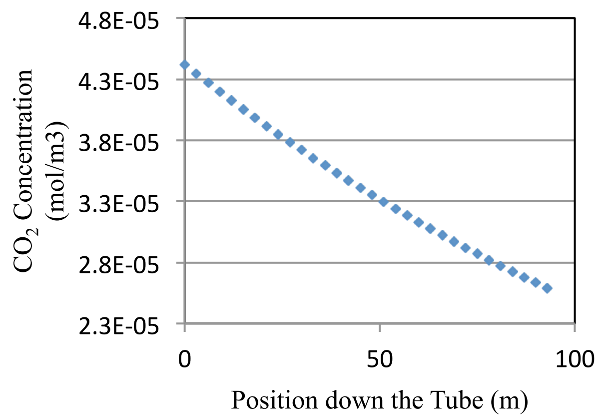


Figure 19: The predicted CO_2 concentration profile down the tubes for a constant CO_2 consumption rate.

Evaluating the Supply Capacity of the Electric Generating Station

Since algae require CO_2 as a carbon source for growth and photosynthesis, a reliable and sufficient source of CO_2 must be identified. The location of the cultivation system near an electric generating station, the W.A. Parish Electric Generating Station, suggests the use of the plant's flue gas emissions as a source of CO_2. Please see table VI for the flue gas composition reported for the power plant.

Table VI: Composition & Flow Rate of Flue Gas

Component	wt.%	Flow Rate (lb s^{-1})
CO_2	0.32	0.142
O_2	0.03	0.0133
N_2	0.58	0.257
H_2O	0.06	0.0266
SO_2	0.01	0.00444
NO_2	0.01	0.00444
Hg	Trace	Trace
Sum	1	0.4437

The plant generates approximately 3,565 MW, making it the largest generator of electricity in the largest electricity generating system in Texas[62]. For the purposes of this project, only emissions from the coal-fired portion of the power plant (plant units 5-8) will be considered, which sum to a total of 2470 MW of energy generation. The W.A. Parish coal fired power plant generates 2.117 lb CO_2 KWh^{-1} as reported in their website, referenced earlier. In order to evaluate the plant's capacity for supplying CO_2 for algae cultivation, one must calculate the rate of CO_2 produced from the coal-fired power plant:

$$3{,}565{,}000 \text{ kW} \times 2.117 \frac{\text{lb}CO_2}{kWh} \times \frac{hour}{3600\ s} = 2{,}096 \frac{lbCO_2}{s}$$

Therefore, at a CO_2 consumption rate of 8.6 lb s^{-1} per field, this plant produces enough CO_2 to provide for 33,270 fields, more than enough to the cover the 107 fields designed for Module I.

$$2{,}096 \frac{lbCO_2}{s} \times \frac{s\ .\ field}{(0.042 \times (1 + 0.5))lbCO_2} = 33{,}270\ fields$$

Fresh Water Consumption (PM, FM)

Depending on the strain of interest, algae can be classified as either freshwater algae, or saltwater-algae; *C. protothecoides* is a freshwater algae strain. Fresh water is required for both the autotrophic and heterotrophic stages of the cultivation process.[9] The fresh water requirement is highly dependent on the amount of PM and FM media recovered from FM and PM clarifiers. For the purposes of this project, the clarifiers are assumed to recover 95% of the media. To prevent the concentration of solid wastes suspended in the media from accumulating in the system, 5% of recovered media will be purged. Therefore, the amount of fresh water supplied must make up for the two sources of water loss from the system: water trapped in the algae slurry and purged waste water. Therefore, the amount of fresh water supply required for both the PM and FM stages is shown below:

- PM Stage

$$\left(186\frac{m^3}{hr}\times(0.05+0.05)\right)=18.6\frac{m^3}{hr}\ =80\frac{gal}{min}perfield$$

- FM Stage

$$\left(191\frac{m^3}{hr}\times(0.05+0.05)\right)=19.1\frac{m^3}{hr}=84\frac{gal}{min}perfield$$

Therefore, the required total fresh water supply rate is 38 $m^3.hr^{-1}$ or 164 gallons per minute *per field*. It is important note that this is a conservative estimate of the amount of water required for cultivation, since the water losses are calculated above as a fraction of the total algae fluid, which includes both algae cells and media. While this assumption may be a good approximation in the PM stage because the algae cell density is no greater than 0.2 g L^1, in the FM stage, the exiting volumetric flow rate of 840 gallons per minute is comprised of 96 gallons per minute of algae cells, and only 744 gallons per minute of FM medium. Taking the algae cells into account, the amount of water purged would be revised to 37 gallons per minute, and the amount of water trapped in the algae slurry would also be 37 gallons per minute. This more exact approach would translate to a water requirement of 74 gallons per minute per field, and not the conservative estimate of 84 gallons per minute as stated above.

Nutrient Consumption (PM)

The nutrient used to nourish the algae is Guillard's f/2 formula, which is used to grow microalgae that are then fed to bivalve larvae. Culturists recommend using Guillard's f/2 formula at a rate of 2 mL per liter of algae cultured[63]. This amount is recommended for algae cultures that have a density of 100,000-200,000 algae cells.

$$\frac{0.002\frac{L_{nutrients}}{Lalgae}\times19{,}925\frac{\mathrm{m}^3}{hr}\times1{,}000\frac{\mathrm{L}}{m^3}\times\frac{hr}{3600\ s}}{107\ fields}=0.103\ \frac{L_{nutrients}}{s}=1.6\frac{gal}{min}perfield$$

Although 90% of the PM media is recycled back to the storage tank, this does not change the amount of nutrients that must be supplied, which was determined to be 0.103 L s^{-1}; for lack of other information, the nutrients are assumed to be fully consumed by the algae, and must therefore be fully replenished.

Oxygen Production (PM)

During the process of photosynthesis, carbon dioxide and water consumed, and in the presence of light, carbon dioxide is fixed into glucose, an organic compound, and oxygen is released. An overall equation for photosynthesis is given by

$$12H_2O + 6CO_2 + \text{light} \rightarrow C_6H_{12}O_6 + 6O_2 + 6\ H_2O$$

The equation shown above implies that for each molecule of CO_2 consumed, a molecule of O_2 is released, however, this 1-to-1 correspondence does not account for the oxygen molecules that result from the splitting of water.[60] Therefore, while it may take 8 photons of light to produce one molecule of O_2, it takes between 8 and 12 photons to fix one CO_2 molecule. Therefore, assuming 10 photons of light are required to fix 1 molecule of CO_2, the rate of oxygen production per field can be calculated as follows:

$$\frac{\frac{8\ molO_2}{10\ molCO_2} \times \frac{32\frac{gO_2}{molO_2}}{\frac{44\ gCO_2}{molCO_2}} \times \frac{1.83\ gCO_2}{galgae} \times \frac{3{,}985.10^3 galgae}{hour} \times \frac{hour}{3600\ s}}{107\ fields} = \frac{11\ \ gO_2}{s} = 0.024\frac{lbO_2}{s} perfield$$

The buildup of oxygen in closed autotrophic cultures is a concern when adequate mass transfer is not easily achieved, which is especially a challenge when the system is scaled up. Therefore, in order to determine whether this would be an issue in the design proposed for this project, the concentration profile of O_2 down the tube was mapped out using the same equations used to map out the CO_2 concentration profile shown above, equations 4 and 5. The concentration profile obtained in shown in figure 20, and as can be seen, the concentration of O_2 down the tube increases exponentially, as dictated by equation 5. The information required to obtain this concentration profile is the algae specific growth rate, the stoichiometric amount of released oxygen (1.07 g_{O2}/g_{Algae}), and the initial concentration of algae in the PM medium. The initial concentration of oxygen in the PM medium can be obt ained from Henry's law: Henry's constant for oxygen in water is given as 769.23 $L.atm.mol^{-1}$, and for an oxygen partial pressure of 0.21 atm, the initial concentration of O_2 in the PM medium is calculated to be $2.73.10^{-7}$ $mol.m^{-3}$. The import information that this concentration profile reveals is that even at the end of the tube, the concentration of O_2 is very low, on the order of 10^{-7} $mol.m^{-3}$, and with the amount of mixing in this system, the buildup of oxygen is assumed to not be a concern. However, if studies performed on pilot plans reveal the contrary, then based on a recommendation by Prof. Wen K. Shieh at the *University of Pennsylvania*, thin and relatively long (3-4 in) plexiglass degassing tubes can be inserted at various points along the tubes for oxygen relief if deemed necessary. These would add negligible cost to the overall design.

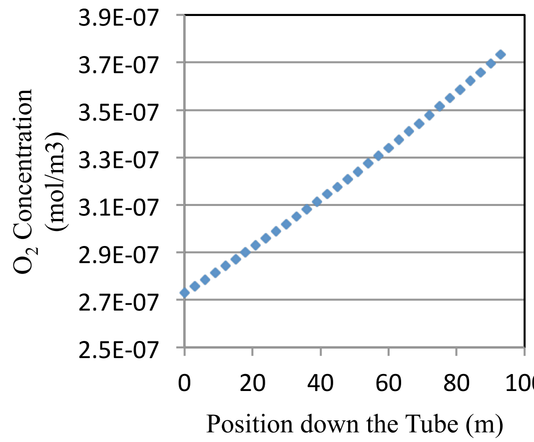


Figure 20: The predicted O_2 concentration profile down the tubes for a constant O_2 consumption rate.

Glycerol Consumption (FM)

Wu et al. recommend placing the algae in glucose, acetate or other organic compound-rich medium to allow for heterotrophic growth, which results in both high biomass and high lipid content in microalgae cells. Grady et al. (2010) examined the use of glycerol as both the only carbon source and in combination with glucose for the cultivation of heterotrophic *C. protothecoides* cultures. Heredia-Arroyo et al. (2010) recommend using a glycerol concentration of 10g L^{-1}, which was shown to result in similar high biomass and lipid content as glucose-rich media.[1] Given that glycerol is one of the by-products of the transesterification of triglycerides, with very little commercial value, a decision was made to use glycerol as the carbon source during the fermentation process. A. To determine the glycerol consumption during the heterotrophic growth (FM) phase *per field*:

$$\frac{10.0\,\frac{gGlycerol}{L}\times\frac{L}{0.123\ kg\ algae}\times 2{,}450{,}718\,\frac{kg\ algae}{hour}\times\frac{hr}{3{,}600\ s}}{107\ fields}=517\,\frac{g}{s}=1.14\,\frac{lb}{s}\,perfield$$

This corresponds to an annual consumption of roughly 16 million kg of glycerol, or 18 thousand tons of glycerol. The amount of glycerol produced as a by-product of the transesterification process is estimated at roughly 95 million kg per year. Therefore, there will be no issues in supplying the required amount of glycerol.

Oxygen Consumption (FM)

Xiong et al. recommend maintaining the dissolved oxygen content in the fermentation tanks over 20% of air saturation[64], which corresponds to a value of 50 μM. Each cultivation field will have 1 fermentation tanks, each of which will have a volume capacity of 1,909 m^3. Oxygen is supplied in the form of compressed air, which generally has an oxygen concentration of 21% by volume. The air flow rate required *per field* is calculated to be 1.02 m^3 of air per hour, as shown below:

$$\frac{\left(\frac{1{,}909\ m^3}{tank}\times\frac{1000L}{1\ m^3}\times\frac{50.10^{-6}moles\ O_2}{L}\times\frac{22.4\ LO2,STP}{1\ moleO_2}\times\frac{1\ LAir}{0.21\ LO_2}\right)}{10\ hr}=\ 1.02\,\frac{m^3Air}{hr}=4.5\,\frac{gal}{min}$$

Henry's law is used to determine the partial pressure of oxygen required in the fermentation tank to ensure a concentration of 50.10^{-6} moles per liter. Henry's constant for O_2 in water at 298°K is given as 769.23 $L.atm.mol^{-1}$. The partial pressure of O_2 in the FM fermentation tank is calculated to be 0.038 atm. Air has an oxygen composition of 23.20% by mass, which results in a total pressure of 0.17 atm. Therefore, to ensure an oxygen concentration of 50.10^{-6} moles per liter, air has to be supplied at a pressure of at least 0.17 atm.

Energy Balance for the Fermentation Tanks (FM)

One of the byproducts of fermentation is heat. It is therefore necessary to determine the amount of heat released in the fermentation tanks under conditions of aerobic respiration.

In order to determine the amount of heat that needs to be removed from the fermentation tanks through auxiliary cooling to maintain an operating temperature of 28±1°C, an energy balance is performed, which yields the following expression:[65]

$$Q_{accumulated} = Q_{fermentation} - Q_{surrounding} \qquad \textit{EQN.2}$$

Where,

$Q_{accumulated}$ = accumulated heat that must be removed
$Q_{fermentation}$ = heat of the fermentation reaction
$Q_{surrounding}$ = heat lost to the surroundings through the walls of the fermentation tank

The heat of fermentation released during glycerol consumption may not be readily found in published literature. Therefore, the heat of fermentation of glucose, given as 121 $kJ.mol_{glucose}^{-1}$ was employed as a close approximation.[66] For a glycerol concentration of 10 gL^{-1}, a volumetric production rate of 20,423 m^3 per tank, a heat capacity of 4.184 $kJ.kg^{-1}°C^{-1}$, the adiabatic temperature rise is calculated to be $°C.hr^{-1}$ as shown below:

$$\frac{\frac{10g_{glu}}{L}\times\frac{20{,}423m^3}{hr}\times\frac{1{,}000\ L}{m^3}\times\frac{121kJ}{mol_{glu}}\times\frac{mol_{glu}}{180g_{glu}}}{107\ Tanks} = 1{,}909m^3\times\frac{1{,}000kg}{m^3}\times\frac{4.184kJ}{kg.°C}\times\Delta T\left(\frac{°C}{hr}\right)$$

The adiabatic increase in temperature is calculated to be 0.17°C per hour.

The amount of heat loss through the tank walls was calculated as shown in Section B of the Appendix on page 274 and found to be negligible in comparison to the amount of heat generated due to fermentation. Therefore, the fermentation contents would be expected to rise in temperature above 28°C without any auxiliary heating or cooling. Chilled water supplied at roughly 4°C will be required to maintain the temperature as close to 28°C as possible. Based on an adiabatic temperature rise of 0.17°C per hour in each fermentation tank, therefore, to determine the chilled water requirement per tank *per field*:

$$1{,}909m^3\times\frac{1{,}000kg}{m^3}\times\frac{4.184kJ}{kg.°C}\times\frac{0.17°C}{hr}\times\frac{Btu}{1.055\ kJ}\times 1\frac{ton}{12{,}000\frac{Btu}{hr}}\times 365\frac{ton-day}{ton} = \frac{39{,}148\ ton-day}{year.tank}$$

Energy Requirements (PM, FM)

Energy input into the algal cultivation system takes on a couple of forms: solar energy required for autotrophic growth of algae and electrical energy required to pump the algae fluid as well as the PM and FM media through the cultivation system. Shown below are the calculations performed to determine the energy requirements *per field*. Bear in mind, the entire cultivation system comprises 107 fields.

Solar Energy (PM)

Xiong et al. (1992) report growing batches of *C. protothecoides* cultures under conditions of 100 μmol m^{-2} s^{-1}or 29.5 $Wm^{-2}$of light exposure.[9] An entire field of 20 acres that contains a mere 12 acres of net reactor area would require roughly 1,400 kW of sunlight energy, or a total of roughly 147,000 kW of sunlight for all 107 fields to produce 20,000 bpd of biodiesel.

$$29.5\,\frac{W}{m^2}\times 12\,\frac{acres_{reactor}}{field}\times 4{,}047\,\frac{m^2}{acre}\times\frac{1kW}{1{,}000\,W} = \;1{,}400\;kW per field$$

Centrifugal Pump Power Requirement (PM/FM Media)

Centrifugal pumps will be used to propel the movement of algae fluid, as well as PM and FM fluid throughout the cultivation system. In order to calculate the power requirement for each pump, the required pump head must first be calculated. Two centrifugal pumps will be used to circulate the PM and FM media in and out of their respective storage tanks to ensure mixing at a turnover rate of once per day. All pumps per field are shown below in figure 21. A detailed description of each pump -its function, its location, its type and power requirement- may be found in the equipment list and unit descriptions below. The calculations performed to determine the power requirement of each of those pumps can be found on the Sample Pump Calculations page in Section E of the Appendix on page 366**.**

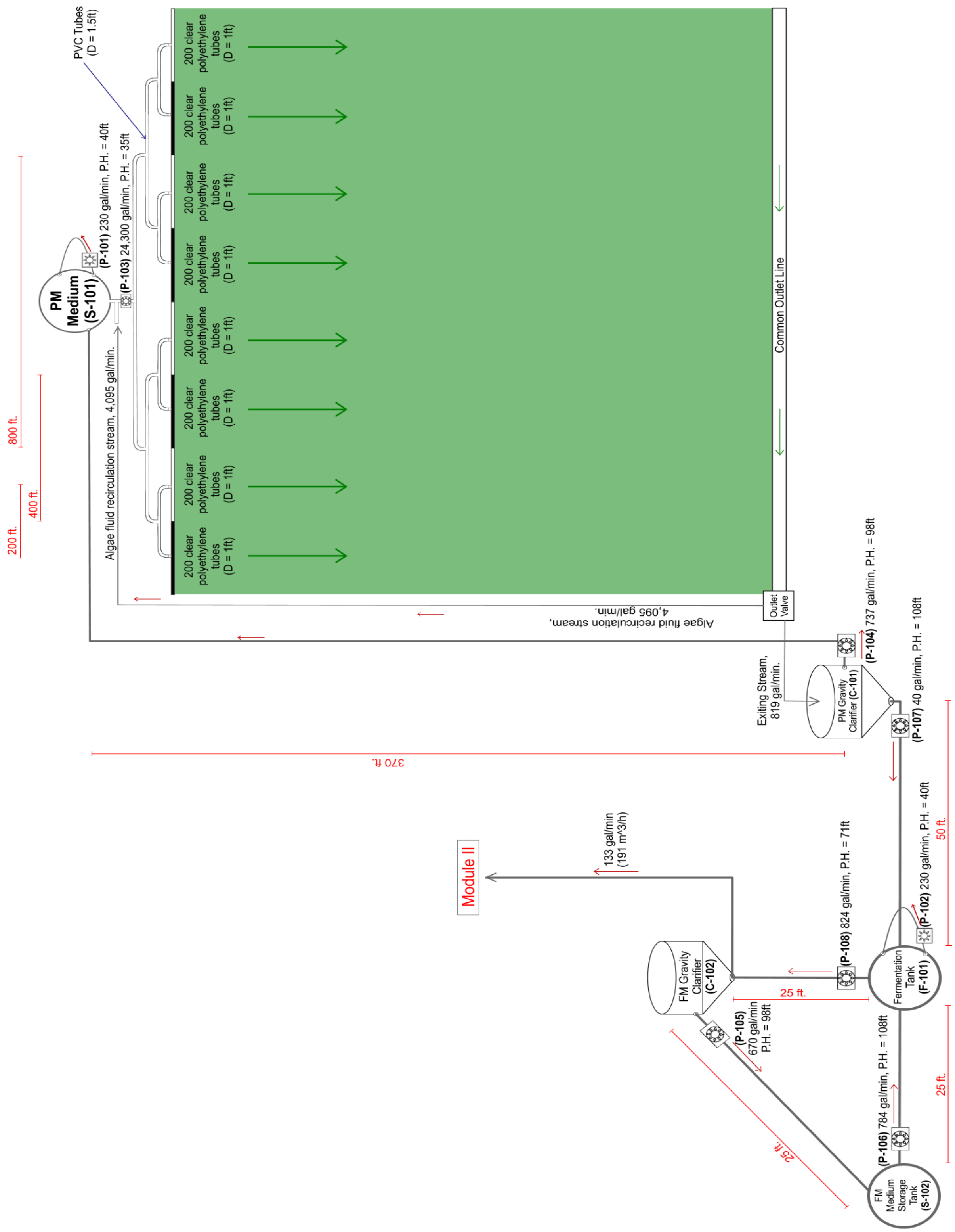


Figure 21: A top-view schematic of a single field with all the pump locations clearly marked.

I. Equipment List

Shown below is a comprehensive list of the equipment required for the PFM cultivation process, which include storage tanks, fermenters, gravity clarifiers and centrifugal pumps. The type, function, size and material of each unit are displayed, as well as the operating temperature and pressure. This information is particularly useful when calculating the cost for each piece of equipment as well as its associated utilities.

Table VII: A List of the Equipment and Relevant Parameters used in PFM Cultivation

Unit #	Equipment Type	Function	Size	Mat'l	Oper T (°F)	Oper P (PSI)
S-101	Storage Tank	Stores PM medium before it is pumped into tubes	D = 40 ft H = 40 ft t_s = 0.625 in V = 819 gal/min	Carbon Steel	77	14.7 - 20
S-102	Storage Tank	Stores FM medium before it is pumped into fermentation tank	D = 40 ft H = 40 ft t_s = 0.625 in V = 744gal/min	Carbon Steel	77	14.7 - 20
C-101	Gravity Clarifier	Separates PM medium from algae cells	D = 40 ft H = 14 ft A_s = 2420 ft^2 V = 819 gal/min	Carbon Steel	77	14.7
C-102	Gravity Clarifier	Separates FM medium from algae cells	D = 40 ft H = 14 ft A_s = 2420 ft^2 V = 841 gal/min	Carbon Steel	77	14.7
F-101	Fermentation Tank	Heterotrophic growth of algae cells in glycerol-rich FM medium	D = 42 ft H = 50 ft t_s = 0.625 in V = 744 gal/min	Carbon Steel	83	14.7-20
P-101	Centrifugal Circulation Pump	Circulates the PM medium in and out of the PM Storage Tank (S-101)	Pc = 4hP V = 230 gal/min L = 40 ft H=40 ft	Cast Iron	77	32
P-102	Centrifugal Circulation Pump	Circulates the FM medium in and out of the FM Storage Tank (S-102)	Pc = 4hP V = 230 gal/min L = 40 ft H=40 ft	Cast Iron	77	32

P-103	Centrifugal Pump	Increases the pressure of the PM medium common inlet line from S-101	Pc = 332hP V =24300 gal/min L = 6630 ft	Cast Iron	77	20
P-104	Centrifugal Pump	Increases the pressure of the PM medium recycle stream from C-101 to S-101	Pc = 31 hP V = 737 gal/min L = 1170 ft H=40 ft	Cast Iron	77	33
P-105	Centrifugal Pump	Increases the pressure of the FM medium recycle stream from C-102 to S-102	Pc = 27 hP V = 670 gal/min L = 25 ft H=40 ft	Cast Iron	77	32
P-106	Centrifugal Pump	Increases the pressure of the FM medium stream from S-102 to F-101	Pc = 31 hP V = 744 gal/min L = 25 ft H= 50 ft	Cast Iron	77	36
P-107	Centrifugal Pump	Increases the pressure of the separated algae slurry stream from C-101 to F-101	Pc = 2 hP V = 40 gal/min L = 50 ft H= 50 ft	Cast Iron	77	36
P-108	Centrifugal Pump	Increases the pressure of the algae fluid stream from F-101 to C-102	Pc = 24 hP V = 840 gal/min L = 25 ft H= 14 ft	Cast Iron	77	20

J. Unit Descriptions

PM Medium Storage Tank (S-101)

The PM medium storage tank stores one day's worth of PM media, a sufficient inventory given that algae do not need a fresh daily supply of PM medium for survival and can remain suspended in the same medium for up to a week at a time. The storage tank is 40 feet high, has a diameter of 40 feet, and has a total volume of 377,000 gallons. The contents of the tank are re-circulated with an external pump to ensure proper mixing. One PM medium storage tank is required per field. The tanks will be constructed of carbon steel, with an operating pressure of 14.7-20 psi, and a bare-module cost of roughly \$1.14MM per tank. Please refer to the pump specification sheet on page 80 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

FM Medium Storage Tank (S-102)

The FM medium storage tank stores one day's worth of FM media, a sufficient inventory given that algae do not need a fresh daily supply of FM medium for survival and can remain suspended in the same medium for up to a week at a time. The storage tank is 40 feet high, has a diameter of 40 feet, and has a total volume of 377,000 gallons. The contents of the tank are re-circulated with an external pump to ensure proper mixing. One FM medium storage tank is required per field. The tanks will be constructed of carbon steel, with an operating pressure of 14.7-20 psi, and a bare-module cost of roughly $1.14MM per tank. Please refer to the pump specification sheet on page 80 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Fermentation Tank (F-101)

The details of fermentation tank design are extensively described in Section F on page 55. Please refer to the pump specification sheet on page 82 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

PM Gravity Clarifier (C-101)

This gravity clarifier separates the algae fluid coming exiting the photosynthetic stage into separate streams of algae slurry and recovered PM medium. The recovered PM medium stream is pumped to the PM medium storage tank, where its f/2 nutrient and CO2 content is replenished. The separated algae slurry cells are collected at the bottom of the gravity clarifier, which is emptied twice or three times a day via a pneumatic valve, and are then sent to the fermentation tank. Five percent of the PM medium is assumed to remain trapped in the separated algae slurry stream while another five percent of the recovered PM medium is purged to prevent the accumulation of solid particulates in the system. Given a volumetric flow rate of 186 gallons per minute, the clarifier has a settling area of 2420 ft^2, and a height of 14 feet. One PM gravity clarifier is required per field. The clarifiers are built of carbon steel, and will each have a bare-module cost of roughly $581,000. Please refer to the pump specification sheet on page 83 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

FM Gravity Clarifier (C-102)

This gravity clarifier separates the algae fluid coming from the fermentation tank into separate streams of algae slurry and recovered FM medium. The recovered FM medium stream is pumped to the FM medium storage tank, where its glycerol and oxygen content is replenished. The separated algae slurry cells are collected at the bottom of the gravity clarifier, which is emptied twice or three times a day via a pneumatic valve, and are then sent to Module II for lipid extraction. 5% of the FM medium is assumed to remain trapped in the separated algae slurry stream while 5% of the recovered FM medium is purged to prevent the accumulation of solid particulates in the system. For a volumetric flow rate of 191 gallons per minute, the clarifier was calculated to have a settling area of 2420 ft^2, and a height of 14 feet. 1 PM gravity clarifier will be required per field. The clarifiers are built of carbon steel, and will each have a bare-module cost of roughly $581,000. Please refer to the pump specification sheet on page 84 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Circulation Pump (P-101)

This pump is used to re-circulate the contents of the PM medium storage tank to ensure proper mixing. To ensure a turnover rate of 1.day-1, the volumetric flow rate of the recirculation stream was determined to be 230 gallons per minute. For 40-ft high storage tanks, a 0.5-ft pipe diameter, a pump efficiency of 0.7, a required pump head of 40 feet was calculated. These specifications correspond to a power requirement of 8 kW at each location *per field*. The pump will be constructed of cast iron, and will have a bare-module cost of roughly $13,600. Please refer to the pump specification sheet on page 85 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Circulation Pump (P-102)

This pump is used to re-circulate the contents of the PM medium storage tank to ensure proper mixing.This is required because the size of the tank is too big to allow for an effective use of stirrers. To ensure a turnover rate of 1.day-1, the volumetric flow rate of the recirculation stream was determined to be 230 gallons per minute. For 40-ft high storage tanks, a 0.5-ft pipe diameter, a pump efficiency of 0.7, a required pump head of 40 feet was calculated. These specifications correspond to a power requirement of 8 kW at each location *per field*. The pump will be constructed of cast iron, and will have a bare-module cost of roughly $13,600. Please refer to the pump specification sheet on page 86 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Pump (P-103)

This centrifugal pump will be required to initiate the movement of the PM medium from the PM medium storage tank into the tubes in the cultivation field. The pressure supplied by this pump has to be sufficient to allow the algae fluid to re-circulate 5 times through the tubes. A pump will not be required to re-circulate the algae fluid in the reverse direction, as the algae fluid will flow downwards by gravity. For an 18-in pipe, an equivalent length of 616 ft, and a volumetric flow rate of 24,300 gallons per minute, a required pump head of 35 feet was calculated. These specifications correspond to a power requirement of 760 kW *per field*. The pump will be constructed of cast iron, and will have a bare-module cost of roughly $216,000. Please refer to the pump specification sheet on page 87 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Pump (P-104)

This centrifugal pump initiates the movement of the recovered PM medium from the PM gravity clarifier back to the PM storage tank For a 0.25-in pipe diameter, a 1170-ft long pipe, a 40-ft high storage tank, and a volumetric flow rate of 737 gallons per minute, a required pump head of 98 feet was calculated. These specifications correspond to a power requirement of 29 kW *per field*. The pump will be constructed of cast iron, and will have a bare-module cost of roughly $27,000. Please refer to the pump specification sheet on page 88 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Pump (P-105)

This centrifugal pump initiates the movement of the FM medium recycle stream from the FM gravity clarifier to the FM medium storage tank. For a 0.25-in pipe diameter, a 25-ft long pipe, a 40-ft high storage tank, and a volumetric flow rate of 670 gallons per minute, a required pump head of 98 feet was calculated. These specifications correspond to a power requirement of 27 kW *per field.* The pump will be constructed of cast iron, and will have a bare-module cost of roughly \$25,500. Please refer to the pump specification sheet on page 89 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Pump (P-106)

This centrifugal pump initiates the movement of the FM medium from the FM storage tank to the fermentation tank. For a 0.25-in pipe diameter, a 25-ft long pipe, a 50-ft high fermentation tank, and a volumetric flow rate of 784 gallons per minute, a required pump head of 108 feet was calculated. These specifications correspond to a power requirement of 31 kW *per field.* The pump will be constructed of cast iron, and will have a bare-module cost of roughly \$27,200. Please refer to the pump specification sheet on page 90 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Pump (P-107)

This pump initiates the movement of the separated algae slurry from the PM gravity clarifier to the fermentation tank.For a 0.25-in pipe diameter, a 50-ft long pipe, a 50-ft high fermentation tank, and a volumetric flow rate of 40 gallons per minute, a required pump head of 108 feet was calculated. These specifications correspond to a power requirement of 1.6 kW *per field.* The pump will be constructed of cast iron, and will have a bare-module cost of roughly \$12,000. Please refer to the pump specification sheet on page 91 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

Centrifugal Pump (P-108)

This pump initiates the movement of the algae fluid stream from the fermentation tank to the FM gravity clarifier.For a 0.25-in pipe diameter, a 25-ft long pipe, a 14-ft high gravity clarifier, and a volumetric flow rate of 824 gallons per minute, a required pump head of 71 feet was calculated. These specifications correspond to a power requirement of 22 kW *per field*. The pump will be constructed of cast iron, and will have a bare-module cost of roughly \$25,000. Please refer to the pump specification sheet on page 92 and the Pump Sample Calculations page in Section E of the Appendix on page 366.

PM Medium Storage Tank

Identification:	Item:	Tank		Date: April 5, 2011
	Item No:	S-101		By: DC/SG/JI
	No. Required:	107		
Function:	Stores 1 day of Photosynthetic Medium (PM) Feed			
Operation:	Continuous			
Materials Handled:				
			Inlet	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (gal/min):		819	819
	Composition:			
		Fresh Water	80	
		F/2 Nutrient	2	2
		CO_2	negligible	negligible
		Recycled PM Medium	737	--
		PM Medium Feed	--	819
Design Data:				
	Type:		Floating Roof	
	Material:		Steel Alloy	
	Volume (ft^3)		44,242	
	C_P	$	372,400	
	C_{BM}	$	1,135,700	

FM Medium Storage Tank

Identification:	Item:	Tank	Date: April 5, 2011	
	Item No:	S-102	By: DC/SG/JI	
	No. Required:	107		
Function:	Stores 1 day of Heterotrophic Medium Feed			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (gal/min):		744	744
	Composition:	Fresh Water	67	--
		Glycerol	7	--
		Oxygen	negligible	negligible
		Recycled FM Medium	670	--
		FM Medium Feed	--	744
Design Data:				
	Type:	Floating Roof		
	Material:	Steel Alloy		
	Volume (ft^3)	60,400		
	C_P	$	372,400	
	C_{BM}	$	1,135,700	

Fermentation Tank

Identification:	Item:	Tank	Date: April 5, 2011
	Item No:	F-101	By: DC/SG/JI
	No. Required:	107	

Function: Fermentation of Algae in Glycerol-Rich FM Medium for 10 Hours

Operation: Continuous

Materials Handled:		**Inlet**	**Outlet**
Inlet Stream ID:		N/A	N/A
Quantity (gal/min):		840	840
Composition:			
	Algae Cells + FM Medium	--	840
	FM Medium	744	--
	Algae Slurry from C-101	40	--

Design Data:

Type:	Floating Roof
Material:	Steel Alloy
Volume (ft^3)	67,500

C_P	$	461,600
C_{BM}	$	1,407,900

PM Gravity Clarifier

Identification:	Item:	Tank	Date: April 5, 2011
	Item No:	C-101	By: DC/SG/JI
	No. Required:	107	

Function: Separates PM medium from algae cells by settling to produce a clear liquid

Operation: Semi-Continuous

Note: Separated algae slurry accumulates at the bottom of the clarifier and is collected and transported to the fermentation tank 2-3 times a day; semi-continuous

Materials Handled:

		Inlet	**Outlet**
Inlet Stream ID:		N/A	
Quantity (gal/min):		819	819
Composition:			
	Algae Cells + PM Medium	819	--
	PM Medium	--	737
	Algae Slurry	--	40
	Purged PM Medium	--	40

Design Data:

Type:	Gravity Clarifier
Material:	Steel Alloy
Settling Area (ft2)	2,420
Height (ft)	14

C_P	$	335,600
C_{BM}	$	580,600

FM Gravity Clarifier

Identification:	Item:	Tank	Date: April 5, 2011
	Item No:	C-102	By: DC/SG/JI
	No. Required:	107	

Function: Separates FM medium from algae cells by settling to produce a clear liquid

Operation: Semi-Continuous

Note: Separated algae slurry accumulates at the bottom of the clarifier and is collected and transported to the fermentation tank 2-3 times a day; semi-continuous

Materials Handled:

		Inlet	**Outlet**
Inlet Stream ID:		N/A	N/A
Quantity (gal/min):		840	840
Composition:			
	Algae Cells + FM Medium	840	--
	FM Medium	--	670
	Algae Slurry	--	133
	Purged FM Medium	--	37

Design Data:

Type:	Gravity Clarifier
Material:	Steel Alloy
Settling Area (ft2)	2,420
Height (ft)	14

C_P	$	335,900
C_{BM}	$	581,100

Centrifugal Pump

Identification:	Item:	Pump	Date: April 5, 2011
	Item No:	P-101	By: DC/SG/JI
	No. Required:	107	

Function:	Circulates the PM medium in and out of the PM Storage Tank (S-101)
Operation:	Continuous

Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		230	230
	Composition:			
		Fresh Water	230	230
		F/2 Nutrient	negligible	negligible
		CO_2	negligible	negligible

Design Data:		
	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	4
	Pressure (psi):	20
	Efficiency:	0.7

C_P	$	900
C_{BM}	$	13,600

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-102		By: DC/SG/JI
	No. Required:	107		
Function:	Circulates the FM medium in and out of the FM Storage Tank (S-102)			
Operation:	Continuous			
Materials Handled:				
			Inlet	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		230	230
	Composition:			
		Fresh Water	230	230
		Glycerol	2	2
		Oxygen	negligible	negligible
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		24	
	Pressure (psi):		20	
	Efficiency:		0.7	
	C_P	$	900	
	C_{BM}	$	13,600	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-103		By: DC/SG/JI
	No. Required:	107		
Function:	Increases the pressure of the PM medium common inlet line from S-101			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		20,210	20,210
	Composition:			
		Fresh Water	230	230
		F/2 Nutrient	negligible	negligible
		CO2	negligible	negligible
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		332	
	Pressure (psi):		20	
	Efficiency:		0.7	
	C_P	$	37,700	
	C_{BM}	$	216,000	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-104		By: DC/SG/JI
	No. Required:	107		
Function:	Increases pressure of PM medium recycle stream from C-101 to S-101			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		728	728
	Composition:			
		Fresh Water	728	728
		F/2 Nutrient	negligible	negligible
		CO2	negligible	negligible
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		31	
	Pressure (psi):		34	
	Efficiency:		0.7	
	C_P	$	4,800	
	C_{BM}	$	27,200	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-105		By: DC/SG/JI
	No. Required:	107		
Function:	Increases pressure of FM medium recycle stream from C-102 to S-102			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		742	742
	Composition:			
		Fresh Water	742	742
		Glycerol	negligible	negligible
		Oxygen	negligible	negligible
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		27	
	Pressure (psi):		32	
	Efficiency:		0.7	
	C_P	$	4,700	
	C_{BM}	$	25,400	

Centrifugal Pump

Identification:	Item:	Pump	Date: April 5, 2011
	Item No:	P-106	By: DC/SG/JI
	No. Required:	107	
Function:	Increases the pressure of the FM medium stream from S-102 to F-101		
Operation:	Continuous		

Materials Handled:

		Inlet	Outlet
Inlet Stream ID:		N/A	N/A
Flow Rate (gal/min):		748	748
Composition:			
	Fresh Water	742	742
	Glycerol	6	6
	Oxygen	negligible	negligible

Design Data:

Type:	Centrifugal Pump
Material:	Cast Iron
Net Work Req (HP)	31
Pressure (psi):	36
Efficiency:	0.7

C_P	$	4,800
C_{BM}	$	27,200

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-107		By: DC/SG/JI
	No. Required:	107		
Function:	Increases pressure of separated algae slurry stream from C-101 to F-101			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		40	40
	Composition:			
		Separated Algae Slurry	40	40
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		2	
	Pressure (psi):		36	
	Efficiency:		0.7	
	C_P	$	3,000	
	C_{BM}	$	11,900	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-108		By: DC/SG/JI
	No. Required:	107		

Function:	Increases the pressure of the algae fluid stream from F-101 to C-102
Operation:	Continuous

Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		824	824
	Composition:			
		Algae Cells + FM Medium	824	824

Design Data:		
	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	24
	Pressure (psi):	20
	Efficiency:	0.7

C_P	$	4,800
C_{BM}	$	25,000

3. Algae Cultivation: Feasibility & Economic Analysis

What follows is a thorough economic analysis of the PFM cultivation process as presented above. An overall economic analysis considering the three modules will be presented in later sections, highlighting cultivation's contribution to the economic feasibility of the overall process.

A. Fixed-Capital Investment Summary

Shown below in table VIII is the equipment cost summary for each of the equipment items of the algae PFM cultivation process. The purchase cost corresponds to the cost of the physical equipment, while the bare module cost is the cost of the equipment after installation costs and fabrication-on-site costs are included. The bare module factor varies depending on the equipment type and size; a listing of such values can be found in Table 22.11 of *Product and Process Design Principles*.[67] The purchase costs for most of these equipment units have been estimated using unit-specific correlations listed in Chapter 22 of *Product and Process Design Principle.s*[67]

Table VIII: Equipment Cost Summary for the Algae PFM Cultivation Process

Designation	Equipment Description	Purchase Cost	Bare Module Factor	Bare Module Cost
STORAGE TANKS				
S-101	PM Storage Tank	39,842,000	3.05	121,519,000
S-102	FM Storage Tank	39,842,000	3.05	121,519,000
GRAVITY CLARIFIERS				
C-101	PM Gravity Clarifier	35,903,900	1.73	62,114,000
C-102	FM Gravity Clarifier	35,937,200	1.73	62,172,000
FERMENTATION TANKS				
F-101	Fermentation Tank	49,389,200	3.05	150,638,000
PUMPS				
P-101	Centrifugal Pump	439,100	3.30	1,450,000
P-102	Centrifugal Pump	439,100	3.30	1,450,000
P-103	Centrifugal Pump	7,000,800	3.30	23,103,000
P-104	Centrifugal Pump	881,500	3.30	2,909,000
P-105	Centrifugal Pump	821,300	3.30	2,711,000
P-106	Centrifugal Pump	881,400	3.30	2,909,000
P-107	Centrifugal Pump	384,300	3.30	1,269,000
P-108	Centrifugal Pump	808,800	3.30	2,670,000
	TOTAL BARE MODULE COST ($)			**556,433,000**

The Fixed Capital Investment Summary shown below in table XI lists the various capital investments that were estimated for the algae PFM cultivation process. The direct permanent investment, C_{DPI}, includes the total bare-module cost of the equipment, the total bare-module cost of the spares, the total bare-module cost of the storage tanks, the cost of site preparation, service facilities cost, and allocated costs for utility plants.[67] The total bare-module cost for the first three listed items is \$590,904,000. The cost of site preparation includes the cost of soil and earthmoving required to place the reactor beds at an incline of roughly 4°, or an elevation of 20 feet at the opposite end point. It was estimated that roughly 185,500 yd^3 of earth will be required to be moved per field, for a total of 20,000,000 yd^3 for all fields. A cost estimate of \$10/yd^3 for the price of earthmoving was obtained from the firm *Langan Engineering & Environmental Services*. Since this cost was based on on-site earthwork that required moving only 4,000 yd^3, this price cannot be directly applied to estimating the cost of moving 20,000,000 yd^3 of earth. Henderson's Law may be used to estimate the price of earthmoving at a significantly larger scale, as shown below in equation 6:

$$p_n = p_1 \times n^{-\alpha} \qquad EQN.3$$

For an α of 0.7, a figure recommended by *Product & Process Design*[67], a cost estimate of \$0.002 per yd^3 was obtained, which translates to an overall expenditure of roughly \$40,000 for all fields. This is a crude estimate of the cost of earthmoving. Therefore, the cost of site preparations was not based entirely on the cost of earthwork and was instead conservatively estimated at 5.0% of total bare-module costs. This translates to a sum that is significantly greater than the cost of earthmoving cited above. Since the algae cultivation fields will be located at a plant with existing utility infrastructure, no allocated costs were considered. The C_{DPI} was calculated to be \$654,396,000.

The total depreciable capital, C_{TDC}, is calculating by adding the cost of contingencies and contractor's fees (18% of C_{DPI}) to C_{DPI}, which was calculated to be \$772,188,000. The total permanent investment, C_{TPI}, is calculated based on the total depreciable capital and other non-depreciable investments such as land and plant startups.[67] Land costs are generally estimated to be approximately 2% of C_{TDC}. However, according to Diversified Energy Corporation (DEC), the capital costs for installation, which includes estimates for the cost of land, harvesting, and product storage for the Simgae™ cultivation system is within the range of \$45,000 to \$60,000 per acre. To remain conservative, the cost of land was estimated at the higher price of $60,000. Startup costs are generally assessed at 10% of C_{TDC}. An initial royalty fee would be paid to Diversified Energy Corporation (DEC) for the use of their licensed algae cultivation technology and design. This royalty is approximately 2% of C_{TDC} as suggested in Section 23.2 of Product and Process Design Principles. The C_{TPI} was calculated to be $916,032,000 as shown below.

It is important to note that the cost of land is on the order of $129 million. This is a significant departure from the result presented in the design report by Carlson et al. (2010) for a phototrophic cultivation model, which allocated roughly $2.2 billion for the cost of land. This significant reduction in cost of land results primarily from adopting an autotrophic-heterotrophic growth model that capitalizes on double CO_2-fixation during the fermentation process. This reduction in cost is also marginally attributed to the increase in the diameter of the reactor tubes, which allows one to optimize land use and field acreage.

To complete the evaluation of the total capital investment, it is necessary to determine the working capital, which is the difference between current assets and liabilities. Working capital is typically calculated by summing 30 days of cash reserves, which cover the cost of manufacture, 30 days of accounts receivable, which cover the month long delay in receipt of payment, and inventory, which is assumed to be 5 days at the sales price. Shown below in table IX is a summary of the production costs, which include feedstocks, utilities, operations, maintenance, operating overhead, property taxes and insurance. Cash reserves are estimated as a twelfth of the total annual cost of manufacture, which was calculated below to be $232,754,000. Also shown in table X is the working capital summary for the algae cultivation process.

Table IX: Cost of Manufacture Summary for the Algae PFM Cultivation Process

Cost of Manufacture			
Feedstocks		$	3,375,000
Utilities		$	78,454,000
Operations		$	3,360,000
Maintenance		$	62,163,000
Operating Overhead		$	8,182,000
Property Taxes and Insurance		$	15,444,000
Depreciation			
	Direct Plant	$	61,776,000
	Allocated Plant	$	0
COST OF MANUFACTURE		**$**	**232,754,000**

Table X: Working Capital Summary for the Algae PFM Cultivation Process

Cash Reserves	$	19,396,167
Accounts Payable	$	6,819,083
WORKING CAPITAL	**$**	**26,216,000**

Unlike variable costs, which scale with production rate, fixed costs are expenses that are charged regardless of production rate. These expenses may be estimated from the C_{TDC} and the number of process sections.[64] When adding the working capital, C_{WC}, to C_{TCI} ,the total capital investment C_{TCI} is calculated to be $1,053,440,000, or roughly $1,060,000,000 as shown below in table XI.

Table XI: Fixed Capital Investment Summary for the Algae PFM Cultivation Process

C_{TBM}	**Total Bare Module Cost**		
C_{PM}	Equipment Bare Module Costs	$	313,395,000
C_{spare}	Total Bare-Module Costs for Spares	$	38,471,000
$C_{storage}$	Total Bare-Module Costs for Storage Tanks	$	243,038,000
		C_{TBM}	594,904,000
C_{DPI}	**Direct Permanent Investment**		
C_{TBM}	Total Bare-Module Cost	$	594,904,000
C_{site}	Site Preparation	$	29,746,000
C_{serv}	Service Facilities	$	29,746,000
C_{alloc}	Allocated Costs for Utility Plants	$	0
		C_{DPI}	654,396,000
C_{TDC}	**Total Depreciable Capital**		
C_{DPI}	Direct Permanent Investment	$	654,396,000
C_{cont}	Contingencies and Contractor's Fees	$	117,792,000
		C_{TDC}	772,188,000
C_{TPI}	**Total Permanent Investment**		
C_{TDC}	Total Depreciable Capital	$	772,188,000
C_{land}	Cost of Land	$	128,400,000
C_{royal}	Cost of Royalties	$	15,444,000
		C_{TPI}	916,032,000
C_{TCI}	**Total Capital Investment**		
C_{TPI}	Total Permanent Investment	$	916,032,000
C_{WC}	Working Capital	$	137,405,000
		C_{TCI}	**1,053,440,000**

B. Operating Costs

Shown below are the calculations used to determine the continuous costs of PFM algae cultivation. The raw materials required for algae cultivation are f/2 Guillard's nutrient medium, carbon dioxide, fresh water, oxygen, and glycerol. Chilled water will be required to maintain the temperature of the fermentation tanks at roughly 28°C. Electrical energy will be supplied to meet the pump power requirement, as well as agitation of the contents of the fermentation tanks.

Nutrients: Guillard's F/2 medium (PM)

The nutrient used to nourish the algae is Guillard's f/2 formula, which is used to grow microalgae that are then fed to bivalve larvae. The nutrient consumption rate was calculated above at 0.0255 liters per second, which translates to annual consumption of 727,056 liters. The main supplier of this medium is Sigma-Aldrich[6], which sells a unit of 10 liters of Guillard's f/2 medium at a price of $18.5. Therefore, at this selling price, the continuous cost of the nutrients is calculated to be:

$$0.103\ \frac{Lnutrients}{s} \times \frac{\$18.5}{10\ L} = \frac{\$0.19}{s} perfield$$

This number would translate to an annual cost of $6MM, however, it does not accurately reflect the cost of the nutrients because it is estimated based on a very small quantity that should not be directly applied to the large scale of operation of the cultivation fields. Industrial culturists recommend preparing the medium from its single components when such large volumes are required, or when the recipe needs to be modified. Shown in Section B of the Appendix on page 274 is the recipe for the Guillard f/2 medium, as well as the costs of the nutrient's individual components. While the costs may vary based on the source of those components, the new estimated cost of the nutrient is now as low as $0.0098 per liter[6], which, *per field*, translates to continuous cost of:

$$0.103\ \frac{Lnutrients}{s} \times \frac{\$0.0098}{L} = \frac{\$0.001}{s} perfield$$

This more accurate estimation translates to an annual cost of \$31,536 per field, which is approximately 190 times less than the original estimate for the annual cost of \$6 MM.

Carbon Dioxide Gas (PM)

The source of CO_2 is a nearby coal-fired plant that emits enough CO_2 to supply around 33,270 fields as shown above. CO_2 is assumed to come with no associated cost. In fact, under a carbon cap-and-trade system, the utilization of CO_2 gas may even become a source of revenue for the algae-to-biodiesel venture, as will be discussed in the *Overall Economic Analysis* Section. The required flue gas volumetric flow rate is calculated to be 16 m^3 per hour *per field.*

Fresh Water (PM, FM)

While the selection of saltwater algae strains allows one to avoid the competitive market for freshwater, the high biomass and lipid content achieved in the study performed by Xu et al. are specific to *C. protothecoides,* which is a freshwater algae strain.[28] The consumption rate of fresh water for both PM and FM stages was calculated above at 18.6 $m^3\ hr^{-1}$ and 19.1$m^3\ hr^{-1}$ per field for the PM and FM stages, respectively, or a total of 38 $m^3\ hr^{-1}$ per field, which translates to a consumption rate of 10.5 liters per second per field. This number accounts for the fact that both PM and FM clarifiers allow for the recovery of roughly 90% of both PM and FM media, respectively. This results in a significant reduction in the continuous cost of fresh water. The cost of fresh water is assumed to be that of process water: \$0.75 per 1000 gallons. Therefore, *per field*:

$$\frac{\$0.75}{1{,}000 gallon} \times \frac{gallon}{3.78\ Liters} \times 10.5 \frac{Liters}{s} = \frac{\$0.002}{s} perfield$$

Glycerol (FM)

As discussed previously, given that glycerol is one of the by-products of the transesterification of triglyecerides, it will be the main carbon source for the fermentation of algae. The annual consumption of glycerol was calculated above at roughly 16 million kg of glycerol, or 18 thousand tons of glycerol. The amount of glycerol produced as a by-product of the transesterification process is estimated at roughly 92 million kg per year. Therefore, there will be no issues in supplying the required amount of glycerol. Moreover, glycerol is assumed to come with no associated cost.

Oxygen in Air (FM)

Xiong et al. recommend maintaining the dissolved oxygen content in the fermentation tanks over 20% of air saturation[9], which corresponds to a value of 50 μM. Each cultivation field will have 1 fermentation tanks, each of which will have a volume capacity of 1,909m^3. Therefore, to maintain this concentration in the fermentation tanks, air with an oxygen content of 21% by volume must be supplied at a volumetric flow rate of 1.02m^3 per hour per field. The cost of air is assumed to be negligible.

Chilled Water Requirement (FM)

The chilled water requirement to cool the fermentation tanks was calculated above as 39,148 ton-day per year per tank. Therefore, given that chilled water at 40°F is priced at $1.20/ton-day[67], the annual operating cost per fermentation tank (or per field) is calculated to be $47,300.

Electrical Energy Requirement (PM, FM)

Electrical energy will be supplied to meet the pump power requirement, as well as to enable agitation of the contents of the fermentation tanks. Shown below in table XII is a summary of the electricity costs for the PFM cultivation of algae. The annual consumption (kW-hr) per field is shown as well as the total annual cost per pump for all 107 fields. These costs are summed to calculate the total annual electricity cost for all the pumps and fermentation tanks on all fields. This sum amounts to $66,643,000.

The variable costs in the PFM algae cultivation process include raw material costs and utilities, as shown in table XIII. Raw material costs include the cost of f/2 Guillard's algae nutrient, compressed air, flue gas and glycerol for an annual total cost of $3,375,000. Compressed air and flue gas are assumed to come at no or negligible cost.

Table XII: Electricity Costs Summary of the Algae PFM Cultivation Process

ELECTRICITY				
Unit	**Equipment**	**Annual Consumption (kW-hr) per Field**	**Price**	**Total Annual Cost ($)**
F-101	Fermentation Tank	1,647,100	$0.060/kW-hr	10,574,400
P-101	Centrifugal Pump	72,800	$0.060/kW-hr	467,400
P-102	Centrifugal Pump	72,800	$0.060/kW-hr	467,400
P-103	Centrifugal Pump	7,688,500	$0.060/kW-hr	49,360,200
P-104	Centrifugal Pump	233,200	$0.060/kW-hr	1,497,100
P-105	Centrifugal Pump	212,000	$0.060/kW-hr	1,361,000
P-106	Centrifugal Pump	240,600	$0.060/kW-hr	1,544,700
P-107	Centrifugal Pump	13,800	$0.060/kW-hr	88,600
P-108	Centrifugal Pump	199,600	$0.060/kW-hr	1,281,400
		TOTAL ANNUAL COST ($)		**66,643,000**

Crude glycerol is one of the by-products of the transesterification process and is also assumed to come at no cost. Utilities cost include the cost of electricity, chilled water, and fresh water with an annual cost of $78,454,000. No byproducts are produced in the process of algae cultivation, and there are no selling/transfer expenses to be included in the general expenses section. These will be accounted for in the economic analysis of the tranesterification process, and shown again in the overall economic analysis section.

Table XIII: Variable Costs Summary of the Algae PFM Cultivation Process

Feedstock (Raw Materials)		
f/2 Guillard's Algae Nutrient	$	3,375,000
Compressed Air	$	0
Flue Gas	$	0
Glycerol	$	0
	$	**3,375,000**
Utilities		
Electricity	$	66,643,000
Chilled Water	$	5,062,000
Fresh Water	$	6,749,000
	$	**78,454,000**
TOTAL VARIABLE COSTS	**$**	**81,829,000**

C. Fixed Costs

Shown in table XIV are the fixed costs associated with PFM algae cultivation, which include operating costs, maintenance costs, operating overhead costs, as well as property taxes and insurance. These costs are incurred independent of the production rate of algae slurry. The values shown in table XIV are calculated based on a specified percentage of the total depreciable capital (C_{TDC}) as recommended in Section 23.2 of *Product and Design Principles*.[67]

The operating costs are calculated with the assumption that there are five operators per shift. This number may need to be revised at some point during implementation of the project. It is assumed that the fields will be monitored and controlled remotely and that any malfunction in the pumps or leaking in the tubes is not a critical issue that must be resolved instantly; there are no vessels on the fields that are operating at either high temperature or pressure, and algae cells are capable of surviving for prolonged periods of time without being continuously pumped through the fields or supplied with fresh media. A means of transportation will be provided to allow the operators to travel across the fields for maintenance if required. A total annual cost of $3,360,000 was calculated for the operating costs. Maintenance costs include costs for wages and benefits, salaries and benefits for the engineering and supervisor employees and personnel, materials and services required to keep equipment in acceptable working order, and maintenance overhead. A total annual cost of $62,163,000 was calculated for the maintenance costs.

Operating Overhead includes the cost of general plant overhead costs, mechanical department, medical services, employees relations' department and business services. The operating overhead includes costs not directly related to plant operating, such as medical and safety services, and may be estimated as a fraction of the combined salary, wages and benefits for maintenance and labor-related operations. A total annual cost of $15,444,000 was calculated for the operating overhead costs. The final item on the fixed costs spread sheet is the property taxes and insurance, which is estimated at 2% of the total capital investment C_{TCI}. The total annual fixed costs amounts to $89,149,000.

Table XIV: Fixed Costs Summary of the Algae PFM Cultivation Process

Operations		
Direct Wages and Benefits	$	1,820,000
Direct Salaries and Benefits	$	273,000
Operating Supplies and Services	$	17,000
Technical assistance to manufacturing	$	600,000
Control laboratory	$	650,000
	$	**3,360,000**
Maintenance		
Wages and Benefits	$	27,027,000
Salaries and Benefits	$	6,757,000
Materials and Services	$	27,027,000
Maintenance Overhead	$	1,352,000
	$	**62,163,000**
Operating Overhead		
General Plant Overhead	$	2,548,000
Mechanical Department Services	$	862,000
Employee Relations Department	$	2,117,000
Business Services	$	2,655,000
	$	**8,182,000**
Property Taxes and Insurance	**$**	**15,444,000**
Total Fixed Costs	**$**	**89,149,000**

V. MODULE II: LIPID EXTRACTION

1. Lipid Extraction: Concept Stage

A. An Overview

Because the conversion of lipids into transportation fuels cannot be performed within algal cells, lipids must be first extracted and separated from the remaining biomass. Since this lipid extraction process consumes a significant amount of energy, designing an energy efficient process has thus far been a bottle neck in the algae-to-biofuel venture. Once the algae fluid in Module I has reached the desired cell density in the fermentation tanks, it sent to the FM gravity clarifiers, from which the recovered algae slurry is sent to the lipid extraction unit. Conventional lipid extraction processes generally consist of three steps: solid separation, dewatering, and extraction; each of these steps is highly energy intensive, and thus, expensive. The separation of algae cells from the surrounding medium is expensive because algae cells and water have very similar densities, and therefore cannot be separated easily by gravity.[78] Additionally, water has a high heat of vaporization, so drying algae cells requires large amounts of heat. Finally, lipid extraction is energy intensive because algal cell walls have a strong exterior and a high modulus of elasticity. This makes it very difficult to efficiently break the cell wall – a prerequisite for lipid extraction. What follows is a thorough discussion of the conventional lipid extraction processes, and the recommended alternative, the Single-Step Extraction™ currently under development by OriginOil™.

B. The Conventional Lipid Extraction Process

Figure 22 is a diagram of a conventional mechanical-solvent extraction system.[6] Algae slurry obtained from an algae cultivation process is first sent to a mechanical solid separations unit, such as a centrifuge, where the algae are dewatered. A 70% water content is generally specified for the output stream.[6] The effluent is then thermally dewatered with a steam dryer to reduce the moisture content to 10%. To initiate cell wall-breakage, the cells are mechanically smashed with a presser, and sent to a hexane extraction unit. The hexane further degrades the cell walls, and two phases form: miscella (solvent with 10-15% oil) and biomass (solid with 30% solvent).

Distillation of the miscella is required to separate the lipids from the rich hexane leaving the extraction unit.[6] Similarly, the hexane must be thermally removed from the biomass. The recovered hexane is then recycled. What follows is thorough description of each of the aforementioned three steps common to any conventional lipid extraction process.

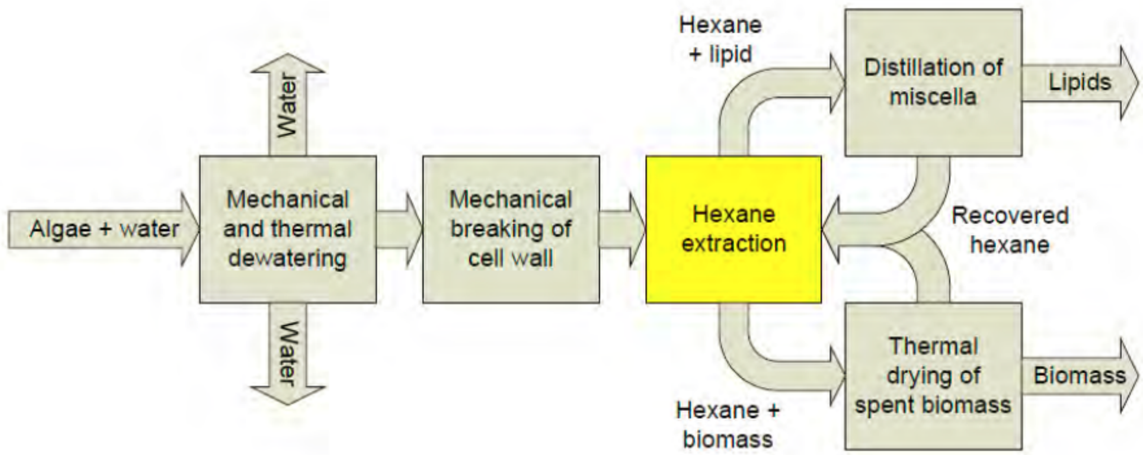


Figure 22: A basic diagram of a typical mechanical, solvent extraction system.[6]

Solids Separation

During solid separation, the majority of the water in the algae slurry is separated from the algae cells. This may be achieved through polymer flocculation, centrifuging, or hydrocycloning[68]. Polymer flocculation is a process that capitalizes on a polymer's ability to attract algae cells, causing them to aggregate[69]. This method is effective for large quantities of algae slurry and is generally less energy intensive than mechanical methods. Since this method does not rely on gravitational separation, it is not limited by the lipid composition of the algae - the greater the lipid content, the closer the algae's density is to that of water. However, flocculating agents are relatively expensive and caustic.[69] The most commonly used separators are centrifuges, which are gravitational separators that rely heavily on differences in density between water and the suspended solids. Although centrifuges are energy and capital intensive, they are effective at separating algae cells from water and functions well for large quantities of algae slurry. A third solid separation technique involves hydrocyclones, which are also gravity separators[70]. In comparison to centrifuge, hydrocyclones consume less energy and have lower operating and capital costs. However, because hydrocyclones are very dependent on density differences, they are limited to algal strains with high specific gravities.

Dewatering

After the initial solids separation, the biomass is sent to a dewatering process. A few common dewatering methods are steam drying, fluidized bed drying, and microwave radiation. In steam drying, hot steam is used to evaporate the moisture in the biomass. While the process is very effective, it is also energy intensive. In a fluidized bed dryer, algae is fluidized in air, which may or may not be heated[71]. As the air passes through the biomass, its moisture content increases. In turn, the water content of the biomass decreases. A fluidized bed dryer is very effective and does not require high maintenance costs. While this process does not require much heating, it is still relatively energy intensive. A third way to dry the biomass involves microwaves. Electromagnetic radiation is used to energize the water molecules trapped in the biomass and results in significant evaporation. This method is more energy efficient than steam drying, but because microwaves lack penetration power, the biomass may be unevenly dried[72].

Extraction

Once the algae slurry has been dewatered, it is sent to the extraction unit where the lipids are separated from the biomass. There are several methods of lipid extraction including mechanical extraction, solvent extraction, and enzyme extraction. Expellers and presses mechanically break up the cell walls and compress the perforated algae cells to extract the lipids. This method does not require any chemical solvents and recovers up to 80% of the lipids[73]. However, the design of each unit is specific to each algal strain, thereby leading to high capital costs. Additionally, this type of extraction involves high energy costs and requires constant maintenance. Although this method does not require chemical input, it is often run in combination with solvent extraction units.[73] In solvent extraction, solvents chemically degrade the cell walls and solubilize the lipids. Once the lipids dissolve into the solvent, the mixture is sent to a distillation unit where the solvent and the lipids are separated. Common solvents used for algae lipid extraction are hexane, benzene, and ether. Solvent extraction is relatively inexpensive and up to 95% effective[74]; however, it requires the use of caustic chemicals. Enzymes are also often used for lipid extraction. This process is quite similar to solvent extraction. Enzymes degrade the cell wall and a water solvent solubilizes the lipids.[68]

C. New Lipid Extraction Technologies

One emerging technology under investigation for the separation of water and biomass is electrophoresis. Electrophoresis removes charged algae cells from solution by applying an electric field, without the need for additional chemicals.[75] Other novel lipid extraction techniques involve the use of acoustics, sonication, mesoporous nanomaterials, and amphiphilic solvents. SRS Inc. is currently exploring some of these new methods in addition to the use of supercritical fluids for lipid extraction at a test facility in New Mexico. They have demonstrated that supercritical fluid extraction is safer and faster than traditional hexane extraction.[76] The OriginOil™ Single-Step Extraction process has also been proposed as an alternative to conventional lipid extraction. Capitalizing on the use of electromagnetic pulses and microbubbles that generate shock waves, the OriginOil™ Single-Step Extraction process is highly energy-efficient and has been suggested to consume just 10% of the energy used during the conventional process.[78] Another technique currently being evaluated is a milking system that manipulates the hydrophobicity of the algal cell membrane. This enables lipid exaction from algae cells without rupturing their cell walls.[77]

D. Proposed Process: OriginOil™ Extraction

Of the many alternatives to conventional lipid-extraction explored above, the OriginOil Single-Step Extraction™ was selected for further investigation and inclusion in the algae-to-biofuel venture. Significantly, the single-step extraction process skips the solid separation and dewatering phase, thereby eliminating two energy-intensive processes. Considering its promise of low energy costs, the Single-Step Extraction Process was deemed to be the most attractive of the various alternatives studied. A diagram of Single-Step Extraction from the OriginOil™ website is shown below in figure 23.[78]

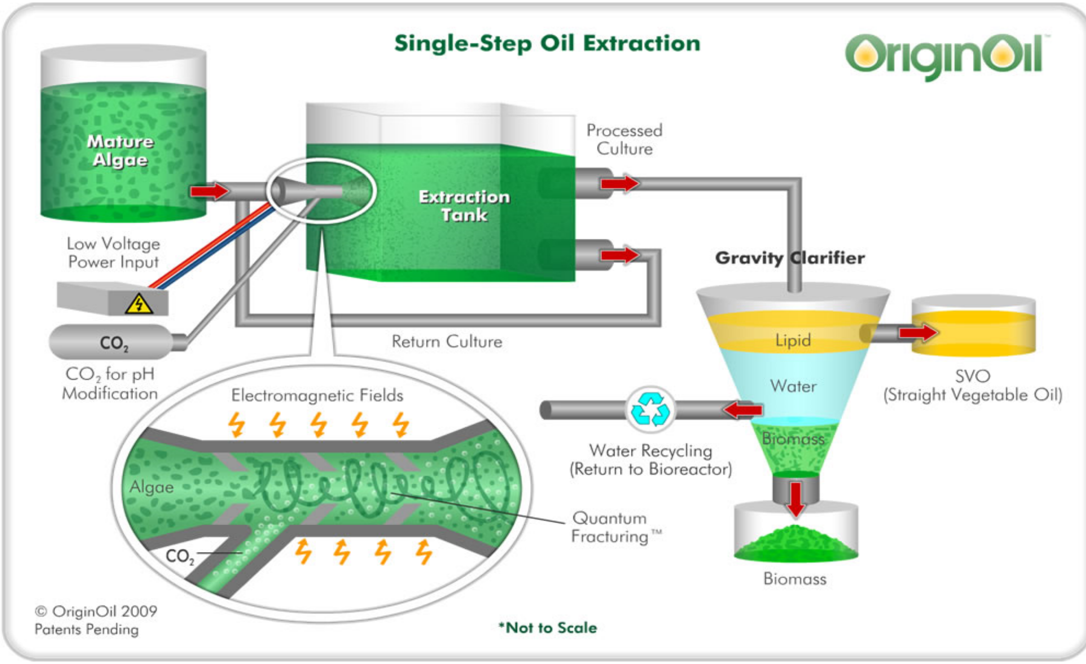


Figure 23[78]: A diagram of the single-step extraction as obtained from OriginOil™.[78]

Quantum Fracturing™

Quantum Fracturing™ was developed by Nicholas Eckelberry and T. Riggs Eckelberry as a less energy intensive method for breaking apart algal cell walls.[79] Quantum Fracturing™ is a fourfold process. First, micro and nano CO_2 bubbles and water are passed in a recirculating pump until a hyper-excited state is reached. Upon reaching this state, the micron mix is expelled into a vessel containing the algal slurry. Upon disgorgement into the vessel, an inward propagating shock wave is produced. First, the shock wave passes over the bubbles and a significant pressure pulse is generated. In response, the bubbles implode and ultrasound waves are produced, enhancing cellular breakdown. Thirdly, friction from cellular breakdown as well as the refraction of shock waves on the cellulose material generates heat. This increases the rate of algal oxidation and further weakens the cell wall. Simultaneously, CO_2 released from the imploding bubbles dissolves in the algal slurry and reduces its pH, further oxidizing and weakening algal cell walls. Together, these events comprise Quantum Fracturing™, a very effective process.[79]

Electromagnetic Pulse

After passing through the Quantum Fracturing™ process, the algae slurry is pumped to an upper manifold chamber connected to a succession of tubes in parallel. The algae flow into the manifold and down the parallel tubes for electromagnetic pulse treatment. This process further weakens the algal cell wall to improve extraction efficiency. Figure 24 identifies a potential tube layout based on the Single-Step Extraction™ patent application. [106]

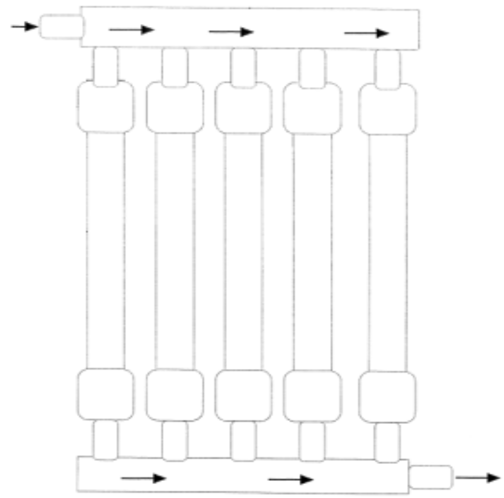


Figure 24: The tube layout proposed in the Single-Step Extraction™ patent application.[1]

Each tube is concentric and operates like an electrical cell. The outer tube behaves as an anode, the inner tube serves as a cathode, and the algae slurry is a naturally conductive medium.[106] Accordingly, when power is supplied and the cell is complete, both electrical and magnetic fields are generated. The power input oscillates, and the algal cells swell and shrink in response. Each electromagnetic pulse weakens the cell wall just as bending a paperclip back and forth weakens the metal link. Eventually, the continous succession of pulses results in the rupture of the cell wall. [106] Apart from experiencing electromagnetic pulses, the algae slurry is also heated as it travels down the tubes. The heat serves two purposes. An increase in temperature not only increases the rate of oxidation and accelerates cell wall degredation, but also alters the specific gravity of the suspended solids in water, heightening the rate at which biomass sinks in aqueous solution to improve their impending separation in a gravity clarifier. Achieving long residence times in the electromagnetic pulse tubes is essential. [106]

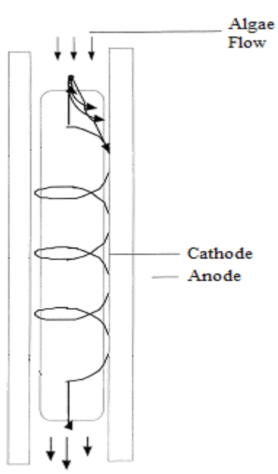


Figure 25: The algae slurry is directed in a spiraling path as it is heated and experiences electromagnetic pulses.

Therefore, the algae slurry is directed into a sprialing path to increase its travel time through the tubes as is shown in Figure 25. [106]

2. Lipid Extraction: Manufacturing & Development Stages

The lipid-extraction process designed for the algae-to-biofuel venture was based extensively on the OriginOil™ Single-Step Extraction process described above. In the section that follows, the process design and its economic feasibility are thoroughly discussed. Since many of the technologies that OriginOil™ has invented are still in the research and development phase and have not yet been commercialized, the company has kept many of the technical details that characterize Single-Step Extraction confidential. The following analysis is based on information limited to that provided by the pending patent application submitted by OriginOil™. Therefore, in order to conduct a complete quantitative analysis, several assumptions were made. These assumptions will be identified and explained in the sections that follow.

A. Process Flow Diagram and the Material Balance

The overall process flow diagram is shown in Figure 26. Notably, lipid-extraction is a bridge between algal cultivation and lipid-processing and is therefore linked to both Modules I and III. The material balance that corresponds to the process flow diagram shown in figure 26 is summarized in table XV. Based on data provided by OriginOil™, it is assumed that 90% of the lipids trapped in the

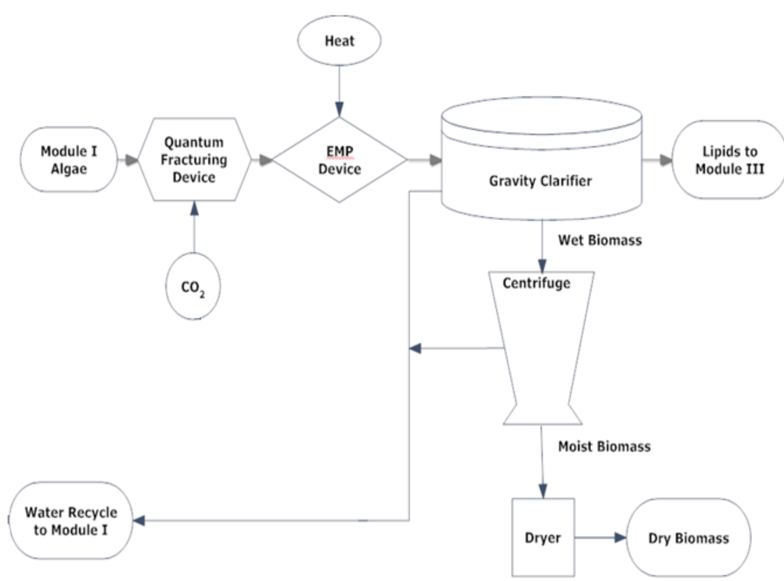


Figure 26: The Lipid-Extraction Process Flow Diagram involves both Quantum Fracturing and Electromagnetic Pulse Treatment and connects Modules I and III.

algae cells exiting Module I are successfully extracted and sent to Module III.[78] This lipid product stream is considered to be pure with only trace concentrations of water and biomass. During the course of lipid-extraction, a biomass byproduct, which must be dewatered to prevent spoiling, is generated. Since achieving specified moisture concentrations for various streams is essential to lipid-extraction, the weight percentage of water in each stream is shown in the material balance below.

Table XV: Summarizes the mass flow rates of each of the exiting streams

Stream	Module I Algae	Lipids	Clarifier Water Recycle	Wet Biomass	Centrifuge Water Recycle	Moist Biomass	Dry Biomass
Mass Flow (lb/hour)	5,403,000	223,700	1,413,000	3,767,000	3,296,000	470,800	418,500
Component Mass Flow							
H_2O	4,803,000	Trace	1,413,000	3,390,000	3,296,000	94,200	41,900
Lipid	276,200	223,700	Trace	52,500	Trace	52,500	52,500
Biomass (Protein+ Carbohydrates)	324,200	Trace	Trace	324,200	Trace	324,200	324,200
H2O Fraction (wt%)	89%	0%	100%	90%	100%	20%	10%

B. Process and Unit Descriptions

This section comprehensively analyzes the operation and design of the lipid-extraction process and the associated equipment, which may be segmented into the following components: 1) Breaking the Cell Wall 2) Gravity Separation 3) Centrifugation 4) Thermal Drying.

Breaking the Cell Wall

Since the lipid extraction facility is located adjacent to the algal cultivation site, algae sludge from Module I may be transported to the lipid extraction feed storage tank by pipeline. The algae sludge, which leaves the storage tank at mass flow rate of 5.4 million lbs per hour and concentration of 7.5 lbs/ft^3, is pumped through the OriginOil™ Single-Step Extraction process, where it is subject to Quantum Fracturing™ and Electromagnetic Pulses. [78] A combination of shockwaves, ultrasound, heat, pH modification, and electromagnetic field generation ruptures the algal cell walls, preparing the slurry for an effective gravity separation.

Gravity Separation

Lysed algae cells from the extraction process enter clarifiers at a mass flow rate of 5.4 million lbs per hour. Each clarifier, designed to facilitate gravitational separation, is circular with a cross sectional area of 102,000 ft^2 and volume of 360,000 ft^3. Typical mixtures of cellular debris in water require more than just gravity to separate. However, OriginOil™ has demonstrated that after being treated with heat and CO_2 micro and nano bubbles, the lipids, water, and biomass separate after just one hour.[78] At elevated temperatures, the specific gravity of biomass changes relative to the other components of the slurry, causing it to sink. Simultaneously, CO_2 aids in releasing any lipids trapped in the cellular debris and bringing them up to the top layer. OriginOil™ claims lipids extraction efficiencies in the range of 85% to 97%. For our analysis, a yield of 90% was assumed. After one hour, most of the biomass has settled to the bottom of the clarifier, while water occupies the middle layer and lipids float on top. The lipids are skimmed off the top layer at a mass flow rate of 224,000 lbs per hour and sent to a lipid-processing facility, where they are converted to biofuel.[78] Water is pumped from the middle layer at a mass flow rate of 1.4 million lbs per hour and into the water storage tank, from which it is recycled for use in algal cultivation. The biomass at the clarifier's bottom is sent to a dewatering system.

Centrifugation

Wet biomass, assumed to be 90% water by mass, from the clarifier is pumped to a centrifuge where a significant fraction of the water is removed. According to Lawrence Wang et al., a high speed continuous solid bowl centrifuge may be used to reduce the moisture content in the biomass to 20%.[80] Five centrifuges, each with a load capacity of 40 solid tons per hour, are used to process 3.8 million lbs of wet biomass, equivalent to 325,000 lbs of solid per hour.[80] From the centrifuge, 3.3 million lbs of water per hour are separated and sent to the water storage tank for recycling to Module I. The moist biomass product is transported to the dryer for additional water removal.

Thermal Drying

Wet biomass enters the dryer with a moisture content of 20% and leaves the dryer with a moisture content of 10%. It is critical that some residual water remain in the biomass to prevent spoiling. To process 470,800 lbs of moist biomass per hour, 17 drum dryers are required. Each drum is 480 ft^2 and has a residence time of 0.5 hours at an operating temperature of 97°F.[80] The biomass byproduct will be discussed in additional detail in Section B of the "Economic and Feasibility Analysis" for Module II.

C. Equipment List and Specification

Shown below is a comprehensive list of the equipment required for the lipid extraction process. The type, function, size and material of each unit are displayed, as well as the operating temperature and pressure. This information is particularly useful when calculating the equipment cost and associated utilities.

Unit #	Equipment Type	Function	Size	Mat'l	Oper T (°F)	Oper P (PSI)
E-101	OriginOil Single Step Extraction Unit	Breaks algae cell wall and extracts lipids	N/A	N/A	N/A	N/A
G-101	Gravity Clarifier	Separates the lipids, water, and biomass	D = 361 ft H = 3.5 ft V = 359,200 ft^3	Concrete	N/A	14.7
G-102	Gravity Clarifier	Separates the lipids, water, and biomass	D = 361 ft H = 3.5 ft V = 359,200 ft^3	Concrete	N/A	14.7
C-101	Continuous Scroll Solid Bowl Centrifuge	Dewaters the moist biomass from the separater	P = 0.01 kWh/kg V = 40 tons/h	Stainless Steel	N/A	14.7
C-102	Continuous Scroll Solid Bowl Centrifuge	Dewaters the moist biomass from the separater	P = 0.01 kWh/kg V = 40 tons/h	Stainless Steel	N/A	14.7
C-103	Continuous Scroll Solid Bowl Centrifuge	Dewaters the moist biomass from the separater	P = 0.01 kWh/kg V = 40 tons/h	Stainless Steel	N/A	14.7
C-104	Continuous Scroll Solid Bowl Centrifuge	Dewaters the moist biomass from the separater	P = 0.01 kWh/kg V = 40 tons/h	Stainless Steel	N/A	14.7

C-105	Continuous Scroll Solid Bowl Centrifuge	Dewaters the moist biomass from the separater	P = 0.01 kWh/kg V = 40 tons/h	Stainless Steel	N/A	14.7
D-101	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-102	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-103	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-104	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-105	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-106	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-107	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-108	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-109	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-110	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-111	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-112	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-113	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h*ft^2	Stainless Steel	97	14.7
D-114	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-115	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7

D-116	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
D-117	Drum Dryer	Evaporates water in moist biomass from centrifuge	A = 480 ft^2 Evp.rate = 40 lb/h.ft^2	Stainless Steel	97	14.7
S-103	Storage Tank	Stores Algae from Module I	D = 200 ft H = 35 ft V = 1,070,000 ft^3	Stainless Steel	97	14.7
S-104	Storage Tank	Stores Recycled Water	D = 80 ft H = 23 ft V = 115,000 ft^3	Stainless Steel	97	14.7
P-109	Centrifugal Pump	Pumps algae from algae storage tank to extraction unit	Pc = 1,994 Hp V = 89,545 gal/min L = 16.5 ft	Cast Iron	N/A	14.7
P-110	Centrifugal Pump	Pumps algae from extraction unit to gravity clarifier	Pc = 634 Hp V = 89,545gal/min H = 1.25 ft	Cast Iron	N/A	14.7
P-111	Centrifugal Pump	Pumps water from clarifier to storage tank	Pc = 18 Hp V = 2,640 gal/min H = 16 ft	Cast Iron	N/A	14.7
P-112	Centrifugal Pump	Pumps Biomass to Centrifuge	Pc = 0.04 Hp V = 86,400 gal/min L = 50 ft	Cast Iron	N/A	14.7

Gravity Clarifier

Identification:

Item:	Clarifier	Date: April 5, 2011
Item No:	G-101, G-102	By: DC/SG/JI
No. Required:	2	

Function: Separates sludge into lipid, biomass, and water streams

Operation: Continuous

Materials Handled:

Stream ID		**Inlet**	**Outlet**	**Outlet**	**Outlet**
		n/a	Lipid	Water	Biomass
Quantity (lb/hr):		5,403,000	223,700	1,413,000	3,767,000
Composition:					
	H2O	89%	0%	100%	90%
	Lipids	5%	100%	0%	1%
	Biomass	6%	0%	0%	9%

Design Data:

Type:	Gravity Clarifier
Material:	Concrete
Volume (ft^3)	359,200

C_P	$	1,737,000
C_{BM}	$	3,579,000

Centrifuge

Identification:	Item:	Centrifuge	Date: April 5, 2011	
	Item No:	C-101 to C-105	By: DC/SG/JI	
	No. Required:	5		

Function: Dewaters the wet biomass from 90% moisture to 20% moisture

Operation: Continuous

Materials Handled:

Stream ID		**Inlet**	**Outlet**	**Outlet**
		Wet Biomass	Moist Biomass	Water
Quantity (lb/hr):		753,300	94,160	659,120
Composition:				
	H_2O	90%	20%	100%
	Lipids	1%	11%	0%
	Biomass	9%	69%	0%

Design Data:

Type:	Continuous Scroll Solid Bowl Centrifuge
Material:	Stainless Steel
Load (tons/hr)	40

C_P	$	380,000
C_{BM}	$	770,400

Dryer

Identification:	Item:	Dryer	Date: April 5, 2011
	Item No:	D-101 to D-117	By: DC/SG/JI
	No. Required:	17	

Function: Dewaters the moist biomass from 20% moisture to 10% moisture

Operation: Continuous

Materials Handled:

Stream ID		**Inlet**	**Outlet**
		Moist Biomass	Dry Biomass
Quantity (lb/hr):		27,700	24,700
Composition:			
	H2O	20%	10%
	Lipids	11%	13%
	Biomass	69%	77%

Design Data:

Type:	Drum Dryer
Material:	Stainless Steel
Heat Transfer Area (ft2)	480
Evaporation Rate (lb/hr-ft2)	6

C_P	$	334,300
C_{BM}	$	688,500

Algae Sludge Storage Tank

Identification:	Item:	Tank		Date: April 5, 2011
	Item No:	S-103		By: DC/SG/JI
	No. Required:	1		
Function:	Stores 1 hour of algae sludge from Module I			
Operation:	Continuous			
Materials Handled:				
			Inlet	**Outlet**
	Inlet Stream ID:		n/a	n/a
	Quantity (gal/min):		n/a	89,545
	Composition:			
		H_2O	89%	89%
		Lipids	5%	5%
		Biomass	6%	6%
Design Data:				
	Type:		Floating Roof	
	Material:		Carbon Steel	
	Volume (ft^3)		359,200	
	C_P	$	1,575,000	
	C_{BM}	$	4,803,700	

Water Storage Tank

Identification:	Item:	Tank	Date: April 5, 2011
	Item No:	S-104	By: DC/SG/JI
	No. Required:	1	

Function: Stores 1 hour of water recycled from clarifier and centrifuge

Operation: Continuous

Materials Handled:

		Inlet	**Outlet**
Inlet Stream ID:		n/a	n/a
Quantity (gal/min):		2,640	n/a
Composition:			
	H2O	100%	100%

Design Data:

Type:	Floating Roof
Material:	Carbon Steel
Volume (ft^3)	115,000

C_P	$	260,600
C_{BM}	$	794,800

Centrifugal Pump

Identification:	Item:	Pump	Date: April 5, 2011
	Item No:	P-109	By: DC/SG/JI
	No. Required:	1	

Function: Pumps algae sludge from storage tank to Single-Step Extraction™ unit

Operation: Continuous

			Inlet	**Outlet**
Materials Handled:				
	Inlet Stream ID:		n/a	n/a
	Flow Rate (gal/min):		89,545	89,545
	Composition:			
		H2O	89%	89%
		Lipids	5%	5%
		Biomass	6%	6%

Design Data:		
	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	1994
	Pressure (psi):	25
	Efficiency:	0.7

C_P	$	511,000
C_{BM}	$	1,686,300

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-110		By: DC/SG/JI
	No. Required:	1		

Function: Pumps algae sludge from Single-Step Extraction™ unit to gravity clarifiers

Operation: Continuous

Materials Handled:

			Inlet	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		89,545	89,545
	Composition:			
		H2O	89%	89%
		Lipids	5%	5%
		Biomass	6%	6%

Design Data:

	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	634
	Pressure (psi):	8
	Efficiency:	0.7

C_P	$	126,500
C_{BM}	$	417,200

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-111		By: DC/SG/JI
	No. Required:	1		

Function: Pumps water recycle stream to water storage tank

Operation: Continuous

Materials Handled:

		Inlet	**Outlet**
Inlet Stream ID:		N/A	N/A
Flow Rate (gal/min):		2,640	2,640
Composition:			
	H2O	100%	100%

Design Data:

Type:	Centrifugal Pump
Material:	Cast Iron
Net Work Req (HP)	18
Pressure (psi):	7
Efficiency:	0.7

C_P	$	10,000
C_{BM}	$	32,800

Centrifugal Pump

Identification:	Item:	Pump	Date: April 5, 2011
	Item No:	P-112	By: DC/SG/JI
	No. Required:	1	

Function: Pumps biomass to centrifuge

Operation: Continuous

			Inlet	**Outlet**
Materials Handled:				
	Inlet Stream ID:		N/A	N/A
	Flow Rate (gal/min):		86,400	86,400
	Composition:			
		H2O	90%	90%
		Lipids	1%	1%
		Biomass	9%	9%

Design Data:		
	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	0.04
	Pressure (psi):	0.0005
	Efficiency:	0.7

C_P	$	424,200
C_{BM}	$	1,399,800

D. Energy Balance and Utility Requirements

Electricity is the only utility necessary for lipid extraction. The following section includes the calculations used to estimate electricity consumption for each item of equipment that comprises the lipid extraction process.

Single-Step Extraction

The most energy intensive step is the Single-Step Extraction™ process consisting of Quantum Fracturing™ and electromagnetic pulse generation. Unfortunately, the details of energy consumption for Single-Step Extraction™ are proprietary information of OriginOil™ and not available to the public. Therefore, an electricity requirement was assumed by scaling up the process discussed by OriginOil™ at the Worldbiofuels Markets conference that occurred on March 15-17, 2010 in Amsterdam.[68] The results presented at the conference are shown below in table XVI.

Table XVI: Electricity requirements for Single-Step Extraction. [81]

Slury Volume	10,000,000 L
Algae Concentration	1 g/L
Energy Requirement	5625 kWh

From the data shown in table XVI, energy consumption per mass of dry algae was calculated for Single-Step Extraction.

$$\frac{5{,}625\ kWh}{10{,}000{,}000\ Liters} \times \frac{1\ Liter}{gram} = 0.000565\,\frac{kWh}{gram}\ or\ 0.5625\,\frac{kWh}{kg}$$

Assuming that the energy consumption required for extraction is directly proportional to the mass of dry algae that enters the process, the rate of energy consumption for Single-Step Extraction™ in Module II was estimated as shown below.

$$Algae\ Dry\ Weight = Lipids + Biomass = 125{,}300\,\frac{kg}{h} + 147{,}100\,\frac{kg}{h} = 272{,}400\,\frac{kg}{h}$$

$$272{,}400\,\frac{kg}{h} \times 0.5625\,\frac{kWh}{kg} = 153{,}200\,\frac{kWh}{h}$$

Clarifier

A gravitational force is responsible for separating the mixture of lipids, water, and biomass that leaves Single-Step Extraction. Therefore, this element of lipid-extraction requires relatively low energy consumption. A small amount of electricity is necessary to run the skimmers that are essential for lipid collection at the top of the clarifiers. Siemens is a global company with experience in constructing and manufacturing clarifiers. Via e-mail correspondence, Mr. Guide Michelutti, a Siemens representative, provided our team with information regarding clarifiers, including an estimated motor power for the skimmer of 2.24kW per clarifier. Using this information, an electricity cost was calculated.

$$2.24kW \times 3600s = 8{,}100\frac{kWh}{h}$$

$$2\ Clarifiers \times 8{,}100\frac{kWh}{h} = 16{,}200\frac{kWh}{h}$$

Centrifuge

In our design, a continuous scroll solid bowl centrifuge, commonly used for sludge dewatering, is employed to remove water from the wet biomass that leaves the clarifer. According to the United States Environmental Protection Agency, the direct electricity cost per mass of dry solid for continuous scroll solid bowl centrifugation is between 0.070kWh/kg to 0.105kWh/kg, while the indirect electricity cost per mass of dry solid is about 0.009kWh/t.[80] An overall value of 0.099kWh/kg dry solids was used for the calculation of centrifugation energy requirements, which are shown below. [80] The calculations made to find the electricity costs of centrifugation are shown below.

$$Algae\ Dry\ Weight = \ Lipids + Biomass = 23{,}800\frac{kg}{h} + 147{,}100\frac{kg}{h} = 170{,}900\frac{kg}{h}$$

$$170{,}900\frac{kg}{h} \times 0.099\frac{kWh}{kg} = 17{,}000\frac{kWh}{h}$$

Dryer

Moist biomass from the centrifuge enters the dryer at a 20% moisture content and must be dried to a 10% moisture content. The mass of water removed was calculated as shown below.

$$170{,}900\frac{kg}{h}\times\left(\frac{0.2\ kg\ H_2O}{0.8\ kg\ solid}-\frac{0.1\ kg\ H_2O}{0.9\ kg\ solid}\right)=23{,}900\frac{kg\ H_2O}{h}$$

The heat of vaporization of water at process conditions is 0.627 kWh. kg^{-1}. This figure may be used to calculate the energy requirement for drying.

$$23{,}900\frac{kg}{h}\times 0.627\frac{kWh}{kg}=14{,}900\frac{kWh}{h}$$

Pumps

There are four centrifugal pumps located throughout the lipid extraction process that are essential for moving fluids from one process unit to the next. One pump empties the algae storage tank, and propels the sludge-like material through the Single-Step Extraction™ unit. A control valve is used to regulate the flow rate of the algal slurry entering the lipid extraction process. A second pump is placed right after the Single-Step Extraction™ unit, where it moves sheared algae into the two clarifiers. After clarification, a third pump is used to transport the spent algae from the clarifier to the centrifuge. A fourth pump recycles water to the water storage tank. The energy requirement for each centrifugal pump is a function of both the volumetric flow rate of pumped fluid in addition to the specified pressure increase. Therefore, each pump has a unique electricity requirement. A sample pressure drop calculation through a liquid-extraction process pipeline is shown below.

$$\Delta P=\frac{128\times Viscosity\times Length\times Volume\ Flow\ Rate}{Diamter^4\times\pi}$$

$$\Delta P(storage\ to\ single\ step)=\frac{128\times 0.001\times 5\times 5.65}{1^4\pi}=1.15\frac{N}{m^2}\ or\ 0.00017\ psi$$

$$With\ Control\ Valve\colon\ \Delta P(storage\ to\ single\ step)=25.00017\ psi$$

Once the pressure drop has been calculated, the pumping power requirement may be computed for each pump. A sample power calculation for a centrifugal pump may be found in Section E, “Equipment Design Calculations,” in the Appendix on page 366.

A detailed summary of the annual electricity consumption and cost for each piece of equipment is shown in Table XVII below. It is important to note that in comparison to other process machinery, the most energy intensive unit is the Single-Step Extraction Process, which is responsible for 75% of the total electricity consumption.

Table XVII: A summary of the annual electricity consumption and equipment cost

ELECTRICITY				
Unit	**Equipment**	**Annual Consumption (kW-hr)**	**Price**	**Annual Cost ($)**
E-101	Single Step Extractor	1,341,772,000	$0.060/kW-hr	80,506,320
G-101	Gravity Clarifier 1	70,550,000	$0.060/kW-hr	4,233,000
G-102	Gravity Clarifier 2	70,550,000	$0.060/kW-hr	4,233,000
C-101	Centrifuge 1	29,633,000	$0.060/kW-hr	1,777,980
C-102	Centrifuge 2	29,633,000	$0.060/kW-hr	1,777,980
C-103	Centrifuge 3	29,633,000	$0.060/kW-hr	1,777,980
C-104	Centrifuge 4	29,633,000	$0.060/kW-hr	1,777,980
C-105	Centrifuge 5	29,633,000	$0.060/kW-hr	1,777,980
D-101	Drum Dryer 1	7,667,000	$0.060/kW-hr	460,020
D-102	Drum Dryer 2	7,667,000	$0.060/kW-hr	460,020
D-103	Drum Dryer 3	7,667,000	$0.060/kW-hr	460,020
D-104	Drum Dryer 4	7,667,000	$0.060/kW-hr	460,020
D-105	Drum Dryer 5	7,667,000	$0.060/kW-hr	460,020
D-106	Drum Dryer 6	7,667,000	$0.060/kW-hr	460,020
D-107	Drum Dryer 7	7,667,000	$0.060/kW-hr	460,020
D-108	Drum Dryer 8	7,667,000	$0.060/kW-hr	460,020
D-109	Drum Dryer 9	7,667,000	$0.060/kW-hr	460,020
D-110	Drum Dryer 10	7,667,000	$0.060/kW-hr	460,020
D-111	Drum Dryer 11	7,667,000	$0.060/kW-hr	460,020
D-112	Drum Dryer 12	7,667,000	$0.060/kW-hr	460,020
D-113	Drum Dryer 13	7,667,000	$0.060/kW-hr	460,020
D-114	Drum Dryer 13	7,667,000	$0.060/kW-hr	460,020
D-115	Drum Dryer 15	7,667,000	$0.060/kW-hr	460,020
D-116	Drum Dryer 16	7,667,000	$0.060/kW-hr	460,020
D-117	Drum Dryer 17	7,667,000	$0.060/kW-hr	460,020
P-101	Centrifugal Pump 1	7,743,000	$0.060/kW-hr	464,580
P-102	Centrifugal Pump 2	193,000	$0.060/kW-hr	11,580
P-103	Centrifugal Pump 3	256,300	$0.060/kW-hr	15,380
P-104	Centrifugal Pump 4	7,743,000	$0.060/kW-hr	464,580

TOTAL	**1,777,311,300**	**$**	**106,638,680**

TOTAL ANNUAL UTILITIES COST	**$**	**106,638,680**

3. Lipid Extraction: Feasibility & Economic Analysis

What follows is a thorough economic analysis of the lipid extraction process as presented above. An overall economic analysis considering the three modules will be presented in later sections, highlighting lipid extraction's contribution to the economic feasibility of the overall process.

A. Fixed Capital Investment Summary

The purchase costs and bare module costs for each item were calculated using recommended estimates from *Product and Process Design Principles*.[67] A sample calculation for each equipment type is shown below.

Gravity Clarifier 1:

$$C_P = 2400(Settling\ Area)^{0.58} = 2400(102{,}200)^{0.58} = \$1{,}737{,}000$$

Continuous Scroll Solid Bowl Centrifuge 1:

$$C_P = 60{,}000(Solid\ Tons\ per\ hour)^{0.5} = 60{,}000(40)^{0.5} = \$380{,}000$$

Drum Dryer 1:

$$C_P = 32{,}000(Heat\ Transfer\ Area)^{0.38} = 32{,}000(480)^{0.38} = \$334{,}300$$

Storage Tank 1:

$$C_P = 475(Volume)^{0.51} = 475(8{,}000{,}000)^{0.51} = \$1{,}575{,}000$$

Centrifugal Pump 1:

$$C_P = 475(Volume)^{0.51} = 475(8{,}000{,}000)^{0.51} = \$1{,}575{,}000$$

The total bare module costs were found to be \$19 million for process machinery, \$7 million for fabricated machinery, \$3.5 million for spares, and \$5.5 million for storage. All equipment is considered to be process machinery with the exception of the two gravity clarifiers and the storage tanks. The required settling area for the gravity clarifier was calculated using table 8-7 from the *Wastewater Engineering: Treatment and Reuse* textbook by Metcalf and Eddy.[59] The table in Metcalf and Eddy gives the necessary design information for typical clarifiers processing an activated sludge process. The valuable information in the table is the average solid loading rate in lb/ft^{2}h. By dividing desired exiting flow rate by this value, the necessary settling area can be calculated. A sample settling area calculation was shown previously in Module I, on pages 62 and 63. As the gravity clarifiers have diameters of 360 ft, transportation of these units is not feasible and they must be fabricated onsite. The extra cost associated with onsite fabrication was taken into account for these items. Sample calculations for the centrifugal pump are shown in Section E of the Appendix, on page 366. All equipment was considered process machinery with the exception of the Single-Step Extraction unit, which will be customized for the required process flow rate. For the purposes of this design project, each single extractor unit was estimated to have a purchase cost of $500,000, with a bare module factor of 2, as shown in the table below. A single spare for each pump was included in these calculations. Table XVIII presents a detailed summary of the purchase and bare-module equipment costs. Based, on the equipment-bare module cost, the fixed capital investment analysis may be performed and the results are shown in table XIX.

Table XVIII: Equipment Cost Summary for the Lipid Extraction Process

Designation	Equipment Description	Purchase Cost	Bare Module Factor	Bare Module Cost
ORIGINOIL EXTRACTION UNIT				
E-101	Single-Step Extractor 1	500,000	2	1,030,000
GRAVITY CLARIFIERS				
G-101	Gravity Clarifier 1	1,737,000	2	3,579,000
G-102	Gravity Clarifier 2	1,737,000	2	3,579,000
CENTRIFUGES				
C-101	Continuous Scroll Bowl 1	380,000	2	770,400
C-102	Continuous Scroll Bowl 2	380,000	2	770,400
C-103	Continuous Scroll Bowl 3	380,000	2	770,400
C-104	Continuous Scroll Bowl 4	380,000	2	770,400
C-105	Continuous Scroll Bowl 5	380,000	2	770,400
THERMAL DRYERS				
D-101	Drum Dryer 1	334,300	2	688,500
D-102	Drum Dryer 2	334,300	2	688,500
D-103	Drum Dryer 3	334,300	2	688,500
D-104	Drum Dryer 4	334,300	2	688,500
D-105	Drum Dryer 5	334,300	2	688,500
D-106	Drum Dryer 6	334,300	2	688,500
D-107	Drum Dryer 7	334,300	2	688,500
D-108	Drum Dryer 8	334,300	2	688,500
D-109	Drum Dryer 9	334,300	2	688,500

D-110	Drum Dryer 10	334,300	2	688,500
D-111	Drum Dryer 11	334,300	2	688,500
D-112	Drum Dryer 12	334,300	2	688,500
D-113	Drum Dryer 13	334,300	2	688,500
D-114	Drum Dryer 14	334,300	2	688,500
D-115	Drum Dryer 15	334,300	2	688,500
D-116	Drum Dryer 16	334,300	2	688,500
D-117	Drum Dryer 17	334,300	2	688,500
STORAGE TANKS				
S-101	Storage Tank 1	1,575,000	3	4,803,700
S-102	Storage Tank 2	260,600	3	794,800
PUMPS				
P-101	Centrifugal Pump 1	511,000	3	1,686,300
P-102	Centrifugal Pump 2	126,500	3	417,200
P-103	Centrifugal Pump 3	10,000	3	32,800
P-104	Centrifugal Pump 4	424,200	3	1,399,800
Spare Pumps	Centrifugal Pumps	1,071,700	3	3,536,100

Total PM ($)	**8,654,800**	**19,092,600**
Total FM ($)	**3,974,000**	**8,188,000**
Total Spares ($)	**1,071,700**	**3,536,100**
Total Storage ($)	**1,835,600**	**5,598,500**

Table XIX: Fixed Capital Investment Summary for the Lipid Extraction Process

$\mathbf{C_{TBM}}$	**Total Bare Module Cost**		
C_{FM}	Equipment Bare Module Costs	$	8,188,000
C_{PM}	Equipment Bare Module Costs	$	19,092,600
C_{spare}	Total Bare-Module Costs for Spares	$	3,536,100
$C_{storage}$	Total Bare-Module Costs for Storage Tanks	$	5,598,500
	$\mathbf{C_{TBM}}$	$	36,415,200

$\mathbf{C_{DPI}}$	**Direct Permanent Investment**		
C_{TBM}	Total Bare-Module Cost	$	36,415,200
C_{site}	Site Preparation	$	1,821,000
C_{serv}	Service Facilities	$	1,821,000
C_{alloc}	Allocated Costs for Utility Plants	$	0
	$\mathbf{C_{DPI}}$	$	40,057,200

C_{TDC}	**Total Depreciable Capital**		
C_{DPI}	Direct Permanent Investment	$	40,057,200
C_{cont}	Contingencies and Contractor's Fees	$	7,211,000
	C_{TDC}	$	47,268,200

C_{TPI}	**Total Permanent Investment**		
C_{TDC}	Total Depreciable Capital	$	47,268,200
C_{land}	Cost of Land	$	945,364
C_{royal}	Cost of Royalties	$	945,364
$C_{startup}$	Cost of Plant Startup	$	4,726,820
	C_{TPI}	$	53,885,748

C_{TCI}	**Total Capital Investment**		
C_{TPI}	Total Permanent Investment	$	53,885,748
	C_{TCI}	**$**	**53,885,748**

The fixed capital investment summary is based on several key assumptions. According to the *Product and Process Design* text, 1 operator is needed for fluids processing, 2 operators are required for solids-fluid processing, and 3 operators are necessary to monitor solids processing.[67] The three main processes in lipid extraction are extraction, solids separation, and dewatering, which involve fluids processing, solid-fluid processing solids processing respectively. Therefore, based on this heuristic, as well as the principle that large continuous-flow processing operations require double the number of operators, a staff of 12 operators per shift may be estimated for lipid-extraction. OriginOil™ does not divulge the costs involved with the licensing of their product. For the purposes of the fixed capital investment analysis, a cost of royalties of 2% of the total depreciable capital was assumed, a figure based on a recommendation from *Product and Process Design Principles.* [67]

B. Operating Costs

The objective of this analysis is to determine the operating costs associated with the lipid-extraction module, which will be essential for conducting an overall economic investigation of biodiesel production. The fixed costs associated with lipid-extraction are presented in table XX, while table XXI shows the overall cost of manufacture.

Table XX: Fixed Costs Summary of the Lipid Extraction Process

Operations		
Direct Wages and Benefits	$	3,679,200
Direct Salaries and Benefits	$	551,880
Operating Supplies and Services	$	220,752
Technical assistance to manufacturing	$	5,000
Control laboratory	$	5,417
	$	**4,463,000**
Maintenance		
Wages and Benefits	$	1,655,000
Salaries and Benefits	$	414,000
Materials and Services	$	1,655,000
Maintenance Overhead	$	83,000
	$	**3,807,000**
Operating Overhead		
General Plant Overhead	$	448,000
Mechanical Department Services	$	152,000
Employee Relations Department	$	372,000
Business Services	$	467,000
	$	**1,439,000**
Property Taxes and Insurance	**$**	**945,324**
Total Fixed Costs	**$**	**10,654,364**

Table XXI: Cost of Manufacture Summary for the Lipid Extraction Process

Cost of Manufacture			
Utilities		$	106,638,680
Operations		$	4,463,000
Maintenance		$	3,807,000
Operating Overhead		$	1,439,000
Property Taxes and Insurance		$	945,364
Depreciation			
	Direct Plant	$	3,781,456
	Allocated Plant	$	0
	TOTAL	**$**	**121,075,000**

Fixed costs are expenses that are charged regardless of production rate, such as operating costs, maintenance costs, operating overhead, property taxes and insurance. These expenses may be estimated from the C_{TDC} and the number of process sections using the guidelines described in section 23.2 *of Product and Design Principles*.[67] Operating costs were calculated based on a total of 12 operators per shift, while maintenance costs include wages and benefits for employees that maintain equipment in proper working order as well as providing for trained supervisors, materials and services and maintenance overhead. The operating overhead includes costs not directly related to plant operating, such as medical and safety services, and may be estimated as a fraction of the combined salary, wages and benefits for maintenance and labor-related operations.[67] Together, the total fixed costs sum to $9,481,000 as shown in table XXI.

One way to analyze the economic feasibility of the lipid-extraction process is to calculate the costs of producing one kilogram of lipid. This calculation is performed by dividing the cost of manufacture, presented in table XXII, by the mass of lipids produced as shown below.

$$Cost\ of\ Lipid\ Production = \frac{Annual\ Cost}{Annual\ Production} = \frac{\$121{,}075{,}000}{1{,}959{,}426{,}000\ lbs\ of\ lipid}$$

$$Cost\ of\ Lipid\ Production = \frac{\$0.06}{lb}$$

OriginOil™ estimates that the cost of lipids-extraction via conventional means is on the order of $0.56/lb[70]. Therefore, the cost of lipid production via Single-Step Extraction™ technology, estimated at $0.06/lb as shown above, indicates the potential for significant savings. Furthermore, if the biomass byproduct is assumed to sell at a price equivalent to soybean meal, the process might be profitable as shown in the calculations below.

$$Revenue\ from\ Lipid\ Production = \frac{Profit - Cost}{Production} = \frac{\$355{,}599{,}000 - \$121{,}075{,}000}{1{,}959{,}426{,}000\ lbs}$$

$$Revenue\ from\ Lipid\ Production = \frac{\$0.12}{lb}$$

C. Other Important Considerations

Byproduct Sales

Once the biomass has been dewatered to 10% moisture content, it may be sold as a byproduct of lipid-extraction. Although there are several potential uses for an algal biomass product, this report considers the distribution of algal biomass as an ingredient of animal feed. Although there are several reports that indicate that algae may be a suitable animal feedstock, a widespread market for algal biomass consumption has not yet been established. Therefore, in order to determine the value biomass from an economic perspective, the prices for other biomass byproducts of lipid-extraction were considered. Table XXII indicates the 2007 prices associated with the biomass byproducts of soybean, cotton, canola and sunflower lipid extraction according to the US Energy Information Administration.[82]

Table XXII: Estimated prices on the byproducts of different biodiesel producing feed[82]

Lipid-Extraction feedstock	Byproduct	$/ LB	Potential Profit (Annual)
Soybeans	Meal (44-48% protein)	0.097	355,599,000
Cotton	Meal (41% protein)	0.088	322,605,000
Canola	Meal (28-36% protein)	0.079	289,611,000
Sunflower	Meal (28% protein)	0.035	128,309,000

As shown in table XXII, biomass byproducts with higher protein contents sell for higher prices. Since the algal biomass produced features a protein content of 50%, even greater than that of the soybean biomass byproduct, it would likely sell for more than 9.7 cents/lb. However, considering the underdeveloped market for algae as an animal feedstock and accounting for an initial period of skepticism during which the safety of algal animal feed is vetted, it would be conservative and appropriate to estimate the algal biomass price at a value equivalent to that of soy meal.

The lipid-extraction process produces 3.7 billion pounds of biomass each year. At a price of $0.097/lb, the sale of biomass is expected to generate approximately $360 million in annual revenue. According to the Center of Livable Futures at Johns Hopkins University[83], the US produced 240 billion pounds of primary animal feedstock in the year 2004. Since the algal biomass generated in module II represents just 1.5% of the current market size, oversaturation of the animal feedstock market is not expected. For the purposes of this design report, it is assumed that all biomass byproduct will be purchased by consumers. Should this not be the case, it would be prudent to account for a storage facility to store the generated algal biomass until demand for its sale increases and its safety as an animal feedstock has been established.

VI. MODULE III: TRANSESTERIFICATION

1. Transesterification: Concept Stage

A. Preliminary Process Synthesis

Direct combustion of lipids and oils has proven to result in nozzle choking, engine deposits, and incomplete combustion in addition to other engine failures. Dilution of lipids in petrodiesel has been shown produce similarly problematic fuels.[84] Therefore, the lipids successfully extracted from algal cells in module II must undergo chemical processing before they may be considered a usable and valuable fuel. Several chemical processes have been developed to transform lipids into biofuel, including pyrolysis, which involves the heating and decomposition of a feedstock in the absence of oxygen, and microemulsification, which involves the solublization of oils in low molecular weight alcohols. Yet, as global interest in the alternative fuel industry mounts, these processes and others have been overshadowed by focus on catalytic transesterifcation and catalytic hydrotreating, two lipid-processing techniques that show high-promise for revolutionizing the transportation fuel industry.[84] As a result, these two alternatives, which are described in detail in this section, were thoroughly explored during process synthesis.

Catalytic Transesterification

Catalytic transesterification involves the reaction of lipids with an alcohol in the presence of a catalyst to produce fatty acid alkyl esters, commonly known as biodiesel fuel, as shown in equation (1). The chemical structure of the alkyl ester product depends on the identity of the alcohol. Methanol, low in price, is the most typically used alcoholic reagent and results in the production of fatty acid methyl esters (FAME). However, in countries where ethanol is inexpensive and readily available, such as Brazil, the products of transesterification are commonly fatty acid ethyl esters.[84]

$$\text{Triglyceride} + 3\ \text{Methanol} \rightarrow 3\ \text{FAME} + \text{Glycerol} \qquad \textit{EQN.}1$$

The large majority of biodiesel producers employ base catalysis, commonly selecting potassium or sodium hydroxide to accelerate the transesterification reaction. The byproduct of transesterification is glycerol, a simple polyol compound commonly used in gums and resins as well as various pharmaceutical and food industry applications.[85] The crude glycerol formed during lipid processing may be refined to meet technical standards for industrial use. However, since glycerol purification is a challenging process that is profitable only at a large scale, it is rarely performed within the battery limits of biodiesel production facilities. Instead, crude glycerol is typically distributed to separate refining plants.

Some lipid mixtures, such as frying oil, may contain free fatty acids (FFA) in addition to triglycerides. These acids may react with a basic catalyst as shown in equation (2) to form soap and water. This reaction consumes catalyst and generates a soapy product that inhibits the separation of the glycerol byproduct from biodiesel. Furthermore, the reaction generates water, which hydrolyzes the FAME product, decreasing its yield.[84]

$$\text{R-COOH} + \text{Basic Catalyst} \rightarrow \text{R-COOK} + H_2O \qquad \textit{EQN.}2$$

To eliminate this side reaction, oils are commonly pretreated with an acid catalyst such as sulfuric acid and methanol. This process esterifies FFA to methyl esters as shown in equation (3) without affecting the other components of the reaction mixture. Then, the remaining oil, free of FFA and consisting of only of triglycerides, may be transesterified with a basic catalyst to produce biodiesel.[84]

$$\text{R-COOH} + CH_3OH \rightarrow \text{R-COOCH}_3 + H_2O \qquad \textit{EQN.}3$$

Since oil from *Chlorella protothecoides*, the algal strain cultivated in Module I, does not have a high FFA content, this side reaction is not expected to impact biodiesel production. Therefore, the acid pretreatment step is omitted in the process design.

Catalytic Hydrotreating

Prior to the rise of the biofuels industry, catalytic hydrotreating was already a standard chemical process used to remove heteroatoms such as sulfur and oxygen from conventional fuels in petroleum refinement. This technology, involving the reaction of a feedstock with hydrogen in a fixed-bed catalytic reactor, has been applied to lipid processing and used to remove carboxyl and carbonyl groups as well as double bonds from triglyceridges.[84] At elevated temperature and pressure, Nickel molybdenum on alumina typically catalyzes three active reaction pathways.

Together, hydrodeoxygenation, decarboxylation and decarbonylation, shown in equations (4), (5), and (6) respectively, result in the production of a mixture of straight chain alkanes, carbon monoxide, carbon dioxide, water, and propane. The n-alkane products, ranging in length from 15 to 18 carbons, are suitable for use as a conventional diesel for or may be upgraded to jet fuel, gasoline or other fuels.[86]

$$\text{Triglyceride} + \text{Hydrogen} \rightarrow \text{Propane} + \text{Water} + \text{n-alkane} \qquad \textit{EQN.}4$$

$$\text{Triglyceride} + \text{Hydrogen} \rightarrow \text{Propane} + CO_2 + \text{n-alkane} \qquad \textit{EQN.}5$$

$$\text{Triglyceride} + \text{Hydrogen} \rightarrow \text{Propane} + \text{water} + \text{CO} + \text{n-alkane} \qquad \textit{EQN.}6$$

Carbon dioxide, carbon monoxide and water are formed as byproducts of catalytic hyrotreating. In the presence of excess hydrogen, theses components participate in both methanation and the water-gas shift reaction as shown in equations (7) and (8). These side reactions consume hydrogen feed and cause methane gas to be an side product of hydrotreating.[6]

$$\text{Carbon Monoxide} + \text{Hydrogen} \rightarrow \text{Methane} + \text{Water} \qquad \textit{EQN.}7$$

$$\text{Carbon Dioxide} + \text{Hydrogen} \rightarrow \text{Carbon Monoxide} + \text{Water} \qquad \textit{EQN.}8$$

Proposed Lipid-Processing Module: Catalytic Transesterification

As all three modules of the algae-to-biofuel were researched simultaneously, the lipid-processing module was designed to operate with the algal cultivation model in a complementary fashion. The "heteroboost" model upon which Module 1 is based, identifies a key fermentative phase, during which algae are supplied an energy source that they may metabolize to increase their productivity and lipid-content. For laboratory-scale feasibility studies, glucose is often supplied as the fermentative energy source. However, glucose is an expensive reagent and from an economic perspective, it would not be an appropriate nutrient to supply on an industrial scale. As thoroughly discussed in Section F, "Design of the Fermentation Tanks," of the Module 1 Concept Stage, Heredia-Arroyo et al. (2010) demonstrated that supplying glycerol during fermentation may enable the growth of algae with a similarly high lipid content as that grown in glucose-rich media.[1, 87] Interestingly, glycerol is the byproduct generated in large volumes during catalytic transesterification, and could be supplied to the algal cultivation site at no cost other than a small initial capital investment for a network of distribution pipelines and control valves. This link between algal cultivation and glycerol production was celebrated as a means of reducing Module 1 operating costs and led to the adoption of catalytic transesterification as the proposed lipid-processing module for the algae-to-biodiesel venture.

B. Facility Design

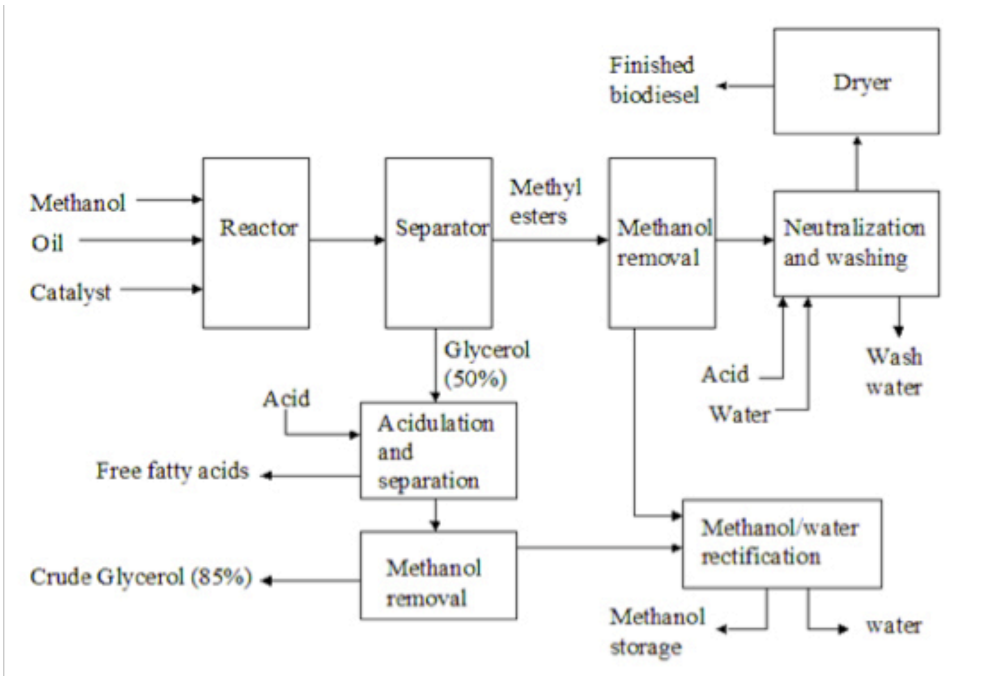


Figure 27. A Typical Industrial Transesterification Route to Biodiesel Fuels[84]

In order to reduce transportation costs and improve efficiency, the biodiesel production facility will be located adjacent to the algal cultivation and oil extraction sites in Thompsons, Texas. As transesterification is distinct from conventional diesel fuel refining, locating the plant near an established oil refinery is not necessary. Since the W.A. Parish Electric Power Generating Station is situated nearby, there will be ample access to electricity and other utilities. The typical industrial transesterification process identified by the *Biodiesel Handbook* is shown above in figure 27.[84] Triglyceride feedstock is initially fed into a reactor along with excess methanol and a homogeneous catalyst. The glycerol byproduct is then separated from the FAME product, which passes through a separation train to be refined into biodiesel that meets industrial standards. Excess methanol is recovered and recycled from both the crude FAME and glycerol streams.

The FAME product is then distributed to consumers. Although transportation via barge is ideal considering the large quantities of liquid product that leave the lipid-processing module, this mode of transport is not feasible due to the inland location of the algae-to-biodiesel venture. While trucks may easily access the biodiesel production facility, their use in shipment is equally unrealistic since they are not appropriate for handling the significant volumes of biodiesel produced on a daily basis.[67] While 815,000 gallons of biodiesel fuel leave the lipid-processing module each day, a tank truck can only accommodate a maximum of 7,500 gallons.[67] When considering both the accessibility of the biodiesel production facility and the high biodiesel production rate, the best means of shipment is by rail, which services the Thompsons, Texas region.[31] Fuel may be transported in special tank cars designed for handling liquids, each of which may be up to 34,500 gallons in size.[67] It is important to note that railway shipments are not continuous. Therefore, storage tanks must accommodate several days of biodiesel product in order to account for the low frequency of railway shipments and to provide adequate capacity in the event of an emergency shutdown. Each step in the process flow diagram shown in figure 27 will be comprehensively evaluated in the Section B of "Feasibility and Developmental Stages."

C. Assembly of Database

Algal oil is not uniform in chemical species, but rather comprised of a variety of triglycerides and free fatty acids. Although researchers rarely measure triglyceride composition, gas chromatography is regularly used to analyze the composition of the biodiesel product, which is shown below for the *Chlorella protothecoides* strain grown in Module 1 in table XXIII.

Table XXIII. Composition of FAME produced from *Chlorella protothecoides*[22]

Component	Chemical Formula	Relative FAME Content (%)
Hexadecanoic acid methyl ester	$C_{17}H_{34}O_2$	10.1
Heptadecanoic acid methyl ester	$C_{18}H_{36}O_2$	0.71
9,12-Octadecadienoic acid methyl ester	$C_{19}H_{34}O_2$	18.33
9-Octadecenoic acid methyl ester	$C_{19}H_{36}O_2$	65.75
Octadecanoic acid methyl ester	$C_{19}H_{38}O_2$	2.85
10-Nonadecenoic acid methyl ester	$C_{10}H_{38}O_2$	1.01
11-Eicosenoic acid methyl ester	$C_{21}H_{40}O_2$	0.67
Eicosanoic acid methyl ester	$C_{21}H_{42}O_2$	0.59

As shown in table XXIII, FAME produced from *C. protothecoides* is mostly comprised of 9-Octadecenoic acid methyl ester, 9, 12-Octadeadienoic acid methyl ester, and Hexadecanoic acid methyl ester.[28] Each fatty acid methyl ester is derived from a fatty acid chain that is part of a larger triglyceride molecule. Therefore, one may work backwards from the data presented in table XXVI to determine what the triglyceride composition of the algal oil may have been prior to transesterification.

Although it is possible for a triglyceride molecule to contain a mixture of fatty acid chain species, it was assumed that triglycerides are uniform in fatty acid-chain composition. Since over 94% of the FAME was comprised of 9-Octadecenoic acid methyl ester, 9, 12-Octadeadienoic acid methyl ester, and Hexadecanoic acid methyl ester, only these species were evaluated when the *C. protothecoides* triglyceride composition, shown below in table XXIV, was determined.

Table XXIV. Composition of Triglcyerides extracted from *Chlorella protothecoides*

Component	Chemical Formula	Fatty Acid Chain Species	Relative Triglyceride Content (%)
Triglyceride 1	$C_{57}H_{104}O_6$	$C_{19}H_{36}O_2$	69.8
Triglyceride 2	$C_{57}H_{98}O_6$	$C_{19}H_{34}O_2$	19.5
Triglyceride 3	$C_{51}H_{98}O_6$	$C_{17}H_{34}O_2$	10.7

Although the overall reaction for each triglyceride may be effectively described by equation (1) above, transesterification is actually a reversible process and involves the formation of two intermediate products. A molecule of triglyceride reacts with one molecule of methanol at a time and releases one fatty acid chain as a methyl ester. After the first fatty acid chain is released, the triglyceride has been converted into a diglyceride. After the second fatty acid chain has been released, the diglyceride has been converted into a monoglyceride. These key intermediates must be accounted for in order to generate an accurate model of catalytic transesterification.

It is important to note that since the transesterification reactions are reversible, a molecule of monoglyceride may react with a molecule of FAME to form a diglcyeride, while a diglyceride may react with FAME in the product mixture to produce a triglyceride. A list of the forward and reverse reactions modeled in the Aspen PLUS simulation of transesterification is shown in table XXV. In modeling these reactions, the Aspen PLUS simulation of catalytic transesterification employed the ELECNRTL property method to fully account for the effects of the dissolved ions in the catalyzed reaction mixture.

Table XV. A List of the Reactions that characterize Transesterification

Reaction	Direction
Triglyceride + MeOH → FAME + Diglyceride	Forward
Diglyceride + FAME → MeOH + Triglyceride	Reverse
Diglyceride + MeOH → FAME + Monoglyceride	Forward
Monoglyceride + FAME → Diglyceride + MeOH	Reverse
Monoglyceride + MeOH → FAME + Glycerol	Forward
Glycerol + FAME → Monoglyceride + MeOH	Reverse

The reactions in table XV will be more thoroughly explored in Section B of "Manufacturing and Developmental Stages." The principle chemicals required for transesterification include the triglyceride feedstock, which is delivered at no cost from the lipid-extraction facility; methanol, priced at $460/MT, concentrated potassium hydroxide catalyst solution, which costs $600/ton, and concentrated hydrochloric acid, which is priced at $130/MT.[107,108,109] Many utilities, including high pressure stream, medium pressure steam, cooling water, chilled water, chilled brine, process water, and wastewater treatment, are also key.

Selection of A Catalyst

Although the transesterification reaction may occur without a catalyst, doing so requires operation at high temperature and pressure and may result in low yields. Therefore, homogeneous catalysts, including acids, bases, and metals, are traditionally used to speed the reaction kinetics and reduce energy requirements.[84] The large majority of biodiesel producers employ base catalysis, which is considered to be the most effective homogeneous catalyst. Acid catalysts have proven to result in slow reaction rates, while metal oxides have short half-lives not suitable for commercial use.[84] In industry, sodium methoxide (sodium methylate), sodium hydroxide (caustic soda) and potassium hydroxide (potash) are the most commonly used bases. A base catalyzes transesterification by reacting with the methanol reagent to form an alkoxide ion, which is then activated to attack the ester group of the triglyceride molecule. This results in the release of a fatty acid methyl ester and the formation of a diglyceride anion. A proton is then transferred from the diglyceride anion to the basic catalyst, resulting in its restoration to its initial chemical structure. This reaction scheme occurs three times per triglyceride molecule until one glycerol molecule and three fatty acid methyl esters have been formed.[75]

D. Bench-Scale Laboratory Work

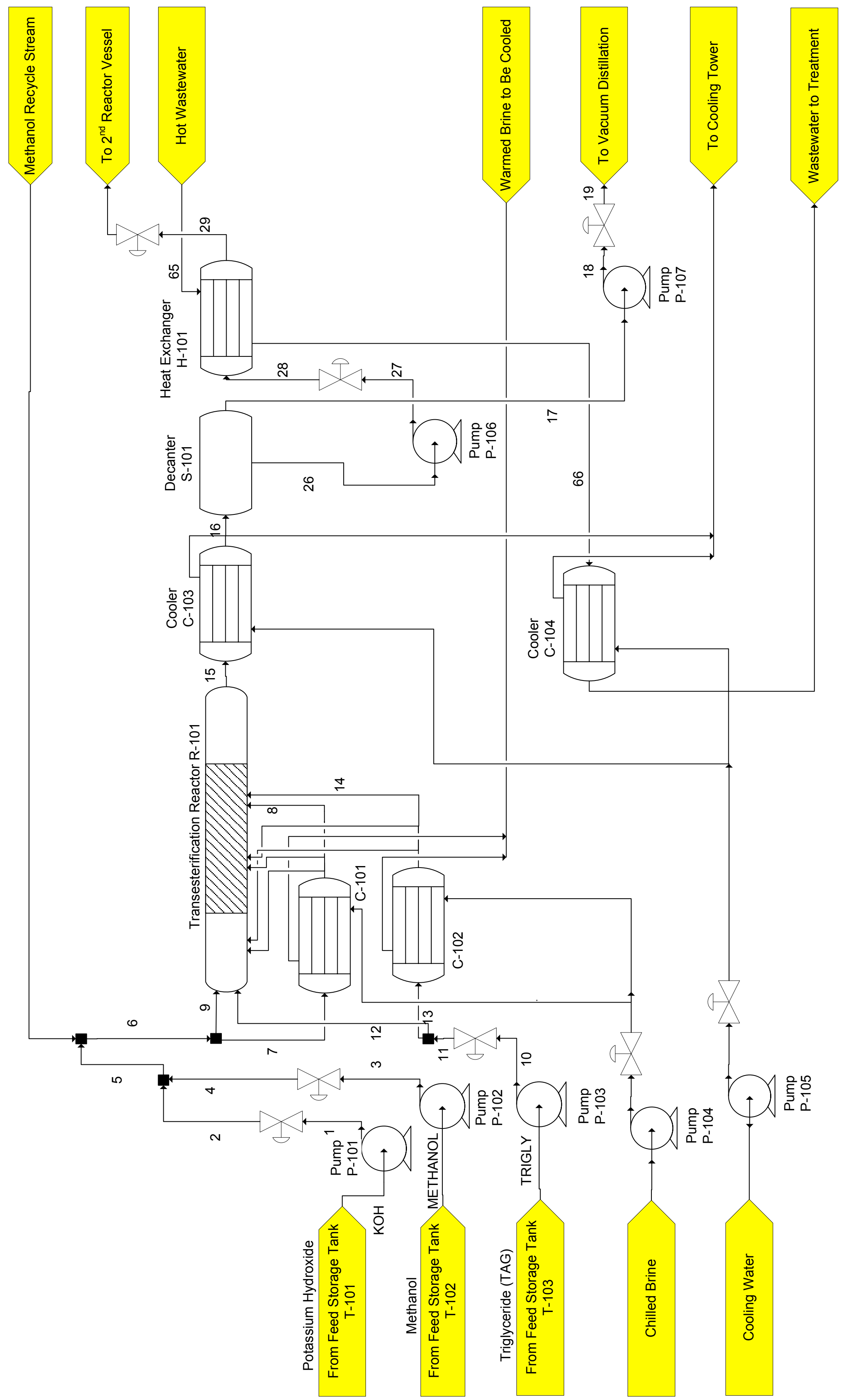
Although the use of transesterification in biodiesel production is well documented in scientific literature, the reaction is often employed for a variety of vegetable oil reagents rather than algal oil specifically derived from *Chlorella protothecoides*. However, many resources were still useful in preparing a model of algal oil transesterification and simulating it using Aspen PLUS software. Y.A. Liu and Ai-Fu Chang wrote an article entitled "Integrated Process Modeling and Product Design of Biodiesel" that proposes a kinetic model for the transesterification of triglycerides and reports reaction rate constants for several vegetable oils at varying temperature and catalyst concentrations.[90] The "Biodiesel Handbook," compiled by editors Gerhard Knothe, Jürgen Krahl, and Jon Van Gerpen proposes a process flow diagram for biodiesel synthesis and describes industrial production in detail.[84] Grace Pokoo-Aikins, Ahmed Nadim, Mahmous M. El-Halwagi, and Vladimir Mahalec have created an ASPEN PLUS simulation of the transesterification process for the algal oil extracted from *Chlorella sp.* and performed a comprehensive economic evaluation of their design.[91]

2. Transesterification: Manufacturing & Development Stages

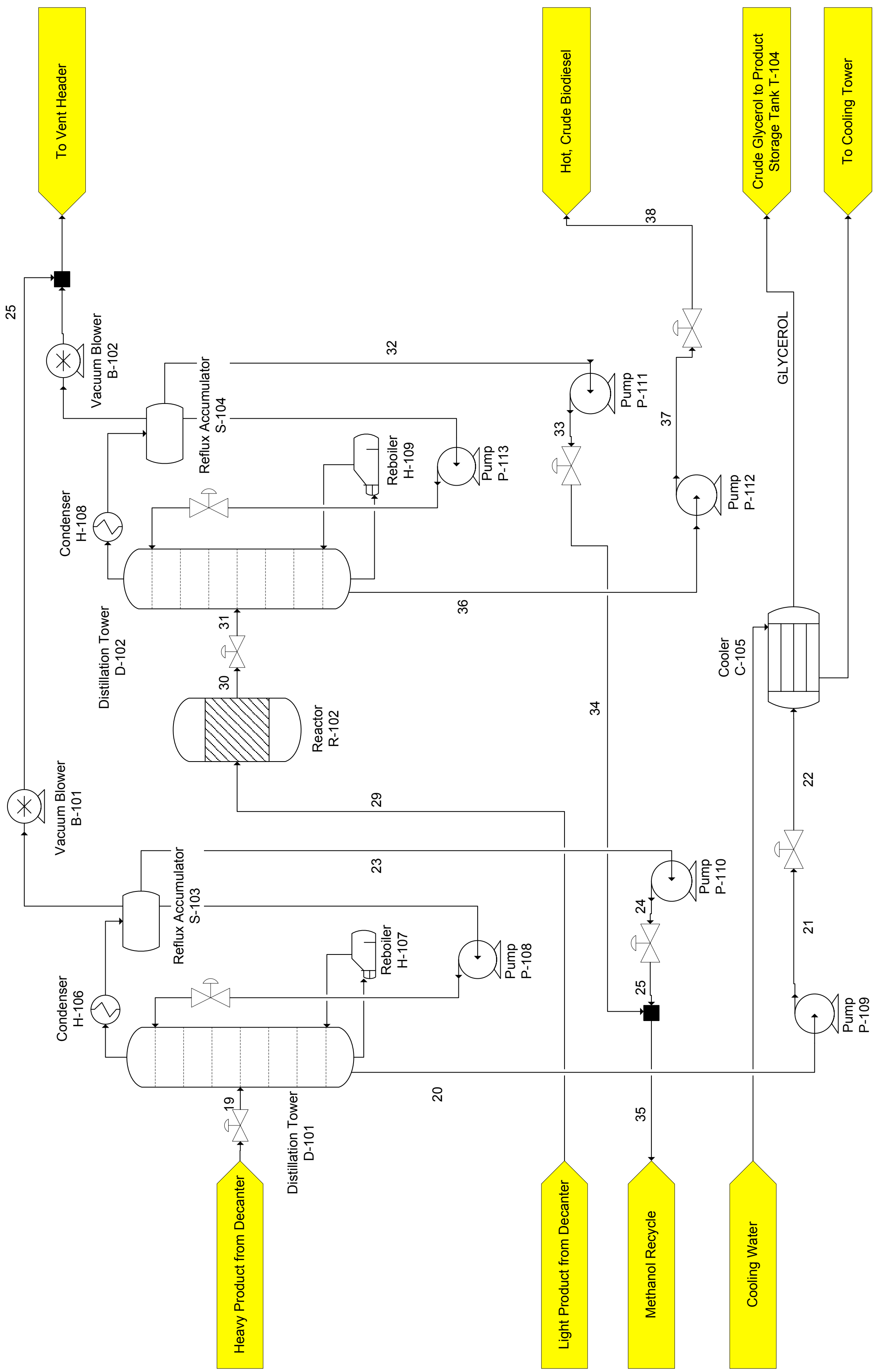
The following sections will describe the process used to transesterify algal oil extracted using Origin Oil™ technology to produce ASTM D6751 grade biodiesel fuel. Section A contains the process flow diagram (PFD) and material balances, while the process description is explained in Sections B. Section C presents the energy balance, along with the associated utility requirements. A list of process equipment along with the corresponding specification sheets and equipment cost summary may be found in Sections D, E, and F respectively. Sections G, H, and I contain the fixed-capital investment summary, other important considerations and the overall operating cost and economic analysis.

A. Process Flow Diagram and the Material Balance

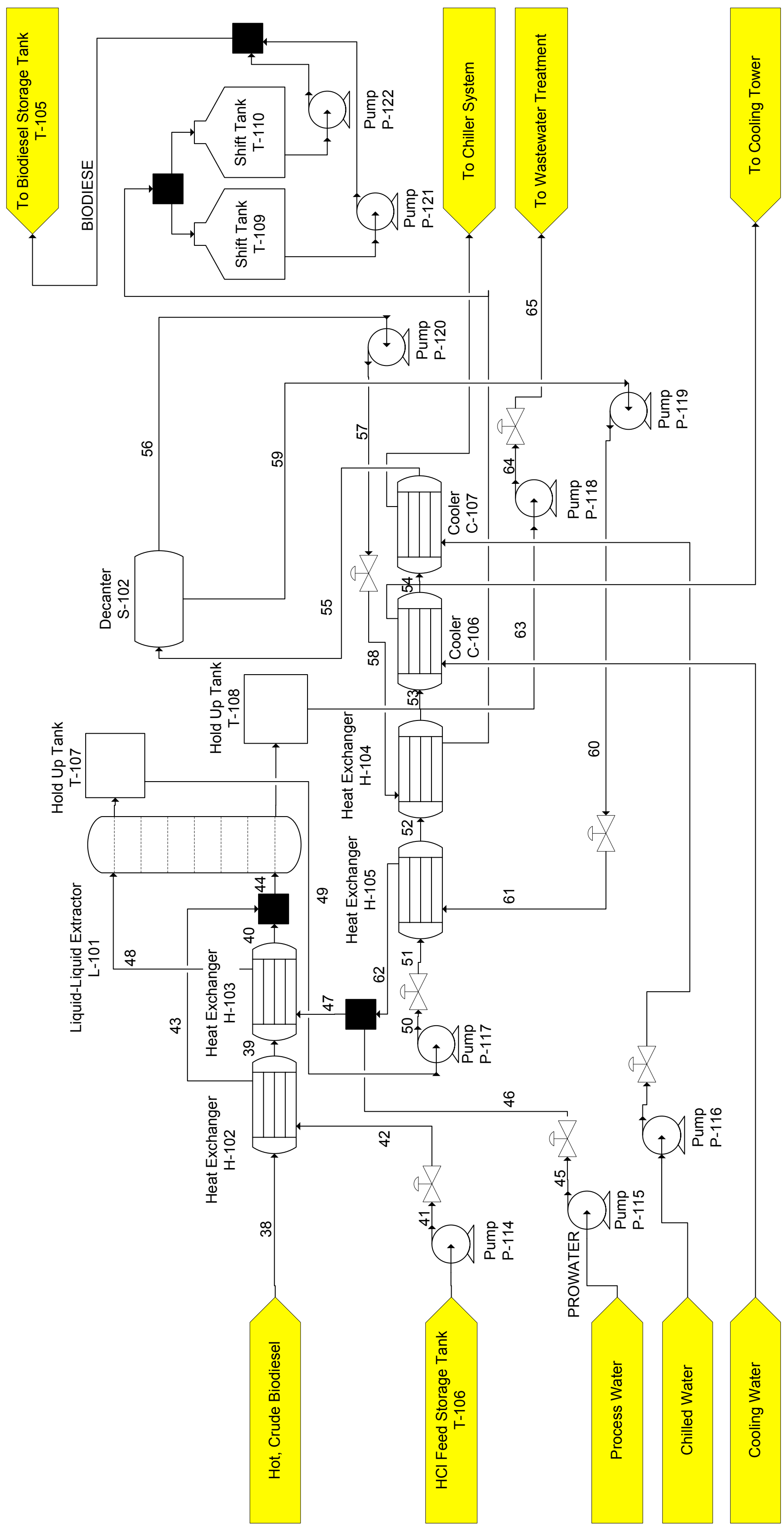
Process Flow Diagram Section One



Process Flow Diagram Section Two



<u>Process Flow Diagram Section Three</u>



		19	20	21	22	23	24	25	26	27
From		V-105	D-101	P-109	V-112	D-101	P-110	V-107	S-101	P-106
To		D-101	P-109	V-112	C-105	P-110	V-107	M4	P-106	V-104
Substream: MIXED										
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Component Mass Flow										
TRIGLY-1	LB/HR	1.76E-04	9.41E-22	0	0	1.76E-04	1.76E-04	1.76E-04	5.309425	5.309425
TRIGLY-2	LB/HR	7.48E-06	2.48E-17	0	0	7.48E-06	7.48E-06	7.48E-06	1.480172	1.480172
TRIGLY-3	LB/HR	2.47E-06	6.15E-19	0	0	2.47E-06	2.47E-06	2.47E-06	0.815264	0.815264
DIGLY-1	LB/HR	0.280394	0.280394	0.280394	0.280394	3.48E-18	0	0	1086.161	1086.161
DIGLY-2	LB/HR	0.151325	0.151325	0.151325	0.151325	3.14E-25	0	0	302.8235	302.8235
DIGLY-3	LB/HR	5.56E-06	5.56E-06	5.56E-06	5.56E-06	2.90E-25	0	0	167.6163	167.6163
MONO-1	LB/HR	0	0	0	0	0	0	0	923.1241	923.1241
MONO-2	LB/HR	0	0	0	0	0	0	0	257.648	257.648
MONO-3	LB/HR	0	0	0	0	0	0	0	144.1034	144.1034
METHANOL	LB/HR	15845.44	49.72337	49.72337	49.72337	15795.72	15795.72	15795.72	8813.413	8813.413
K+	LB/HR	11.98411	11.98411	11.98411	11.98411	1.99E-26	0	0	689.8871	689.8871
OH-	LB/HR	5.213199	5.213199	5.213199	5.213199	1.43E-07	1.45E-07	1.45E-07	300.1074	300.1074
WATER	LB/HR	1188.024	889.9948	889.9948	889.9948	298.0289	298.0289	298.0289	41.4943	41.4943
HCL	LB/HR	0	0	0	0	0	0	0	0	0
CL-	LB/HR	0	0	0	0	0	0	0	0	0
H3O+	LB/HR	6.17E-15	1.28E-11	1.25E-17	1.25E-17	1.60E-07	1.62E-07	1.62E-07	8.39E-14	2.34E-21
FAME-1	LB/HR	0	0	0	0	0	0	0	1.55E+05	1.55E+05
FAME-2	LB/HR	0	0	0	0	0	0	0	43229.58	43229.58
FAME-3	LB/HR	0	0	0	0	0	0	0	23820.08	23820.08
GLYCEROL	LB/HR	22954.85	22954.85	22954.85	22954.85	1.05E-13	0	0	2.312088	2.312088
Mass Flow	LB/HR	40005.95	23912.2	23912.2	23912.2	16093.75	16093.75	16093.75	2.35E+05	2.35E+05
Temperature	F	100	334.7182	334.9849	334.9849	123.1999	123.8344	123.8871	100	-13.7028
Pressure	PSIA	17.7	9.88	48.88	23.88	8	79.7	54.7	42.7	71.7
Molar Enthalpy	BTU/LBMOL	-1.62E+05	-2.49E+05	-2.49E+05	-2.49E+05	-1.02E+05	-1.02E+05	-1.02E+05	-2.45E+05	-2.45E+05
Mass Enthalpy	BTU/LB	-3273.36	-3131.05	-3130.8	-3130.8	-3239.94	-3239.26	-3239.26	-1123.83	-1123.69
Enthalpy Flow	BTU/HR	-1.31E+08	-7.49E+07	-7.49E+07	-7.49E+07	-5.21E+07	-5.21E+07	-5.21E+07	-2.64E+08	-2.64E+08
Mass Density	LB/CUFT	64.2135	68.40657	68.39599	68.39599	47.28657	47.25974	47.25744	49.58	53.00518
Average Molecular Weight		49.36993	79.48995	79.48995	79.48995	31.58673	31.58673	31.58673	218.0926	218.0926

		28	29	30	31	32	33	34	35	R35
From		V-104	H-101	R-102	V-106	D-102	P-111	V-108	M4	
To		H-101	R-102	V-106	D-102	P-111	V-108	M4		M2
Substream: MIXED										
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Component Mass Flow										
TRIGLY-1	LB/HR	5.309425	5.309425	1.81E-04	1.81E-04	3.88E-05	3.88E-05	3.88E-05	2.15E-04	0
TRIGLY-2	LB/HR	1.480172	1.480172	5.03E-05	5.03E-05	6.67E-06	6.67E-06	6.67E-06	1.42E-05	0
TRIGLY-3	LB/HR	0.815264	0.815264	2.77E-05	2.77E-05	4.18E-06	4.18E-06	4.18E-06	6.65E-06	0
DIGLY-1	LB/HR	1086.161	1086.161	10.81165	10.81165	3.77E-17	0	0	0	0
DIGLY-2	LB/HR	302.8235	302.8235	3.01431	3.01431	8.51E-21	0	0	0	0
DIGLY-3	LB/HR	167.6163	167.6163	1.668453	1.668453	5.03E-22	0	0	0	0
MONO-1	LB/HR	923.1241	923.1241	22.87408	22.87408	9.31E-25	0	0	0	0
MONO-2	LB/HR	257.648	257.648	6.383639	6.383639	9.30E-25	0	0	0	0
MONO-3	LB/HR	144.1034	144.1034	3.571105	3.571105	8.69E-26	0	0	0	0
METHANOL	LB/HR	8813.413	8813.413	8534.535	8534.535	8413.39	8413.39	8413.39	24209.11	24513.2
K+	LB/HR	689.8871	689.8871	689.8871	689.8871	1.03E-26	0	0	0	0
OH-	LB/HR	300.1074	300.1074	300.1074	300.1074	9.78E-13	2.20E-12	2.48E-09	1.36E-07	0
WATER	LB/HR	41.4943	41.4943	41.4943	41.4943	6.610399	6.610399	6.610399	304.6393	0
HCL	LB/HR	0	0	0	0	0	0	0	0	0
CL-	LB/HR	0	0	0	0	0	0	0	0	0
H3O+	LB/HR	1.56E-18	8.51E-18	9.93E-18	9.99E-18	1.09E-12	2.46E-12	2.77E-09	1.52E-07	0
FAME-1	LB/HR	1.55E+05	1.55E+05	1.57E+05	1.57E+05	4.04E-12	0	0	0	0
FAME-2	LB/HR	43229.58	43229.58	43725.97	43725.97	4.35E-13	0	0	0	0
FAME-3	LB/HR	23820.08	23820.08	24093.67	24093.67	4.49E-10	0	0	0	0
GLYCEROL	LB/HR	2.312088	2.312088	571.1636	571.1636	8.24E-15	0	0	0	0
Mass Flow	LB/HR	2.35E+05	2.35E+05	2.35E+05	2.35E+05	8420	8420	8420	24513.75	24513.2
Temperature	F	100	140	143.7277	143.9038	122.3769	123.4043	124.2636	124.0273	77
Pressure	PSIA	46.7	39.7	34.7	9.7	8	79.7	54.7	49.7	79.65204
Molar Enthalpy	BTU/LBMOL	-2.38E+05	-2.35E+05	-2.35E+05	-2.35E+05	-1.02E+05	-1.02E+05	-1.02E+05	-1.02E+05	-1.03E+05
Mass Enthalpy	BTU/LB	-1089.5	-1075.46	-1075.46	-1075.46	-3176.11	-3175.21	-3175.21	-3217.26	-3211.749
Enthalpy Flow	BTU/HR	-2.56E+08	-2.53E+08	-2.53E+08	-2.53E+08	-2.67E+07	-2.67E+07	-2.67E+07	-7.89E+07	-7.87E+07
Mass Density	LB/CUFT	50.75902	49.94818	50.47362	50.46994	47.12338	47.07948	47.04272	47.18305	49.00393
Average Molecular Weight		218.0926	218.0926	218.0926	218.0926	32.02259	32.02259	32.02259	31.73509	32.04216

		36	37	38	39	40	41	42	43	44
From		D-102	P-112	V-109	H-102	H-103	P-114	V-110	H-102	M5
To		P-112	V-109	H-102	H-103	M5	V-110	H-102	M5	L-101
Substream: MIXED										
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Mixed	Liquid	Liquid
Component Mass Flow										
TRIGLY-1	LB/HR	1.42E-04	1.42E-04	1.42E-04	1.42E-04	1.42E-04	0	0	0	1.42E-04
TRIGLY-2	LB/HR	4.37E-05	4.37E-05	4.37E-05	4.37E-05	4.37E-05	0	0	0	4.37E-05
TRIGLY-3	LB/HR	2.35E-05	2.35E-05	2.35E-05	2.35E-05	2.35E-05	0	0	0	2.35E-05
DIGLY-1	LB/HR	10.81165	10.81165	10.81165	10.81165	10.81165	0	0	0	10.81165
DIGLY-2	LB/HR	3.01431	3.01431	3.01431	3.01431	3.01431	0	0	0	3.01431
DIGLY-3	LB/HR	1.668453	1.668453	1.668453	1.668453	1.668453	0	0	0	1.668453
MONO-1	LB/HR	22.87408	22.87408	22.87408	22.87408	22.87408	0	0	0	22.87408
MONO-2	LB/HR	6.383639	6.383639	6.383639	6.383639	6.383639	0	0	0	6.383639
MONO-3	LB/HR	3.571105	3.571105	3.571105	3.571105	3.571105	0	0	0	3.571105
METHANOL	LB/HR	121.1454	121.1454	121.1454	121.1454	121.1454	0	0	0	121.1454
K+	LB/HR	689.8871	689.8871	689.8871	689.8871	689.8871	0	0	0	689.8871
OH-	LB/HR	300.1074	300.1074	300.1074	300.1074	300.1074	1.13E-17	1.39E-12	4.19E-16	2.32E-16
WATER	LB/HR	34.8839	34.8839	34.8839	34.8839	34.8839	787.3039	1057.262	729.4546	1403.168
HCL	LB/HR	0	0	0	0	0	117.8712	664.2316	0.791476	6.990962
CL-	LB/HR	0	0	0	0	0	552.8729	21.60826	666.7178	660.6896
H3O+	LB/HR	8.66E-19	8.74E-19	8.74E-19	8.30E-19	1.29E-21	296.6475	11.59405	357.7318	18.83874
FAME-1	LB/HR	1.57E+05	1.57E+05	1.57E+05	1.57E+05	1.57E+05	0	0	0	1.57E+05
FAME-2	LB/HR	43725.97	43725.97	43725.97	43725.97	43725.97	0	0	0	43725.97
FAME-3	LB/HR	24093.67	24093.67	24093.67	24093.67	24093.67	0	0	0	24093.67
GLYCEROL	LB/HR	571.1636	571.1636	571.1636	571.1636	571.1636	0	0	0	571.1636
Mass Flow	LB/HR	2.26E+05	2.26E+05	2.26E+05	2.26E+05	2.26E+05	1754.696	1754.696	1754.696	2.28E+05
Temperature	F	428.2219	428.8306	428.8306	425.4596	156.83	76.73	76.73	156.83	165.5326
Pressure	PSIA	9.88	66.73	41.73	34.73	27.73	62.7	37.7	30.7	22.73
Molar Enthalpy	BTU/LBMOL	-2.44E+05	-2.44E+05	-2.44E+05	-2.45E+05	-2.77E+05	-1.09E+05	-1.03E+05	-1.08E+05	-2.62E+05
Mass Enthalpy	BTU/LB	-877.797	-877.489	-877.489	-879.192	-993.97	-4868.96	-4591.36	-4811.09	-1023.33
Enthalpy Flow	BTU/HR	-1.99E+08	-1.99E+08	-1.99E+08	-1.99E+08	-2.25E+08	-8.54E+06	-8.06E+06	-8.44E+06	-2.34E+08
Mass Density	LB/CUFT	46.80712	46.79216	46.79216	46.87494	53.06467	70.9943	0.627994	72.87197	52.91721
Average Molecular Weight		278.2043	278.2043	278.2043	278.2043	278.2043	22.46048	22.46048	22.46048	255.8063

		45	46	47	48	49	50	51	52	53
From		P-115	V-111	M6	H-103	L-101	P-117	V-116	H-105	H-104
To		V-111	M6	H-103	L-101	P-117	V-116	H-105	H-104	C-106
Substream: MIXED										
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Component Mass Flow										
TRIGLY-1	LB/HR	0	0	0	0	5.15E-05	5.15E-05	5.15E-05	5.15E-05	5.15E-05
TRIGLY-2	LB/HR	0	0	0	0	2.56E-05	2.56E-05	2.56E-05	2.56E-05	2.56E-05
TRIGLY-3	LB/HR	0	0	0	0	9.54E-06	9.54E-06	9.54E-06	9.54E-06	9.54E-06
DIGLY-1	LB/HR	0	0	0	0	1.125944	1.125944	1.125944	1.125944	1.125944
DIGLY-2	LB/HR	0	0	0	0	0.313915	0.313915	0.313915	0.313915	0.313915
DIGLY-3	LB/HR	0	0	0	0	0.173756	0.173756	0.173756	0.173756	0.173756
MONO-1	LB/HR	0	0	0	0	2.382146	2.382146	2.382146	2.382146	2.382146
MONO-2	LB/HR	0	0	0	0	0.664803	0.664803	0.664803	0.664803	0.664803
MONO-3	LB/HR	0	0	0	0	0.371901	0.371901	0.371901	0.371901	0.371901
METHANOL	LB/HR	0	0	0	0	4.17E-08	4.17E-08	4.17E-08	4.17E-08	4.17E-08
K+	LB/HR	0	0	0	0	3.07E-04	3.07E-04	3.07E-04	3.07E-04	3.07E-04
OH-	LB/HR	3.50E-04	3.50E-04	5.73E-04	2.30E-03	8.73E-04	0.03468	0.034734	0.022375	1.31E-03
WATER	LB/HR	2.05E+05	2.05E+05	3.09E+05	3.09E+05	1.03E+05	1.03E+05	1.03E+05	1.03E+05	1.03E+05
HCL	LB/HR	0	0	0	0	0.724522	1.50E-15	1.50E-15	9.80E-16	1.50E-13
CL-	LB/HR	0	0	0	0	2.94E-04	0.704798	0.704798	0.704798	0.704798
H3O+	LB/HR	3.91E-04	3.91E-04	6.41E-04	2.57E-03	9.85E-04	0.416803	0.416863	0.40304	0.379477
FAME-1	LB/HR	0	0	0	0	1.57E+05	1.57E+05	1.57E+05	1.57E+05	1.57E+05
FAME-2	LB/HR	0	0	0	0	43724.96	43724.96	43724.96	43724.96	43724.96
FAME-3	LB/HR	0	0	0	0	24093.01	24093.01	24093.01	24093.01	24093.01
GLYCEROL	LB/HR	0	0	0	0	1.52E-08	1.52E-08	1.52E-08	1.52E-08	1.52E-08
Mass Flow	LB/HR	2.05E+05	2.05E+05	3.09E+05	3.09E+05	3.28E+05	3.28E+05	3.28E+05	3.28E+05	3.28E+05
Temperature	F	76.92283	76.92283	81.07401	165.286	165.4	165.6947	165.7802	143.8025	134.7561
Pressure	PSIA	59.69595	34.69595	29.69595	22.69595	14.69595	70.69595	45.69595	38.69595	31.69595
Molar Enthalpy	BTU/LBMOL	-1.23E+05	-1.23E+05	-1.23E+05	-1.21E+05	-1.41E+05	-1.41E+05	-1.41E+05	-1.42E+05	-1.42E+05
Mass Enthalpy	BTU/LB	-6825.45	-6825.45	-6821.31	-6737.14	-2798.42	-2798.17	-2798.17	-2809.49	-2814.13
Enthalpy Flow	BTU/HR	-1.40E+09	-1.40E+09	-2.11E+09	-2.08E+09	-9.18E+08	-9.18E+08	-9.18E+08	-9.22E+08	-9.23E+08
Mass Density	LB/CUFT	62.26054	62.25571	62.21607	60.88886	55.14676	55.14371	55.1406	55.60877	55.79645
Average Molecular Weight		18.01528	18.01528	18.01528	18.01528	50.41035	50.41035	50.41035	50.41035	50.41035

		R62	63	64	65	66	BIODIESE	GLYCEROL	HCL	KOH
From			L-101	P-118	V-114	H-101	H-104	C-105		
To		M6	P-118	V-114	H-101	C-104			P-114	P-101
Substream: MIXED										
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Component Mass Flow										
TRIGLY-1	LB/HR	0	9.02E-05	9.02E-05	9.02E-05	9.02E-05	5.14E-05	0	0	0
TRIGLY-2	LB/HR	0	1.80E-05	1.80E-05	1.80E-05	1.80E-05	2.56E-05	0	0	0
TRIGLY-3	LB/HR	0	1.40E-05	1.40E-05	1.40E-05	1.40E-05	9.54E-06	0	0	0
DIGLY-1	LB/HR	0	9.68234	9.68234	9.68234	9.68234	1.125943	0.2803943	0	0
DIGLY-2	LB/HR	0	2.699455	2.699455	2.699455	2.699455	0.3139151	0.1513248	0	0
DIGLY-3	LB/HR	0	1.494177	1.494177	1.494177	1.494177	0.1737556	5.56E-06	0	0
MONO-1	LB/HR	0	20.48481	20.48481	20.48481	20.48481	2.382146	0	0	0
MONO-2	LB/HR	0	5.716846	5.716846	5.716846	5.716846	0.6648031	0	0	0
MONO-3	LB/HR	0	3.198091	3.198091	3.198091	3.198091	0.371901	0	0	0
METHANOL	LB/HR	0	121.1454	121.1454	121.1454	121.1454	3.99E-09	49.72337	0	0
K+	LB/HR	0	689.8867	689.8867	689.8867	689.8867	3.07E-04	11.98411	0	701.8712
OH-	LB/HR	2.26E-04	1.43E-03	2.40E-07	2.40E-07	1.56E-07	2.25E-18	5.213199	1.15E-17	305.3206
WATER	LB/HR	1.03E+05	2.07E+05	2.07E+05	2.07E+05	2.07E+05	126.3571	889.9948	787.1794	1229.518
HCL	LB/HR	0	6.264271	3.97E-06	3.97E-06	3.99E-06	0.6617162	0	117.6192	0
CL-	LB/HR	0	660.6893	666.7805	666.7805	666.7805	0.0613624	0	553.118	0
H3O+	LB/HR	2.53E-04	18.84033	22.107	22.107	22.107	0.0327751	8.65E-21	296.7791	1.70E-16
FAME-1	LB/HR	0	3.625658	3.625658	3.625658	3.625658	1.57E+05	0	0	0
FAME-2	LB/HR	0	1.007952	1.007952	1.007952	1.007952	43724.96	0	0	0
FAME-3	LB/HR	0	0.663134	0.663134	0.663134	0.663134	24093.01	0	0	0
GLYCEROL	LB/HR	0	571.1636	571.1636	571.1636	571.1636	0	22954.85	0	0
Mass Flow	LB/HR	1.03E+05	2.09E+05	2.09E+05	2.09E+05	2.09E+05	2.25E+05	23912.2	1754.696	2236.71
Temperature	F	89.33	165.4	165.5827	165.5828	149.6472	76.73	100	77	77
Pressure	PSIA	131.9696	14.69595	54.99595	29.99595	22.99595	14.7	16.88	14.7	14.7
Molar Enthalpy	BTU/LBMOL	-1.23E+05	-1.21E+05	-1.21E+05	-1.21E+05	-1.22E+05	-2.92E+05	-2.61E+05	-1.09E+05	-1.16E+05
Mass Enthalpy	BTU/LB	-6813.07	-6695.7	-6695.53	-6695.53	-6711.32	-1004.713	-3277.498	-4868.775	-5384.83
Enthalpy Flow	BTU/HR	-7.04E+08	-1.40E+09	-1.40E+09	-1.40E+09	-1.40E+09	-2.26E+08	-7.84E+07	-8.54E+06	-1.20E+07
Mass Density	LB/CUFT	62.14955	61.14352	61.14867	61.1438	61.46248	54.76641	76.89274	70.98207	90.03089
Average Molecular Weight		18.01528	18.12463	18.12463	18.12463	18.12463	290.578	79.48995	22.46048	21.47543

		54	55	56	57	58	59	60	61	62
From		C-106	C-107	S-102	P-120	V-115	S-102	P-119	V-113	H-105
To		C-107	S-102	P-120	V-115	H-104	P-119	V-113	H-105	
Substream: MIXED										
Phase:		Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid	Liquid
Component Mass Flow										
TRIGLY-1	LB/HR	5.15E-05	5.15E-05	5.14E-05	5.14E-05	5.14E-05	4.41E-08	5.08E-09	0	0
TRIGLY-2	LB/HR	2.56E-05	2.56E-05	2.56E-05	2.56E-05	2.56E-05	2.57E-09	0	0	0
TRIGLY-3	LB/HR	9.54E-06	9.54E-06	9.54E-06	9.54E-06	9.54E-06	5.32E-10	0	0	0
DIGLY-1	LB/HR	1.125944	1.125944	1.125943	1.125943	1.125943	1.42E-06	1.42E-06	1.42E-06	1.42E-06
DIGLY-2	LB/HR	0.313915	0.313915	0.313915	0.313915	0.313915	3.21E-07	3.21E-07	3.21E-07	3.21E-07
DIGLY-3	LB/HR	0.173756	0.173756	0.173756	0.173756	0.173756	2.19E-10	0	0	0
MONO-1	LB/HR	2.382146	2.382146	2.382146	2.382146	2.382146	2.30E-26	0	0	0
MONO-2	LB/HR	0.664803	0.664803	0.664803	0.664803	0.664803	7.86E-27	0	0	0
MONO-3	LB/HR	0.371901	0.371901	0.371901	0.371901	0.371901	6.66E-27	0	0	0
METHANOL	LB/HR	4.17E-08	4.17E-08	3.99E-09	3.99E-09	3.99E-09	3.78E-08	3.78E-08	3.78E-08	3.78E-08
K+	LB/HR	3.07E-04	3.07E-04	3.07E-04	3.07E-04	3.07E-04	8.88E-10	8.88E-10	8.88E-10	8.88E-10
OH-	LB/HR	7.77E-03	1.91E-03	4.88E-10	8.09E-14	8.18E-14	1.41E-15	1.03E-04	1.03E-04	2.26E-04
WATER	LB/HR	1.03E+05	1.03E+05	126.0302	126.0302	126.0302	1.03E+05	1.03E+05	1.03E+05	1.03E+05
HCL	LB/HR	3.64E-16	1.00E-16	2.29E-09	1.03E-04	1.03E-04	8.32E-11	2.24E-19	2.24E-19	4.90E-19
CL-	LB/HR	0.704798	0.704798	0.704796	0.704695	0.704696	2.04E-06	2.04E-06	2.04E-06	2.04E-06
H3O+	LB/HR	0.386702	0.38015	0.378014	0.37796	0.37796	1.09E-06	1.16E-04	1.16E-04	2.54E-04
FAME-1	LB/HR	1.57E+05	1.57E+05	1.57E+05	1.57E+05	1.57E+05	1.02E-18	0	0	0
FAME-2	LB/HR	43724.96	43724.96	43724.96	43724.96	43724.96	2.37E-19	0	0	0
FAME-3	LB/HR	24093.01	24093.01	24093.01	24093.01	24093.01	5.22E-18	0	0	0
GLYCEROL	LB/HR	1.52E-08	1.52E-08	1.81E-14	0	0	1.52E-08	1.52E-08	1.52E-08	1.52E-08
Mass Flow	LB/HR	3.28E+05	3.28E+05	2.25E+05	2.25E+05	2.25E+05	1.03E+05	1.03E+05	1.03E+05	1.03E+05
Temperature	F	100	54.5	54.5	53.33	53.56392	54.5	53.33	53.33	89.33
Pressure	PSIA	24.69595	17.69595	14.69595	46.7	21.7	14.69595	71.7	46.7	39.7
Molar Enthalpy	BTU/LBMOL	-1.43E+05	-1.44E+05	-3.16E+05	-2.94E+05	-2.94E+05	-1.25E+05	-1.23E+05	-1.23E+05	-1.23E+05
Mass Enthalpy	BTU/LB	-2831.45	-2853.54	-1086.05	-1011.49	-1011.49	-6928.21	-6849.03	-6849.03	-6813.07
Enthalpy Flow	BTU/HR	-9.29E+08	-9.36E+08	-2.44E+08	-2.27E+08	-2.27E+08	-7.16E+08	-7.08E+08	-7.08E+08	-7.04E+08
Mass Density	LB/CUFT	56.4892	57.3088	55.0957	55.26174	55.25681	62.73947	62.41609	62.41105	62.13198
Average Molecular Weight		50.41035	50.41035	290.578	290.578	290.578	18.01528	18.01528	18.01528	18.01528

		METHANOL	PROWATER	TRIGLY	WASTE
From					C-104
To		P-102	P-115	P-103	
Substream: MIXED					
Phase:		Liquid	Liquid	Liquid	Liquid
Component Mass Flow					
TRIGLY-1	LB/HR	0	0	1.56E+05	9.02E-05
TRIGLY-2	LB/HR	0	0	43534.7	1.80E-05
TRIGLY-3	LB/HR	0	0	23978.44	1.40E-05
DIGLY-1	LB/HR	0	0	0	9.68234
DIGLY-2	LB/HR	0	0	0	2.699455
DIGLY-3	LB/HR	0	0	0	1.494177
MONO-1	LB/HR	0	0	0	20.48481
MONO-2	LB/HR	0	0	0	5.716846
MONO-3	LB/HR	0	0	0	3.198091
METHANOL	LB/HR	24429.51	0	0	121.1454
K+	LB/HR	0	0	0	689.8867
OH-	LB/HR	0	3.48E-04	0	3.13E-08
WATER	LB/HR	0	2.05E+05	0	2.07E+05
HCL	LB/HR	0	0	0	3.97E-06
CL-	LB/HR	0	0	0	666.7805
H3O+	LB/HR	0	3.89E-04	0	22.107
FAME-1	LB/HR	0	0	0	3.625658
FAME-2	LB/HR	0	0	0	1.007952
FAME-3	LB/HR	0	0	0	0.663134
GLYCEROL	LB/HR	0	0	0	571.1636
Mass Flow	LB/HR	24429.51	2.05E+05	2.24E+05	2.09E+05
Temperature	F	77	76.73	77	100
Pressure	PSIA	14.7	14.69595	14.7	15.99595
Molar Enthalpy	BTU/LBMOL	-1.03E+05	-1.23E+05	-7.65E+05	-1.23E+05
Mass Enthalpy	BTU/LB	-3211.886	-6825.645	-873.864	-6760.4
Enthalpy Flow	BTU/HR	-7.85E+07	-1.40E+09	-1.95E+08	-1.41E+09
Mass Density	LB/CUFT	49.00393	62.25357	3.30426	62.2674
Average Molecular Weight		32.04216	18.01528	875.2	18.12463

B. Process Description

What follows is a comprehensive analysis of the operation and design of the catalytic transesterification process, which may be segmented into the following components: 1) Preparation of Reagents 2) Transesterification Reaction 3) Stream Separation 4) Product Refining. Once the transesterification process was designed, it was simulated using the Aspen PLUS process modeling software. Please see Sections D4, D5 and D6 of the Appendix on pages 288, 289, and 292 for the Aspen PLUS process flow sheet, Aspen PLUS property estimation details and the complete simulation results.

As mentioned in the "Preliminary Process Synthesis" section above, both catalytic hydrotreating and transesterification are cited as promising lipid-processing modules from technical and economic perspectives. Most small diesel hydrotreaters operate at a throughput of roughly 20,000 barrels per day, a rate equivalent to roughly 225,000 lb/hr of diesel production.[8] To accurately compare a catalytic hydrotreating module to a tranesterification module from a financial standpoint, the transesterification process was also designed for a throughput of 225,000 kg/hr. Since FAME is less dense than the n-alkanes produced by catalytic hydrotreating, this mass flow rate translates to a volumetric flow rate of only 17,560 barrels per day of biodiesel fuel.

Reagent Preparation

- ***Triglyceride Feed***

The triglycerides extracted from algae via the OriginOil™ process are stored in a feed storage tank (T-103) and pumped (P-103) into the transesterification process at a mass flow rate of 223,700 lb/hr, pressure of 105 psi and temperature of 77ºF. For the purposes of the ASPEN PLUS simulation, it was assumed that the triglyceride feedstock is pure, with no significant quantity of trace metals, contaminants or leftover components from algae cultivation. A triglyceride consists of three fatty acid chains and a glycerol backbone. As previously discussed in the Section C of the Concept Stage, "Assembly of Database," there is no single triglyceride that includes all of the various fatty acid chains found in the lipids extracted from *Chlorella protothecoides*. Therefore, three triglycerides, each comprised of three identical fatty acid chains were modeled in Aspen PLUS using group contribution property estimation, which involves the breakdown of a large molecule into smaller components with known properties. Since this technique can only approximate the chemical, physical and thermodynamic properties of an uncharacterized species, it is important to note that the Aspen PLUS simulation results are close approximations, but not exact predictions, of the transesterification process. Please see section D5 of the Appendix on page 289 for additional information regarding the breakdown of the triglyceride, diglyceride, monoglyceride and FAME chemical structures into their various chemical groups.

The amount of each triglyceride in the feed stream was set according to the weight percentage reported in table XXVII. One sixth of the triglycerides are sent to the tubular transesterification reactor at room temperature, while the remaining portion loses a significant amount of heat to chilled brine in Cooler C-102 and is decreased in temperature to 10ºF. These triglycerides are then split into three cold shots that enter the tubular reactor at various locations. This scheme prevents the reactor temperature from increasing significantly during transesterificaiton, which is moderately exothermic.

- ***Methanol Feed***

Methanol, stored in a feed storage tank (T-102), is pumped (P-102) at a constant mass flow rate of 24,820 lb/hr, pressure of 105 PSI and temperature of 77ºF. This feed stream is mixed with the catalyst feed stream before being split into various portions. One sixth enters the transesterification reactor at 77ºF, while the remaining amount loses a significant amount of heat to chilled brine in Cooler C-101 and is decreased in temperature to 10ºF. This feed mixture is split into three cold shots that enter the tubular reactor at various locations in order to prevent the temperature from increasing significantly during transesterificaiton, which is moderately exothermic.

- ***Potassium Hydroxide Feed***

A 45% concentrated potassium hydroxide solution is used as a catalyst for the transesterification process. Since the dissolution of solid potassium hydroxide pellets in water is highly exothermic and requires a carefully controlled process design, concentrated potassium hydroxide, which is readily available for purchase, will not be made within battery limits. The 45% potassium hydroxide solution, stored in Tank-101, is pumped (P-101) at a mass flow rate of 2,240 lb/hr, exactly 1% of the triglyceride stream flow rate, and temperature of 77ºF. The catalyst is mixed with the methanol feed stream before being split into various portions and entering the transesterification reactor.

Tranesterification Reaction

In the transesterification reactors (R-101, R-102), a liquid phase reaction occurs. Triglycerides react with methanol in the presence of a homogeneous potassium hydroxide catalyst at an average temperature of 140ºF and pressure of 59 PSI to form fatty methyl esters and glycerol. This temperature is safely below the methanol boiling point of 218ºF at 59 PSI.

Based on the various triglyceride chemical formulae reported in table XXVII as well as the three reversible equations that each triglyceride undergoes as presented in table XXVIII, the transesterification reaction network may be defined for the lipids extracted from *Chlorella protothecoides*. Since each triglyceride participates in 3 reversible reactions and there are three triglycerides species present in the algal oil, there are a total of 9 forward reactions and 9 reverse reactions, which are shown below in table XXIX. The rate constants that correspond to the reactions are also listed in table XXIX. While rate constants are readily available for a wide variety of vegetable oils, there is no available kinetic data for oil extracted from *Chlorella protothecoides*. For the purposes of the Aspen PLUS simulation, it was assumed that the rate constants measured by Leevijit et al. for palm oil at 60ºC and 1% catalyst concentration (by weight) approximate the rate constants for *Chlorella protothecoides* oil at the same conditions.[88]

Using the rate constants listed above in table XXVI, six differential equations may be written to represent the transesterification reaction network. Each differential equation corresponds to the concentration of one of the six species formed or consumed: triglycerides, diglycerides, monoglycerides, FAME, methanol, and glycerol. These equations, shown in detail in Section D1 of the Appendix, were solved simultaneously on MATLAB to give plots of concentration versus reaction time. Since the initial species concentrations are known and the steady-state species concentrations may be read from the MATLAB plots at long times, the steady-state conversion for each reaction may be calculated.

Table XXVI: The transesterification reaction network and associated rate constants

Rxn #	Stoichiometry	Rate Constant Name	Rate Constant (L/mol•min)
1	$C_{57}H_{104}O_6 + CH_3OH \rightarrow C_{19}H_{36}O_2 + C_{39}H_{72}O_5$	k1f	0.634
1	$C_{39}H_{72}O_5 + C_{19}H_{36}O_2 \rightarrow C_{57}H_{104}O_6 + CH_3OH$	k1r	0.000
2	$C_{39}H_{72}O_5 + CH_3OH \rightarrow C_{19}H_{36}O_2 + C_{21}H_{40}O_4$	k2f	7.104
2	$C_{21}H_{40}O_4 + C_{19}H_{36}O_2 \rightarrow C_{39}H_{72}O_5 + CH_3OH$	k2r	4.192
3	$C_{21}H_{40}O_4 + CH_3OH \rightarrow C_{19}H_{36}O_2 + C_3H_8O_3$	k3f	7.860
3	$C_3H_8O_3 + C_{19}H_{36}O_2 \rightarrow C_{21}H_{40}O_4 + CH_3OH$	k3r	0.121
4	$C_{57}H_{98}O_6 + CH_3OH \rightarrow C_{19}H_{34}O_2 + C_{39}H_{68}O_5$	k1f	0.634
4	$C_{39}H_{68}O_5 + C_{19}H_{34}O_2 \rightarrow C_{57}H_{98}O_6 + CH_3OH$	k1r	0.000
5	$C_{39}H_{68}O_5 + CH_3OH \rightarrow C_{19}H_{34}O_2 + C_{21}H_{38}O_4$	k2f	7.104
5	$C_{21}H_{38}O_4 + C_{19}H_{34}O_2 \rightarrow C_{39}H_{68}O_5 + CH_3OH$	k2r	4.192
6	$C_{21}H_{38}O_4 + CH_3OH \rightarrow C_{19}H_{34}O_2 + C_3H_8O_3$	k3f	7.860
6	$C_{19}H_{36}O_2 + C_{19}H_{34}O_2 \rightarrow C_{21}H_{38}O_4 + CH_3OH$	k3r	0.121
7	$C_{51}H_{98}O_6 + CH_3OH \rightarrow C_{17}H_{34}O_2 + C_{35}H_{68}O_5$	k1f	0.634
7	$C_{35}H_{68}O_5 + C_{17}H_{34}O_2 \rightarrow C_{51}H_{98}O_6 + CH_3OH$	k1r	0.000
8	$C_{35}H_{68}O_5 + CH_3OH \rightarrow C_{17}H_{34}O_2 + C_{19}H_{38}O_4$	k2f	7.104
8	$C_{19}H_{38}O_4 + C_{17}H_{34}O_2 \rightarrow C_{35}H_{68}O_5 + CH_3OH$	k2r	4.192
9	$C_{19}H_{38}O_4 + CH_3OH \rightarrow C_{17}H_{34}O_2 + C_3H_8O_3$	k3f	7.860
9	$C_3H_8O_3 + C_{17}H_{34}O_2 \rightarrow C_{21}H_{38}O_4 + CH_3OH$	k3r	0.121

Significantly, the forward reactions were observed to dominate, resulting in a high overall conversion of roughly 97.5%. The results are summarized in table XXVII below.

Table XXVII. MATLAB Results for the Transesterification Reaction Model

Reaction	Conversion at Steady-State (140ºF, 1% Cat w/w)
(1) TG + MeOH $\leftrightarrow$ FAME + DG	0.99997
(2) DG + MeOH $\leftrightarrow$ FAME + MG	0.99008
(3) MG + MeOH $\leftrightarrow$ FAME + GL	0.98517
(Overall) TG + 3 MeOH $\rightarrow$ 3 FAME + GL	0.97537

Although an overall conversion on the order of 97.5% is very high for the transesterification reacton, in order to meet industrial standards for biodiesel consumption, the final concentration of diglycerides and monoglycerides in solution must be less than 0.2% and 0.8% by weight respectively.[89] In order to meet this rigorous specification, two reactors in series are required. Effluent from the first reactor is cooled and sent to a decanter, where the heavy glycerol product readily separates by gravity from the light phase, comprised of unreacted triglycerides, diglycerides and monoglycerides, methanol and FAME.[84] Then, the light phase is sent to a second reactor, where the reaction essentially goes to completion, reaching an overall conversion on the order of 99.97%. Enough methanol and catalyst remains in the light phase stream leaving the decanter for an efficient reaction in the second reactor.

Effective inter-stage removal of glycerol is essential for the overall conversion to be high for the series of reactions.[90] By removing glycerol, an end product of transesterification, from the effluent from the first reactor, reverse reactions involving the unreacted diglycerides and monoglycerides are prevented. This ensures that forward reactions will prevail in the second reactor. For additional details regarding the reactor design and geometry, please see Section D, “Equipment List and Unit Descriptions,” on page 167.

Stream Separation

The hot effluent from the first reactor passes through Cooler C-103, where it is decreased in temperature to 100ºF. The cooled stream enters the decanter (S-101), where the insoluble glycerol product readily separates from the unreacted triglycerides, diglycerides, monoglycerides and FAME product. Excess methanol is present in both the heavy and light phases, although it predominantly leaves the decanter along with the crude glycerol. A small amount of water and potassium hydroxide is also present in the crude glycerol, although the majority of the catalyst remains in the light phase.

The crude glycerol is sent to a vacuum distillation column (D-101), where methanol is removed and recycled. The bottoms product is then cooled (C-105) before being pumped into the crude glycerol storage tank (T-104). The light phase leaving the decanter is heated (H-101) and enters the second reactor (R-102). The effluent from the second reactor is sent to be refined.

Product Refining

- ***Alcohol Removal***

The crude biodiesel stream exiting the second reactor (R-102) contains several impurities, including methanol, glycerol, water, potassium ions, hydroxide ions, and very small unreacted quantities of triglycerodes diglcyerides, and monoglycerides. Since methanol boils at just 150ºF at atmospheric pressure, it may be effectively separated from the crude FAME stream, which boils between 500-650ºF –depending on composition– in a distillation column (D-102).[84] To reduce the temperature throughout the column, distillation is operated under a vacuum at 8 PSI. The methanol recovered from the crude biodiesel product is combined with the methanol removed from the crude glycerol stream and recycled.

- ***Neutralization and Washing***

Fatty acid methyl esters are very susceptible to being hydrolyzed to free fatty acids in the presence of the basic potassium hydroxide catalyst.[91] Therefore, the crude biodiesel stream must be neutralized to guarantee that the product has a low free fatty acid content. Hydrochloric acid is mixed with the basic biodiesel solution, resulting in the formation of a neutralized mixture that contains water as well as dissolved potassium and chloride ions. This stream is sent to a liquid-liquid extractor (L-101), where the crude biodiesel is washed with a significant quantity of process water at elevated temperature, which readily dissolves potassium and chloride ions in addition to methanol, glycerol and a significant portion of the remaining diglcyerides and monoglycerides. The resulting biodiesel product stream is free of nearly all impurities, but retains a large amount of water from the washing step. It is cooled below room temperature and sent to a decanter (S-102), where the heavy water phase readily separates from the light FAME product. The light phase is heated to ambient temperature and sent to shift tanks (T-109, T-110) for complete chemical analysis before being deposited in a product storage tank (T-105).

C. Energy Balance and Utility Requirements

The energy requirements for the transesterification module may be determined largely from the Aspen PLUS simulation results. Tables XXVIII through XXXV list the annual utility requirements and the associated expenditures.

TABLE XXVIII. Annual Electricity Requirements for Transesterification

ELECTRICITY				
Unit	**Equipment**	**Annual Consumption (kW-hr)**	**Price**	**Annual Cost ($)**
P-101	Centrifugal Pump 1	1,600	$0.060/kW-hr	90
P-102	Centrifugal Pump 2	42,800	$0.060/kW-hr	2,570
P-103	Centrifugal Pump 3	265,500	$0.060/kW-hr	15,930
P-104	Centrifugal Pump 4	149,600	$0.060/kW-hr	8,970
P-105	Centrifugal Pump 5	1,307,600	$0.060/kW-hr	78,450
P-106	Centrifugal Pump 6	1,489,200	$0.060/kW-hr	89,350
P-107	Centrifugal Pump 7	126,300	$0.060/kW-hr	7,570
P-108	Centrifugal Pump 8	800	$0.060/kW-hr	50
P-109	Centrifugal Pump 9	9,700	$0.060/kW-hr	580
P-110	Centrifugal Pump 10	17,700	$0.060/kW-hr	1,060
P-111	Centrifugal Pump 11	9,100	$0.060/kW-hr	540
P-112	Centrifugal Pump 12	187,800	$0.060/kW-hr	11,260
P-113	Centrifugal Pump 13	400	$0.060/kW-hr	20
P-114	Centrifugal Pump 14	900	$0.060/kW-hr	50
P-115	Centrifugal Pump 15	2,188,400	$0.060/kW-hr	131,300
P-116	Centrifugal Pump 16	316,100	$0.060/kW-hr	18,960
P-117	Centrifugal Pump 17	243,200	$0.060/kW-hr	14,590
P-118	Centrifugal Pump 18	105,300	$0.060/kW-hr	6,310
P-119	Centrifugal Pump 19	1,109,200	$0.060/kW-hr	66,550
P-120	Centrifugal Pump 20	790,300	$0.060/kW-hr	47,410
P-121	Centrifugal Pump 21	10,000	$0.060/kW-hr	600
P-122	Centrifugal Pump 22	10,000	$0.060/kW-hr	600
B-101	Blower 1	15,700	$0.060/kW-hr	940
B-102	Blower 2	11,500	$0.060/kW-hr	690
R-102	Reactor 2	69,600	$0.060/kW-hr	4,170

TOTAL		**8,478,300**	**$**	**509,000**

TABLE XXIX. Annual High Pressure Steam Requirements for Transesterification

HIGH PRESSURE STEAM (450 PSIG)				
Unit	**Equipment**	**Annual Consumption (lb)**	**Price**	**Annual Cost ($)**
H-109	Dist Tower 2 Reboiler	361,336,000	$6.60/1000 lb	2,385,000

TOTAL		**361,336,000**	**$**	**2,385,000**

TABLE XXX Annual Medium Pressure Steam Requirements for Transesterification

MEDIUM PRESSURE STEAM (150 PSIG)				
Unit	**Equipment**	**Annual Consumption (lb)**	**Price**	**Annual Cost ($)**
H-107	Dist Tower 1 Reboiler	131,737,000	$4.80/1000 lb	633,000
TOTAL		**131,737,000**	**$**	**633,000**

TABLE XXXI. Annual Cooling Water Requirements for Transesterification

COOLING WATER				
Unit	**Equipment**	**Annual Consumption (gal)**	**Price**	**Annual Cost ($)**
C-103	Cooler 3	320,121,000	$0.075/1000 gal	25,000
C-104	Cooler 4	358,888,000	$0.075/1000 gal	27,000
C-105	Cooler 5	122,821,000	$0.075/1000 gal	10,000
C-106	Cooler 6	198,985,000	$0.075/1000 gal	15,000
H-106	Dist Tower 1 Condenser	310,672,000	$0.075/1000 gal	24,000
H-108	Dist Tower 2 Condenser	159,213,000	$0.075/1000 gal	12,000
TOTAL	**Cooling Water**	**1,470,696,000**		**113,000**

TABLE XXXII. Annual Process Water Requirements for Transesterification

PROCESS WATER				
Unit	**Equipment**	**Annual Consumption (gal)**	**Price**	**Annual Cost ($)**
L-101	Liq-Liq Extractor	213,847,000	$0.75/1000 gal	161,000
TOTAL	**Process Water**	**213,847,000**		**161,000**

TABLE XXXIII. Annual Chilled Water Requirements for Transesterification

CHILLED WATER				
Unit	**Equipment**	**Annual Consumption (ton-day)**	**Price**	**Annual Cost ($)**
C-102	Cooler 2	220,500	$1.20/ton-day	265,000
TOTAL	**Chilled Water**	**220,500**		**265,000**

TABLE XXXIV. Annual Chilled Brine Requirements for Transesterification

CHILLED BRINE				
Unit	**Equipment**	**Annual Consumption (ton-day)**	**Price**	**Annual Cost ($)**
C-101	Cooler 1	67,600	$2.40/ton-day	163,000
C-102	Cooler 2	134,100	$2.40/ton-day	322,000
TOTAL	**Chilled Brine**	**201,700**		**485,000**

TABLE XXXV. Annual Wastewater Treatment Requirements for Transesterification

WASTEWATER				
Unit	**Equipment**	**Annual Consumption (lb)**	**Price**	**Annual Cost ($)**
W-101	Wastewater Treatment	18,294,000	$0.15/lb	2,745,000
TOTAL	**Wastewater Treatment**	**18,294,000**		**2,745,000**

TOTAL ANNUAL UTILITIES COST	**$**	**5,913,000**

The major utilities required for transesterification include electricity, high-pressure steam, medium-pressure steam, cooling water, process water, chilled water, chilled brine, and wastewater treatment. Since the transesterification plant is located in the vicinity of a large power plant as well as industrial scale desalination facilities, these utilities should be readily available. There is no need to build a distinct utility plant or incur significant allocated costs.

Electricity is used to supply energy mainly to the centrifugal pumps present throughout the transesterification facility. These pumps are required to maintain the steady flow of material from one process unit to the next. Pumps P-106, P-107, P-112, P-119 and P-120 pump material that is changing in height as it enters a vessel, such as a distillation column or liquid-liquid extraction unit. Therefore, they consume large amounts of energy. Since pumps P-105 P-115, and P-116 handle very large volumetric flow rates, they consume significant amounts of electricity as well. In addition to pumps, two centrifugal blowers require electricity to generate vacuum pressure in the two distillation columns, a condition essential for operation within a reasonable temperature range. Electricity requirements were considered at $0.06/kWh.[67]

High-pressure steam is required in the kettle vaporizer H-109, which is part of the second vacuum distillation column apparatus, D-102. Saturated team at 450 PSIG has a temperature of 460ºF and latent heat of 766 BTU/lb.[92] Its condensation is required to generate the large source of heat energy needed to return a portion of the bottoms product as vapor to the bottom tray of the distillation tower. The price of high-pressure steam, $6.60/lb, was assessed based on Table 23.1 of *Product and Process Design Principles.*[67]

Medium pressure steam is required in the kettle reboiler H-107, which is part of the first vacuum distillation column apparatus, D-101. Saturated team at 150 PSIG has a temperature of 366ºF and latent heat of 857 BTU/lb.[82] Its condensation is required to generate the large source of heat energy needed to return a portion of the bottoms product as vapor to the bottom tray of the distillation tower. The price of medium pressure steam, $4.80/lb, was assessed based on Table 23.1 of *Product and Process Design Principles*.[67]

Cooling water is essential to transesterification and is used to condense vapors at the top of the vacuum distillation towers as well as to cool hot liquid products leaving the vacuum distillation and liquid-liquid extraction columns. Cooling water is supplied at 90ºF and is heated to a maximum of 120ºF. Limiting the temperature range of the cooling water prevents dissolved ions and salts from fouling pipelines or heat exchanger tubing. The heated water is sent to a cooling tower and may be recycled or returned to a nearby body of water. The price of cooling water, $0.075/1000 gal, was assessed based on Table 23.1 of *Product and Process Design Principles*.[67]

Process water is only required for the water-washing phase of biodiesel refining that occurs in the liquid-liquid extraction vessel, L-101. A mass flow rate of process water approximately 140% of the crude biodiesel flow rate is required for adequate washing and impurity removal, resulting in significant water consumption at a price, based on Table 23.1 of *Product and Process Design Principles*, of $0.75/1000 gal.[67] Once the process water is contaminated with dissolved ions, glycerol, methanol, diglycerides and monoglycerides, it becomes a wastewater stream that contains 1% impurities and must be treated. Based on a price of $0.15 per lb of organic material that must be removed, wastewater treatment requires an annual expenditure of nearly 2.75 million dollars, a very large portion of the overall utility costs.

Since the decanter used to remove excess water from the purified biodiesel product operates at 55ºF, chilled water is required to cool the inlet stream. Entering a heat exchanger (C-107) at 40ºF, the chilled water is heated to 90ºF, a temperature at which it may be combined with the cooling water stream for regular use in cooling applications. Chilled water, which costs $1.20/ton-day to refrigerate, is significantly more expensive than cooling or process water, but is necessary to bring the crude biodiesel stream to temperatures low enough to encourage the formation of two distinct liquid phases in the decanter.[67]

The transesterification reactor R-101 was designed to operate at an average temperature of 140ºF, a condition maintained by feeding cold shots to various inlets along the tubular vessel. Chilled brine at -30ºF is required to cool the methanol and triglyceride cold shot streams to 10ºF. The brine is heated to 0ºF, refrigerated to -30ºF, and recycled. Chilled brine, which costs $2.40/ton-day to refrigerate, is significantly more expensive than cooling or process water.[67] However, its use is necessary to keep the reactor temperature within an acceptable range to control to the exothermic reaction and ensure that the methanol reactant remains a liquid at the operating pressure.

D. Equipment List and Unit Descriptions

TABLE XXXVI. A List of Equipment used in the Transesterification Process

Unit #	Equipment Type	Function	Size	Mat'l	Oper T (°F)	Oper P (PSI)
D-101	Distillation Tower	Removes methanol from the crude glycerol product stream	D = 3 ft H = 48 ft ts = 0.5 in	Carbon Steel	335	8
D-102	Distillation Tower	Removes methanol from the FAME product stream	D = 2.75 ft H = 64 ft ts = 0.625 in	Carbon Steel	428	8
P-101	Centrifugal Pump	Increases the pressure of the KOH feed stream	Pc = 0.23 hP V = 2.25 gal/min H = 105 ft	Cast Iron	77	105
P-102	Centrifugal Pump	Increases the pressure of the methanol feed stream	Pc = 6.6 hP V = 65 gal/min H = 265 ft	Cast Iron	77	105

P-103	Centrifugal Pump	Increases the pressure of the triglyceride feed stream	Pc = 41 hP V = 600 gal/min H = 250 ft	Cast Iron	77	95
P-104	Centrifugal Pump	Increases the pressure of the chilled brine stream	Pc = 23 hP V = 780 gal/min H =65 ft	Cast Iron	-30	50
P-105	Centrifugal Pump	Increases the pressure of the cooling water stream	Pc = 201 hP V = 6170 gal/min H =90 ft	Cast Iron	90	52
P-106	Centrifugal Pump	Increases the pressure of the feed stream to the second reactor	Pc = 228 hP V = 575 gal/min H = 1345 ft	Cast Iron	100	520
P-107	Centrifugal Pump	Increases the pressure of the feed stream to Distillation Tower D-101	Pc = 19 hP V = 68 gal/min H = 665 ft	Cast Iron	100	382
P-108	Centrifugal Pump	Reflux Pump for Distillation Tower D-101	Pc = 0.12 hP V = 5 gal/min H = 87.5 ft	Cast Iron	123	36
P-109	Centrifugal Pump	Increases the pressure of the bottoms stream from Distillation Tower D-101	Pc = 1.5 hP V = 44 gal/min H = 85 ft	Cast Iron	335	50
P-110	Centrifugal Pump	Increases the pressure of the distillate stream from Distillation Tower D-101	Pc = 3 hP V = 45 gal/min H = 220 ft	Cast Iron	123	80
P-111	Centrifugal Pump	Increases the pressure of the distillate stream from Distillation Tower D-102	Pc = 1.6 hP V = 23 gal/min H = 220 ft	Cast Iron	122	80
P-112	Centrifugal Pump	Increases the pressure of the bottoms stream from Distillation Tower D-102	Pc = 29 hP V = 605 gal/min H = 175 ft	Cast Iron	428	67
P-113	Centrifugal Pump	Reflux Pump for Distillation Tower D-102	Pc = 0.75 hP V = 2.25 gal/min H = 87.5 ft	Cast Iron	122	36
P-114	Centrifugal Pump	Increases the pressure of the HCl feed stream	Pc = .17 hP V = 3.5 gal/min H = 98 ft	Cast Iron	77	63
P-115	Centrifugal Pump	Increases the pressure of the process water stream	Pc = 335 hP V = 415 gal/min H = 2235 ft	Cast Iron	77	985
P-116	Centrifugal Pump	Increses the pressure of the chilled water feed stream	Pc = 48 hP V = 1410 gal/min H = 95 ft	Cast Iron	40	55

P-117	Centrifugal Pump	Increases the pressure of the lean phase leaving the Liquid-Liquid Extractor	Pc = 37 hP V = 745 gal/min H = 155 ft	Cast Iron	166	74
P-118	Centrifugal Pump	Increases the pressure of the waste-water stream moving to treatment	Pc = 16.1 hP V = 430 gal/min H =105 ft	Cast Iron	165	58
P-119	Centrifugal Pump	Increases pressure of the process water recycle stream	Pc = 170 hP V = 210 gal/min H = 2245 ft	Cast Iron	54	985
P-120	Centrifugal Pump	Increases the pressure of the Biodiesel product leaving the transesterification process	Pc = 121 hP V = 515 gal/min H =735 ft	Cast Iron	77	295
P-121	Centrifugal Pump	Increases the pressure of Biodiesel product leaving shift tank T-109	Pc = 1.5 hP V = 65 gal/min H =74 ft	Cast Iron	77	43
P-122	Centrifugal Pump	Increases the pressure of Biodiesel product leaving shift tank T-110	Pc = 1.5 hP V = 65 gal/min H =74 ft	Cast Iron	77	43
B-101	Blower	Maintains vacuum pressure for distillation tower D-101	Pc = 2.4 hP	Cast Iron	123	8
B-102	Blower	Maintains vacuum pressure for distillation tower D-102	Pc = 1.75 hP	Cast Iron	122	8
H-101	Shell and Tube Heat Exchanger	Increases the temperature of the feed stream to R-102	Q = 3,297,000 BTU/hr A = 3635 ft^2	Carbon Steel, Brass	140	47
H-102	Double-Pipe Heat Exchanger	Increases temperature of HCl feed stream	Q = 386,000 BTU/hr A = 50 ft^2	Carbon Steel, Stainless Steel	156	42
H-103	Shell and Tube Heat Exchanger	Increases the temperature of the process water stream entering the liq-liq extractor	Q = 25,989,000 BTU/hr A = 6960 ft^2	Carbon Steel, Brass	165	35
H-104	Shell and Tube Heat Exchanger	Increases the temperature of the Biodiesel product stream	Q = 1,522,000 BTU/hr A = 1375 ft^2	Carbon Steel, Brass	77	39

H-105	Shell and Tube Heat Exchanger	Increases the temperature of the process water recycle stream	Q = 3,715,000 BTU/hr A = 1785 ft^2	Carbon Steel, Brass	89	47
H-106	Condenser	Reflux condenser for distillation tower D-101	Q =8,832,000 BTU/hr A = 3445 ft^2	Carbon Steel, Brass	123	23
H-107	Kettle Vaporizer	Kettle Vaporizer for distillation tower D-101	Q = 12,888,000 BTU/hr A = 225 ft^2	Carbon Steel, Brass	335	25
H-108	Condenser	Reflux condenser for distillation tower D-102	Q =4,526,000 BTU/hr A = 1975 ft^2	Carbon Steel, Brass	122	23
H-109	Kettle Vaporizer	Kettle Vaporizer for distillation tower D-102	Q = 31,596,000 BTU/hr A = 2,760 ft^2	Carbon Steel, Brass	428	25
C-101	Shell and Tube Heat Exchanger	Decreases the temperature of the methanol cold shots	Q = 2,222,000 BTU/hr A = 660 ft^2	Carbon Steel, Brass	10	65
C-102	Shell and Tube Heat Exchanger	Decreases the temperature of the triglycerides cold shot streams	Q = 4,409,000 BTU/hr A = 2,230 ft^2	Carbon Steel, Brass	10	65
C-103	Shell and Tube Heat Exchanger	Decreases the temperature of the inlet stream to decanter S-101	Q = 9,143,000 BTU/hr A = 13,110 ft^2	Carbon Steel, Brass	100	53
C-104	Shell and Tube Heat Exchanger	Decreases the temperature of the wastewater stream	Q = 10,250,000 BTU/hr A =2840 ft^2	Carbon Steel, Brass	100	23
C-105	Shell and Tube Heat Exchanger	Decreases the temperature of the crude glcyerol product stream	Q = 3,508,000 BTU/hr A = 350 ft^2	Carbon Steel, Brass	100	24
C-106	Shell and Tube Heat Exchanger	Decreases the temperature of the crude biodiesel stream entering the decanter S-102	Q = 5,683,000 BTU/hr A = 4,940 ft^2	Carbon Steel, Brass	100	32
C-107	Shell and Tube Heat Exchanger	Further decreases the temperature of the crude biodiesel stream entering the decanter S-102	Q = 7,249,000 BTU/hr A = 5,990 ft^2	Carbon Steel, Brass	54.5	25
T-101	Storage Tank	KOH Feed Storage tank	V = 3010 ft^3	Steel Alloy	77	15

T-102	Storage Tank	Methanol feed storage tank	V = 59,900 ft^3	Steel Alloy	77	15
T-103	Storage Tank	Triglyceride feed storage tank	V = 133,680 ft^3	Steel Alloy	77	15
T-104	Storage Tank	Glycerol product storage tank	V = 37,350 ft^3	Steel Alloy	77	15
T-105	Storage Tank	Biodiesel product storage tank	V = 133,680 ft^3	Steel Alloy	77	15
T-106	Storage Tank	Hydrochloric Acid feed storage tank	V = 2,975 ft^3	Steel Alloy	77	15
T-107	Surge Drum	Stores Liquid-Liquid Extractor Light Phase Surge	V = 500 ft^3	Carbon Steel	155	20
T-108	Surge Drum	Stores Liquid-Liquid Extractor Heavy Phase Surge	V = 285 ft^3	Carbon Steel	155	20
T-109	Shift Tank	Stores 8 hours of biodiesel product for analysis	V = 32,840 ft^3	Steel Alloy	77	15
T-110	Shift Tank	Stores 8 hours of biodiesel product for analysis	V = 32,840 ft^3	Steel Alloy	77	15
S-101	Decanter	Separates glycerol from reactor R-101 effluent	D = 9.75 ft L = 40 ft	Carbon Steel	100	46
S-102	Decanter	Separates water from liquid-liquid extractor light phase	D = 9.75 ft L = 40 ft	Carbon Steel	54.5	15
L-101	Liquid-Liquid Extractor	Removes potassium and hydroxide ions from crude biodiesel stream	D = 13.5 ft H = 66 ft	Carbon Steel	165	15
S-103	Reflux Accumulator	Accumulates reflux for distillation tower D-101	D = 2.75 ft L = 5.5 ft	Carbon Steel	123	8
S-104	Reflux Accumulator	Accumulates reflux for distillation tower D-102	D = 2.25 ft L = 4.5 ft	Carbon Steel	122	8
R-101	Tubular Reactor	Transesterification Reactor that converts triglycerides into FAME	D = 8 in L = 580 ft 4 parallel tubes	Carbon Steel	140	53
R-102	Stirred Tank Reactor	Transesterification Reactor that converts triglycerides into FAME	V = 470 ft^3 Stir Speed = 300 rpm	Carbon Steel	140	18

Distillation Tower 1 (D-101)

Distillation Tower 1 separates methanol and crude glycerol at moderate temperatures by employing the use of vacuum pressure. Pressure and temperature varies in the column from 10 PSI and 335 ºF at the bottom to 8 PSI and 123 ºF at the top. The distillation column is very effective and recovers 99.7% by mass of the methanol in the inlet stream. Sized based on operation at 85% of its flooding velocity, the tower was priced as a pressurized vessel with trays separated by 24 inches, ladders and platforms. The total bare-module cost of this unit is $289,900. Please refer to the Distillation Tower 1 Specification Sheet on page 189 and the Sample Calculations in Section E of the Appendix on page 366.

Distillation Tower 2 (D-102)

Similar to Distillation Tower 1, Distillation Tower 2 separates methanol and crude biodiesel at moderate temperatures by employing vacuum distillation technology. Pressure and temperature varies in the column from 10 PSI and 428 ºF at the bottom to 8 PSI and 122 ºF at the top. The distillation column is very effective and recovers 98.6% by mass of the methanol in the inlet stream. The tower was sized based on operation at 85% of its flooding velocity and priced as a pressurized vessel with trays, separated by 24 inches. The cost of ladders and platforms was included as well. The total bare-module cost of this unit is $384,800. Please refer to the Distillation Tower 2 Specification Sheet on page 190 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 1 (P-101)

Pump P-101 is used to pump the potassium hydroxide feedstock from storage tank T-101 and through mixers and heat exchangers to the reactor so that it may serve as a homogeneous catalyst. Assuming a 65 PSI pressure drop between T-101 and R-101 and the need for pressure in the reactor to exceed 50 PSI, a pressure head of 105 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $15,600. Please refer to the Centrifugal Pump 1 Specification Sheet on page 191 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 2 (P-102)

Pump P-102 is used to pump the methanol feedstock from storage tank T-102 and through mixers and heat exchangers to the reactor so that it may act as a reactant during transesterification. Assuming a 65 PSI pressure drop between T-102 and R-101 and the need for pressure in the reactor to exceed 50 PSI, a pressure head of 265 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $13,600. Please refer to the Centrifugal Pump 2 Specification Sheet on page 192 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 3 (P-103)

Pump P-103 is used to pump the triglyceride feedstock from storage tank T-103 through mixers and heat exchangers to the reactor so that it may be transformed into the fatty methyl ester product via transesterification. Assuming a 65 PSI pressure drop between T-103 and R-101 and the need for pressure in the reactor to exceed 50 PSI, a pressure head of 250 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $33,500. Please refer to the Centrifugal Pump 3 Specification Sheet on page 193 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 4 (P-104)

Pump P-104 is used to pump chilled brine to the shell and tube heat exchangers C-101 and C-102 to cool portions of the triglyceride and methanol feed streams before they enter the first reactor (R-101). Assuming a 35 PSI pressure drop en route to the cooling unit, a pressure head of 65 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $24,800. Please refer to the Centrifugal Pump 4 Specification Sheet on page 194 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 5 (P-105)

Pump P-105 is used to pump cooling water at 90ºF through a central conduit to deliver water to four shell and tube heat exchangers (C-103, C-104, C-105 and C-106) as well as two condensers (H-106 and H-108). The cooling water is essential for lowering the temperature of heated streams or condensing vapors into liquids. Assuming a 37 PSI pressure drop through the control valves, piping and heat exchanger units, a pressure head of 90 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $118,500. Please refer to the Centrifugal Pump 5 Specification Sheet on page 195 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 6 (P-106)

Pump P-106 is used to pump the feed stream consisting of FAME, methanol, potassium hydroxide catalyst and unreacted diglycerides and monoglycerides from the decanter (S-101) through a pre-heater to the second reactor unit (R-102) and onto Distillation Tower 2. Assuming a 475 PSI pressure drop between the decanter and the vacuum distillation column, a pressure head of 1375 ft. is calculated. This significant pressure drop is a consequence of the Distillation Tower 2 height, since the inlet stream must be pumped 32 vertical feet halfway up the column. This unit is a centrifugal pump made of cast iron. The total bare module cost is $303,900. Please refer to the Centrifugal Pump 6 Specification Sheet on page 196 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 7 (P-107)

Pump P-107 is used to pump the crude glycerol stream from the decanter (S-101) through a pre-heater to the second reactor unit (R-102) and onto Distillation Tower 1. Assuming a 340 PSI pressure drop between the decanter and the vacuum distillation column, a pressure head of 665 ft. is calculated. This significant pressure drop is a consequence of the Distillation Tower 1 height, since the inlet stream must be pumped 24 vertical feet halfway up the column. This unit is a centrifugal pump made of cast iron. The total bare module cost is $38,300. Please refer to the Centrifugal Pump 7 Specification Sheet on page 197 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 8 (P-108)

Pump P-108 is the reflux pump associated with Distillation Tower 1. Assuming a 28 PSI pressure drop, a pressure head of 87.5 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $13,800. Please refer to the Centrifugal Pump 8

Specification Sheet on page 198 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 9 (P-109)

Pump P-109 is the bottoms product pump associated with Distillation Tower 1 and pumps hot, crude glycerol through a control valve and two heat exchanges. Assuming a 40 PSI pressure drop, a pressure head of 85 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $11,900. Please refer to the Centrifugal Pump 9 Specification Sheet on page 199 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 10 (P-110)

Pump P-110 is the distillate product pump associated with Distillation Tower 1 and pumps a liquid methanol recycle stream. Assuming a 72 PSI pressure drop, a pressure head of 220 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $12,600. Please refer to the Centrifugal Pump 10 Specification Sheet on page 200 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 11 (P-111)

Pump P-111 is the distillate product pump associated with Distillation Tower 2 and pumps a liquid methanol recycle stream. Assuming a 72 PSI pressure drop, a pressure head of 220 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $11,800. Please refer to the Centrifugal Pump 11 Specification Sheet on page 201 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 12 (P-112)

Pump P-112 is the bottoms product pump associated with Distillation Tower 2 and pumps crude FAME to the separation train for refining. Assuming a 57 PSI pressure drop, a pressure head of 175 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $28,200. Please refer to the Centrifugal Pump 12 Specification Sheet on page 202 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 13 (P-113)

Pump P-113 is the reflux pump associated with Distillation Tower 1. Assuming a 28 PSI pressure drop, a pressure head of 87.5 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $16,200. Please refer to the Centrifugal Pump 13 Specification Sheet on page 203 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 14 (P-114)

Pump P-114 is used to pump the hydrochloric acid feedstock from storage tank T-106 to a mixer, where it neutralizes the crude biodiesel stream. Assuming a 48 PSI pressure drop, a pressure head of 98 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $14,100. Please refer to the Centrifugal Pump 14 Specification Sheet on page 204 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 15 (P-115)

Pump P-115 is used to pump process water to the liquid-liquid extractor unit (L-101) to wash the crude biodiesel product. Since the process water enters the 66 ft tall extractor column at the top, a significant pressure of 969 PSI must be delivered. This requires a pressure head of 2,235 ft. This unit is a centrifugal pump made of cast iron. The total bare module cost is $318,400. Please refer to the Centrifugal Pump 5 Specification Sheet on page 205 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 16 (P-116)

Pump P-116 is used to pump chilled water to the shell and tube heat exchanger C-107 to cool the wet biodiesel product before it enters the decanter (S-102). Assuming a 40 PSI pressure drop en route to the cooling unit, a pressure head of 95 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $39,100. Please refer to the Centrifugal Pump 16 Specification Sheet on page 206 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 17 (P-117)

Pump P-117 is used to pump the lean light phase from the liquid-liquid extraction vessel (L-101) through a series of heat exchangers to the decanter (S-102). Assuming a 59 PSI pressure drop, a pressure head of 155 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $32,100. Please refer to the Centrifugal Pump 17 Specification Sheet on page 207 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 18 (P-118)

Pump P-118 is used to pump the wastewater from the liquid-liquid extractor (L-101) to be treated. Assuming a 43 PSI pressure drop, a pressure head of 105 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $20,600. Please refer to the Centrifugal Pump 18 Specification Sheet on page 208 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 19 (P-119)

Pump P-119 is used to pump the process water recycle stream from the decanter (S-102) to be mixed with fresh process water. Assuming a 970 PSI pressure drop, a pressure head of 2245 ft. is calculated. This substantial head is required to return the process water to the top of the 66 ft. liquid-liquid extractor. This unit is a centrifugal pump made of cast iron. The total bare module cost is $226,100. Please refer to the Centrifugal Pump 19 Specification Sheet on page 209 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 20 (P-120)

Pump P-120 is used to pump the biodiesel product stream from the decanter (S-102) to shift-tanks (T-109, T-110) to be analyzed for quality assurance. Assuming a 281 PSI pressure drop, a pressure head of 735 ft. is calculated. This substantial head is required to pump the biodiesel to the top of the 20 ft shift storage tanks. This unit is a centrifugal pump made of cast iron. The total bare module cost is $103,800. Please refer to the Centrifugal Pump 20 Specification Sheet on page 210 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 21 (P-121)

Pump P-121 is used to empty the shift tank pump T-109 and deliver refined biodiesel fuel to the product storage tank T-105. Assuming a 28 PSI pressure drop, a pressure head of 74 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $12,000. Please refer to the Centrifugal Pump 21 Specification Sheet on page 211 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Pump 22 (P-122)

Pump P-122 is used to empty the shift tank pump T-110 and deliver refined biodiesel fuel to the product storage tank T-105. Assuming a 28 PSI pressure drop, a pressure head of 74 ft. is calculated. This unit is a centrifugal pump made of cast iron. The total bare module cost is $12,000. Please refer to the Centrifugal Pump 22 Specification Sheet on page 212 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Blower 1 (B-101)

The centrifugal blower B-101 maintains vacuum pressure in distillation tower 1. Based on the tower's volume, air is expected to leak into the column and rise with the vapor stream at a rate of 7.4 lb/hr. The centrifugal blower removes this air along with some methanol from the vapor space in the reflux accumulator and operates at a compression ratio of 2. This unit is a centrifugal blower made of cast iron. The total bare module cost is $4,300. Please refer to the Centrifugal Blower 1 Specification Sheet on page 213 and the Sample Calculations in Section E of the Appendix on page 366.

Centrifugal Blower 2 (B-102)

The centrifugal blower B-102 maintains vacuum pressure in distillation tower 2. Based on the tower's volume, air is expected to leak into the column and rise with the vapor stream at a rate of 8.5 lb/hr. The centrifugal blower removes this air along with some methanol from the vapor space in the reflux accumulator and operates at a compression ratio of 2. This unit is a centrifugal blower made of cast iron. The total bare module cost is $3,400. Please refer to the Centrifugal Blower 2 Specification Sheet on page 214 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 1 (H-101)

Heat exchanger 1 is a fixed head shell and tube heat exchanger that removes heat from hot wastewater to preheat the feed stream, consisting of triglycerides, diglycerides, monoglycerides, methanol, FAME and potassium hydroxide, to the second reactor (R-102). The feed stream is heated from 100ºF to 140ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $176,200. Please refer to the Heat Exchanger 1 Specification Sheet on page 215 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 2 (H-102)

Heat exchanger 2 is a double-pipe heat exchanger that removes heat from hot crude biodiesel to preheat the HCl feed stream prior to neutralization and liquid-liquid extraction. The HCl stream is heated from 77ªF to 157ºF. Based on a carbon outer pipe and stainless steel inner pipe construction, the total bare-module cost of this unit is $12,900. Please refer to the Heat Exchanger 2 Specification Sheet on page 216 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 3 (H-103)

Heat exchanger 3 is a fixed head shell and tube heat exchanger that removes heat from hot crude biodiesel to preheat the process water feed stream. The process water stream is heated from 81ªF to 165ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $292,600. Please refer to the Heat Exchanger 3 Specification Sheet on page 217 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 4 (H-104)

Heat exchanger 4 is a fixed head shell and tube heat exchanger that heats the refined biodiesel product leaving the chilled decanter (S-102) with the warm lean light phase stream leaving the liquid-liquid extractor. The biodiesel stream is heated from 54ªF to 77ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $95,900. Please refer to the Heat Exchanger 4 Specification Sheet on page 218 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 5 (H-105)

Heat exchanger 5 is a fixed head shell and tube heat exchanger that heats the water phase leaving the chilled decanter (S-102) with the warm lean light phase stream leaving the liquid-liquid extractor. The water stream, which is recycled, is heated from 53ªF to 89ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $111,000. Please refer to the Heat Exchanger 5 Specification Sheet on page 219 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 6 (H-106)

Heat exchanger 6 is a fixed head shell and tube heat exchanger that uses cooling water to condense the vapor product of Distillation Tower 1 into a liquid stream at 123ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $168,400. Please refer to the Heat Exchanger 6 Specification Sheet on page 220 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 7 (H-107)

Heat exchanger 6 is a kettle vaporizer that uses medium-pressure stream to vaporize a portion of the liquid bottoms product of Distillation Tower 1 at 335ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $176,200. Please refer to the Heat Exchanger 7 Specification Sheet on page 221 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 8 (H-108)

Heat exchanger 8 is a fixed head shell and tube heat exchanger that uses cooling water to condense the vapor product of Distillation Tower 2 into a liquid stream at 122ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $116,900. Please refer to the Heat Exchanger 8 Specification Sheet on page 222 and the Sample Calculations in Section E of the Appendix on page 366.

Heat Exchanger 9 (H-109)

This is a kettle vaporizer that uses high pressure stream to vaporize a portion of the liquid bottoms product of Distillation Tower 1. For carbon steel shell and brass tube construction, the total bare-module cost of this unit is $317,400. Please refer to the Heat Exchanger 9 Specification Sheet on page 223 and the Sample Calculations in Section E of the Appendix on page 366.

Cooler 1 (C-101)

Cooler 1 is a fixed head shell and tube heat exchanger that uses chilled brine to decrease the temperature of the methanol feed to the first reactor (R-101). The stream is cooled from 77ªF to 10ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $68,900. Please refer to the Cooler 1 Specification Sheet on page 224 and the Sample Calculations in Section E of the Appendix on page 366.

Cooler 2 (C-101)

Cooler 2 is a fixed head shell and tube heat exchanger that uses chilled brine to decrease the temperature of the methanol feed to the first reactor (R-101). The stream is cooled from 77ªF to 10ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $127,400. Please refer to the Cooler 2 Specification Sheet on page 225 and the Sample Calculations in Section E of the Appendix on page 366.

Cooler 3 (C-103)

Cooler 3 is a fixed head shell and tube heat exchanger that uses cooling water to decrease the temperature of the effluent from the first reactor before it enters the decanter (S-101). The effluent stream is cooled from 180ªF to 100ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $523,000. Please refer to the Cooler 3 Specification Sheet on page 226 and the Sample Calculations in Section E of the Appendix on page 366.

Cooler 4 (C-104)

Cooler 4 is a fixed head shell and tube heat exchanger that uses cooling water to decrease the temperature of the hot wastewater stream before it is sent for treatment. The water stream is cooled from 150ªF to 100ºF Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $147,700. Please refer to the Cooler 4 Specification Sheet on page 227 and the Sample Calculations in Section E of the Appendix on page 366.

Cooler 5 (C-105)

Cooler 5 is a fixed head shell and tube heat exchanger that uses cooling water to decrease the temperature of the crude glycerol stream before it is sent to a product storage tank. The glycerol stream is cooled from 335ªF to 100ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $55,900. Please refer to the Cooler 5 Specification Sheet on page 228 and the Sample Calculations in Section E of the Appendix on page 366.

Cooler 6 (C-106)

Cooler 6 is a fixed head shell and tube heat exchanger that uses cooling water to decrease the temperature of the wet biodiesel stream that enters the decanter (S-102). The biodiesel stream is cooled from 135ªF to 100ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $221,200. Please refer to the Cooler 6 Specification Sheet on page 229 and the Sample Calculations in Section E of the Appendix on page 366.

Cooler 7 (C-107)

Cooler 7 is a fixed head shell and tube heat exchanger that uses chilled water to further decrease the temperature of the wet biodiesel stream that enters the decanter (S-102). The biodiesel stream is cooled from 100ªF to 55ºF. Based on a carbon steel shell and brass tube construction, the total bare-module cost of this unit is $257,700. Please refer to the Cooler 7 Specification Sheet on page 230 and the Sample Calculations in Section E of the Appendix on page 366.

Feed Storage Tank 1 (T-101)

Feed Storage Tank 1 is a floating roof tank that stores the potassium hydroxide catalyst before it is required for use in the transesterification reactor. The tank has a capacity to store 22,500 gallons of potassium hydroxide, which is adequate storage for 5 days of inventory. Using a steel alloy construction, the total bare-module cost of this unit is $289,000. Please refer to the Feed Storage Tank 1 Specification Sheet on page 231 and the Sample Calculations in Section E of the Appendix on page 366.

Feed Storage Tank 2 (T-102)

Feed Storage Tank 2 is a floating roof tank that stores methanol before it is required for use in the transesterification reactor. The tank has a capacity to store 450,000 gallons of methanol, which is adequate storage for 5 days of inventory. Using a steel alloy construction, the total bare-module cost of this unit is $1,329,000. Please refer to the Feed Storage Tank 2 Specification Sheet on page 232 and the Sample Calculations in Section E of the Appendix on page 366.

Feed Storage Tank 3 (T-103)

Feed Storage Tank 3 is a set of 2 floating roof tanks that store the triglyceride feedstock from the Origin Oil™ lipid extraction facility before it is required for use in the transesterification reactor. Each tank has a capacity to store 1,000,000 gallons of triglycerides, a total that is adequate storage for 2 days of inventory. The residence time is reduced because triglycerides are delivered from the nearby lipid extraction facility and not an outside location, minimizing potential transportation disruptions. Using a steel alloy construction, the total bare-module cost of these units is $3,993,000. Please refer to the Feed Storage Tank 3 Specification Sheet on page 233 and the Sample Calculations in Section E of the Appendix on page 366.

Product Storage Tank 4 (T-104)

Product Storage Tank 4 stores crude glycerol product before it is transported for use in the fermentation component of algal cultivation or to a glycerol refinery. The tank has a capacity to store 280,000 gallons of glycerol, which is adequate storage for 5 days of inventory. Using a steel alloy construction, the total bare-module cost of this unit is $1,043,000. Please refer to the Product Storage Tank 4 Specification Sheet on page 234 and the Sample Calculations in Section E of the Appendix on page 366.

Product Storage Tank 5 (T-105)

Product Storage Tank 5 is a set of 4 floating roof tanks that store the biodiesel product before it is transported to consumers. Each tank can store 1,000,000 gallons of biodiesel, which is adequate storage for 5 days of inventory. For a steel alloy construction, the total bare-module cost of these units is $7,985,000. Please refer to the Product Storage Tank 5 Specification Sheet on page 235 and the Sample Calculations in Section E of the Appendix on page 366.

Feed Storage Tank 6 (T-106)

Feed Storage Tank 6 is a floating roof tank that stores hydrochloric acid before it is required for use in neutralization. The tank has a capacity to store 2975 ft^3, or 22,200 gallons of HCl, which is adequate storage for 5 days of inventory. Using a steel alloy construction, the total bare-module cost of this unit is $305,000. Please refer to the Feed Storage Tank 6 Specification Sheet on page 236 and the Sample Calculations in Section E of the Appendix on page 366.

Surge Tank 7 (T-107)

Surge Tank 7 is a cone roof tank that holds the lean light phase exiting the liquid-liquid extractor to even out disturbances in the flow and provide a steady feed to the succeeding heat exchanger and decanter units. The tank has a capacity to store 500 ft^3, or 3,750 gallons, which allows for 5 minutes of residence time. Using a carbon steel construction, the total bare-module cost of this unit is $54,000. Please refer to the Surge Tank 7 Specification Sheet on page 237 and the Sample Calculations in Section E of the Appendix on page 366.

Surge Tank 8 (T-108)

Surge Tank 8 is a cone roof tank that holds the rich heavy phase exiting the liquid-liquid extractor to even out disturbances in the flow and provide a steady feed to the succeeding heat exchanger and storage units. The tank has a capacity to store 285 ft^3, or 2,130 gallons, which allows for 5 minutes of residence time. Using a carbon steel construction, the total bare-module cost of this unit is $41,000. Please refer to the Surge Tank 8 Specification Sheet on page 238 and the Sample Calculations in Section E of the Appendix on page 366.

Shift Storage Tank 9 (T-109)

Shift Storage Tank 9 is a floating roof tank that stores one shift worth of refined biodiesel product so that it may be analyzed for quality before being sent to the product storage tank (T-105) and on to consumers. The tank has a capacity to store 246,000 gallons of biodiesel, which is adequate storage for 8 hours of inventory. Using a steel alloy construction, the total bare-module cost of this unit is $1,953,000. Please refer to the Product Storage Tank 9 Specification Sheet on page 239 and the Sample Calculations in Section E of the Appendix on page 366.

Shift Storage Tank 10 (T-110)

Shift Storage Tank 10 is a floating roof tank that stores one shift worth of refined biodiesel product so that it may be analyzed for quality before being sent to the product storage tank (T-105) and on to consumers. The tank has a capacity to store 246,000 gallons of biodiesel, which is adequate storage for 8 hours of inventory. Using a steel alloy construction, the total bare-module cost of this unit is $1,953,000. Please refer to the Product Storage Tank 10 Specification Sheet on page 240 and the Sample Calculations in Section E of the Appendix on page 366.

Decanter 1 (S-101)

Decanter 1 is a liquid-liquid separation unit that removes 99.99% of the glycerol from the reactor 1 effluent at 100ºF, a sufficient amount to enable the transesterification reaction to go to completion in the second reactor. The decanter is modeled as a horizontal pressure vessel with a residence time of 30 minutes as suggested by industrial consultants. The total bare-module cost of this unit is $258,700. Please refer to the Decanter 1 Specification Sheet on page 241 and the Sample Calculations in Section E of the Appendix on page 366.

Decanter 2 (S-102)

Decanter 2 is a liquid-liquid separation unit that removes water from the crude biodiesel stream at 55ºF. The decanter removes 99.88% of the water in the biodiesel stream. The biodiesel exiting the decanter contains 0.049% water by volume, which less than the maximum 0.05% allowed in biodiesel fuel by ASTM D6751 standards. The decanter is modeled as a horizontal pressure vessel with a residence time of 30 minutes as suggested by industrial consultants. The total bare-module cost of this unit is $258,700. Please refer to the Decanter 2 Specification Sheet on page 242 and the Sample Calculations in Section E of the Appendix on page 366.

Reflux Accumulator 1 (S-103)

This unit, part of the distillation tower 1 apparatus, holds the liquid reflux leaving the condenser H-106 for a residence time of 5 minutes and is half-filled with liquid at steady state. Methanol and any air that leaked the column due to its vacuum pressure occupy the reflux accumulator vapor space. The reflux accumulator is modeled as a horizontal pressure vessel and has a total bare-module cost of $32,800. Please refer to the Reflux Accumulator 1 Specification Sheet on page 243 and the Sample Calculations in Section E of the Appendix on page 366.

Reflux Accumulator 2 (S-104)

This unit, part of the distillation tower 2 apparatus, holds the liquid reflux leaving the condenser H-108 for a residence time of 5 minutes and is half-filled with liquid at steady state. Methanol and any air that leaked the column due to its vacuum pressure occupy the reflux accumulator vapor space. The reflux accumulator is modeled as a horizontal pressure vessel and has a total bare-module cost of $28,900. Please refer to the Reflux Accumulator 2 Specification Sheet on page 244 and the Sample Calculations in Section E of the Appendix on page 366.

Liquid-Liquid Extractor 1 (L-101)

The liquid-liquid extractor is a washing unit that removes dissolved potassium and chloride ions as well as other impurities from the crude biodiesel stream. In the extractor, process water flows countercurrent against the liquid biodiesel stream and uptakes nearly 100% of the dissolved ions, glycerol and methanol and a substantial portion of unreacted diglycerides and monoglycerides. The bottom stream leaves the extractor as wastewater that is 1% organic material by mass and is sent to wastewater treatment. The liquid-liquid extractor is modeled as a rotating-disk contactor with an HETP of 4 ft and a throughput of 72 $ft^3/hr\text{-}ft^2$. The total bare-module cost of this unit is $2,586,800. Please refer to the Liquid-Liquid Extractor Specification Sheet on page 245 and the Sample Calculations in Section E of the Appendix on page 366.

Reactor 1 (R-101)

In Reactor 1, triglyceride and methanol feeds react to produce FAME and a glycerol byproduct at an overall conversion of 97.5%. The first reactor is comprised of 4 tubular vessels with lengths of 580 ft and diameters of 8 inches as is shown in Section D3 of the Appendix on page 287. At these dimensions, the Reynolds number for the feed stream, 1800, is small enough that the reactants approximate plug-flow conditions. Since the vessels are of considerable length, they are segmented into smaller pieces and arranged in a serpentine fashion with U-shaped connectors. The transesterification reactor was designed for a temperature of 140ºF, but the reaction is exothermic. To maintain the temperature within an acceptable range (80-180ºF), the triglyceride and methanol feed streams are each split into 4 streams, 3 of which are chilled to 10ºF.

When the methanol and triglycerides first enter the reactor, they begin to climb in temperature. Once the temperature becomes excessive, the next portion of methanol and triglcyeride feed is delivered to the reactor as a "cold shot," which sharply decreases the reactor temperature. The temperature climbs and dipsπ as additional cold shots are fed to the vessel. The resulting temperature profile may have a saw-tooth appearance, but overall, gives an average temperature of roughly 140ºF. Since the reaction kinetics will be faster than expected at temperatures above 140ºF and slower at temperatures below 140ºF, the two deviations are expected to balance each other, enabling the 140ºF reaction kinetics to be applied to the entire length of the vessel.

The reactor is sized based on the MATLAB results discussed in the Section D1 of the Appendix on page 284. The first cold shot is delivered at a distance of 32 ft, the second is delivered at 131 ft and the third at 230 ft, for a total length of 580 ft. The design calculations and temperature and concentration profiles in the tubular reactor are described in Section D2 of the Appendix on page 286. Based on a recommendation for the cost of pipe by Professor Fabiano, the total bare module cost for this unit is $1,525,000. Please refer to the Tubular Reactor Specification Sheet on page 246 and the Sample Calculations in Section E of the Appendix on page 366.

Reactor 2 (R-102)

In the second reactor, unreacted triglycerides, diglycerides and monoglycerides react with the excess methanol in the feed stream to bring the overall conversion to 99.97%. Since the majority of the feed stream is highly viscous FAME, plug flow may not be approximated in a tubular reactor. Therefore, an adiabatic, stirred tank reactor is employed, with an agitation rate of 300 rpm, a volume of 470 ft^3, a residence time of 6 minutes, and an operating temperature of 140ºF.[68] Since the extent of reaction is small in comparison to the large volume of reactants, the adiabatic temperature rise is only 3ºF according to the ASPEN PLUS simulation, an increase that is very tolerable. Please refer to the Stirred Tank Reactor Specification Sheet on page 247 and the Sample Calculations in Section E of the Appendix on page 366.

E. Specification Sheets

The following pages list the specification sheets that detail each unit in the lipid-processing transesterification module.

Distillation Tower

Identification:	Item:	RadFrac		Date: April 5, 2011
	Item No:	D-101		By: DC/SG/JI
	No. Required:	1		

Function: Removes methanol from the crude glycerol product stream

Operation: Continuous

Materials Handled:

		Inlet	Outlet	
Inlet Stream ID:		19	23	20
Quantity (lb/hr):		40,006	16,094	23,912
Composition:				
	GLYCEROL	22955	0	22,955
	METHANOL	15845	15,796	50
	OH-	5	0	5
	WATER	1188	298	890
	K+	12	0	12
	DIGLY-1	0.3	0	0
	DIGLY-2	0.2	0	0

	Inlet	Outlet	
Temperature (°F)	100	123	335
Pressure (PSI)	18	8	10
Vapor Fraction	0	0	0

Design Data:

Type:	Pressure Vessel
Material:	Carbon Steel
Height (ft):	48
Diameter (ft):	3
Number of Actual Trays:	12

C_P	$	69,700
C_{BM}	$	289,900

Distillation Tower

Identification:	Item:	RadFrac		Date: April 5, 2011
	Item No:	D-102		By: DC/SG/JI
	No. Required:	1		

Function: Removes methanol from the FAME product stream

Operation: Continuous

Materials Handled:

		Inlet	**Outlet**	
Inlet Stream ID:		31	32	36
Quantity (lb/hr):		234,851	8420	226431
Composition:				
	FAME-1	156,846	0	156,846
	GLYCEROL	571	0	571
	METHANOL	8,535	8,413	121
	FAME-2	43,726	0	43,726
	FAME-3	24,094	0	24,094
	OH-	300	0	300
	WATER	41	7	35
	K+	690	0	690
	DIGLY-1	11	0	11
	DIGLY-2	3	0	3
	DIGLY-3	2	0	2
	MONO-1	23	0	23
	MONO-2	6	0	6
	MONO-3	4	0	4

	Inlet	Outlet	
Temperature (°F)	144	122	428
Pressure (PSI)	10	8	10
Vapor Fraction	0	0	0

Design Data:

Type:	Pressure Vessel
Material:	Carbon Steel
Height (ft):	64
Diameter (ft):	2.75
Number of Actual Trays:	12

C_P	$	92,500
C_{BM}	$	384,800

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-101		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the KOH feed stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		KOH	1
	Quantity (lb/hr):		2,237	2,237
	Composition:			
		OH-	305	305
		WATER	1,230	1,230
		K+	702	702
		H3O+	0	0
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		0.2	
	Pressure (psi):		105	
	Efficiency:		0.7	
	C_P	$	4,800	
	C_{BM}	$	15,600	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-102		By: DC/SG/JI
	No. Required:	1		
Function:	Initiates movement of Methanol feed			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		METHANOL	3
	Quantity (lb/hr):		24,430	24,430
	Composition:			
		METHANOL	24,430	24,430
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		6.6	
	Pressure (psi):		105	
	Efficiency:		0.7	
	C_P	$	4,200	
	C_{BM}	$	13,600	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-103		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the triglyceride feed stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		TRIGLY	10
	Quantity (lb/hr):		223,678	223,678
	Composition:			
		TRIGLY-1	223,678	223,678
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		41	
	Pressure (psi):		95	
	Efficiency:		0.7	
	C_P	$	10,200	
	C_{BM}	$	33,500	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-104		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the chilled brine utility stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (lb/hr):		486266.2	486266.2
	Composition:			
		WATER	486266.2	486266.2
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		23	
	Pressure (psi):		50	
	Efficiency:		0.7	
	C_P	$	7,500	
	C_{BM}	$	24,800	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-105		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the cooling water utility stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (lb/hr):		1398121	1398121
	Composition:			
		WATER	1398121	1398121
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		201	
	Pressure (psi):		52	
	Efficiency:		0.7	
	C_P	$	35,900	
	C_{BM}	$	118,500	

Centrifugal Pump

Identification:	Item: Pump		Date: April 5, 2011
	Item No: P-106		By: DC/SG/JI
	No. Required: 1		
Function:	Increases the pressure of the feed stream to the second reactor		
Operation:	Continuous		

Materials Handled:

		Inlet	Outlet
Inlet Stream ID:		26	27
Quantity (lb/hr):		234851.4	234,851
Composition:			
	TRIGLY-1	5	5
	FAME-1	155,065	155,065
	GLYCEROL	2	2
	METHANOL	8,813	8,813
	FAME-2	43,230	43,230
	TRIGLY-2	1	1
	FAME-3	23,820	23,820
	TRIGLY-3	1	1
	OH-	300	300
	WATER	41	41
	K+	690	690
	DIGLY-1	1,086	1,086
	DIGLY-2	303	303
	DIGLY-3	168	168
	MONO-1	923	923
	MONO-2	258	258
	MONO-3	144	144

Design Data:

Type:	Centrifugal Pump
Material:	Cast Iron
Net Work Req (HP)	228
Pressure (psi):	520
Efficiency:	0.7

C_P	$	92,100
C_{BM}	$	303,900

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-107		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the feed stream to Distillation Tower D-101			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		17	18
	Quantity (lb/hr):		40,006	40,006
	Composition:			
		GLYCEROL	22,955	22,955
		METHANOL	15,845	15,845
		OH-	5	5
		WATER	1,188	1,188
		K+	12	12
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		19	
	Pressure (psi):		382	
	Efficiency:		0.7	
	C_P	$	11,600	
	C_{BM}	$	38,300	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-108		By: DC/SG/JI
	No. Required:	1		
Function:	Reflux Pump for Distillation Tower D-101			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (lb/hr):		1,609	1,609
	Composition:			
		METHANOL	1,580	1,580
		WATER	30	30
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		0.1	
	Pressure (psi):		36	
	Efficiency:		0.7	
	C_P	$	4,200	
	C_{BM}	$	13,800	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-109		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the bottoms stream from Distillation Tower D-101			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		20	21
	Quantity (lb/hr):		23,912	23,912
	Composition:			
		GLYCEROL	22,955	22,955
		METHANOL	50	50
		OH-	5	5
		WATER	890	890
		K+	12	12
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		1.5	
	Pressure (psi):		50	
	Efficiency:		0.7	
	C_P	$	3,600	
	C_{BM}	$	11,900	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-110		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the distillate stream from Distillation Tower D-101			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		23	24
	Quantity (lb/hr):		16,094	16,094
	Composition:			
		METHANOL	15,796	15796
		WATER	298	298
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		3	
	Pressure (psi):		80	
	Efficiency:		0.7	
	C_P	$	3,800	
	C_{BM}	$	12,400	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-111		By: DC/SG/JI
	No. Required:	1		

Function:	Increases the pressure of the distillate stream from Distillation Tower D-102
Operation:	Continuous

Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		32	33
	Quantity (lb/hr):		8,420	8,420
	Composition:			
		METHANOL	8,414	8,414
		WATER	6	6

Design Data:		
	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	1.6
	Pressure (psi):	80
	Efficiency:	0.7

C_P	$	3,600
C_{BM}	$	11,800

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-112		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the bottoms stream from Distillation Tower D-102			
Operation:	Continuous			

Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		36	37
	Quantity (lb/hr):		226431.4	226,431
	Composition:			
		FAME-1	156,846	156,846
		GLYCEROL	571	571
		METHANOL	121	121
		FAME-2	43,726	43,726
		FAME-3	24,094	24,094
		OH-	300	300
		WATER	35	35
		K+	690	690
		DIGLY-1	11	11
		DIGLY-2	3	3
		DIGLY-3	2	2
		MONO-1	23	23
		MONO-2	6	6
		MONO-3	4	4

Design Data:		
	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	29
	Pressure (psi):	67
	Efficiency:	0.7

C_P	$	8,600
C_{BM}	$	28,200

Centrifugal Pump

Identification:	Item:	Pump	Date: April 5, 2011	
	Item No:	P-113	By: DC/SG/JI	
	No. Required:	1		
Function:	Reflux Pump for Distillation Tower D-102			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (lb/hr):		842	842
	Composition:			
		METHANOL	841	841
		WATER	1	1
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		0.75	
	Pressure (psi):		36	
	Efficiency:		0.7	
	C_P	$	4,900	
	C_{BM}	$	16,200	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-114		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the HCl feed stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		HCL	41
	Quantity (lb/hr):		1,755	1,755
	Composition:			
		WATER	787	787
		CL-	553	553
		H3O+	297	297
		HCL	118	118
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		0.17	
	Pressure (psi):		63	
	Efficiency:		0.7	
	C_P	$	4,300	
	C_{BM}	$	14,100	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-115		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the process water utility stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		PROWATER	45
	Quantity (lb/hr):		205,458	205,458
	Composition:			
		WATER	205,458	205457.6
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		335	
	Pressure (psi):		985	
	Efficiency:		0.7	
	C_P	$	101,500	
	C_{BM}	$	334,800	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-116		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the chilled water feed stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (lb/hr):		701,693	701,693
	Composition:			
		WATER	701,693	701,693
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		48	
	Pressure (psi):		55	
	Efficiency:		0.7	
	C_P	$	11,900	
	C_{BM}	$	39,100	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-117		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of the lean phase leaving the Liquid-Liquid Extractor			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		49	50
	Quantity (lb/hr):		328,115	328,115
	Composition:			
		FAME-1	156843	156843
		FAME-2	43725	43725
		FAME-3	24093	24093
		WATER	103449	103449
		DIGLY-1	1.1	1.1
		DIGLY-2	0.3	0.3
		DIGLY-3	0.2	0.2
		MONO-1	2.4	2.4
		MONO-2	0.7	0.7
		MONO-3	0.4	0.4
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		37	
	Pressure (psi):		74	
	Efficiency:		0.7	
	C_P	$	9,800	
	C_{BM}	$	32,100	

Centrifugal Pump

Identification:	Item:	Pump	Date: April 5, 2011
	Item No:	P-118	By: DC/SG/JI
	No. Required:	1	
Function:	Increases the pressure of the waste-water stream moving to treatment		
Operation:	Continuous		

Materials Handled:

		Inlet	Outlet
Inlet Stream ID:		63	64
Quantity (lb/hr):		208830.3	208,830
Composition:			
	FAME-1	4	4
	GLYCEROL	571	571
	METHANOL	121	121
	FAME-2	1	1
	FAME-3	1	1
	WATER	206,714	206,711
	K+	690	690
	CL-	661	667
	H3O+	19	22
	HCL	6	0
	DIGLY-1	10	10
	DIGLY-2	3	3
	DIGLY-3	1	1
	MONO-1	20	20
	MONO-2	6	6
	MONO-3	3	3

Design Data:

Type:	Centrifugal Pump
Material:	Cast Iron
Net Work Req (HP)	16
Pressure (psi):	58
Efficiency:	0.7

C_P	$	6,300
C_{BM}	$	20,600

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-119		By: DC/SG/JI
	No. Required:	1		
Function:	Increases pressure of the process water recycle stream			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		59	60
	Quantity (lb/hr):		103322.8	103,323
	Composition:			
		WATER	103,323	103,323
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		170	
	Pressure (psi):		985	
	Efficiency:		0.7	
	C_P	$	68,500	
	C_{BM}	$	226,100	

Centrifugal Pump

Identification:	Item:	Pump	Date: April 5, 2011
	Item No:	P-120	By: DC/SG/JI
	No. Required:	1	

Function:	Increases the pressure of the biodiesel stream leaving the transesterification process and moving towards the shift tanks
Operation:	Continuous

Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		56	57
	Quantity (lb/hr):		224,793	224,793
	Composition:			
		FAME-1	156,842	156,842
		FAME-2	43,725	43,725
		FAME-3	24,093	24,093
		CL-	1	1
		H3O+	0.4	0.4
		DIGLY-1	1	1
		DIGLY-2	0.3	0.3
		DIGLY-3	0.2	0.2
		MONO-1	2.6	2.6
		MONO-2	0.7	0.7
		MONO-3	0.4	0.4

Design Data:		
	Type:	Centrifugal Pump
	Material:	Cast Iron
	Net Work Req (HP)	121
	Pressure (psi):	295
	Efficiency:	0.7

C_P	$	31,500
C_{BM}	$	103,800

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-121		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of Biodiesel product leaving shift tank T-109			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (lb/hr):		27,451	27,451
	Composition:			
		Biodiesel	27,451	27,451
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		1.5	
	Pressure (psi):		43	
	Efficiency:		0.7	
	C_P	$	3,700	
	C_{BM}	$	12,000	

Centrifugal Pump

Identification:	Item:	Pump		Date: April 5, 2011
	Item No:	P-122		By: DC/SG/JI
	No. Required:	1		
Function:	Increases the pressure of Biodiesel product leaving shift tank T-110			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	N/A
	Quantity (lb/hr):		27,451	27,451
	Composition:			
		Biodiesel	27,451	27,451
Design Data:				
	Type:		Centrifugal Pump	
	Material:		Cast Iron	
	Net Work Req (HP)		1.5	
	Pressure (psi):		43	
	Efficiency:		0.7	
	C_P	$	3,700	
	C_{BM}	$	12,000	

Centrifugal Blower

Identification:	Item:	Blower	Date: April 5, 2011
	Item No:	B-101	By: DC/SG/JI
	No. Required:	1	

Function:	Maintains vacuum pressure for distillation tower D-101
Operation:	Continuous

Materials Handled:			**Inlet**
	Inlet Stream ID:		N/A
	Quantity (lb/hr):		127
	Composition:		
		METHANOL	120
		O2	2
		N2	5

Design Data:		
	Type:	Centrifugal Blower
	Material:	Cast Iron
	Net Work Req (HP):	2.4
	Inlet Pressure (psi):	8
	Compression Ratio:	2

C_P	$	2,000
C_{BM}	$	4,300

Centrifugal Blower

Identification:	Item:	Blower	Date: April 5, 2011
	Item No:	B-102	By: DC/SG/JI
	No. Required:	1	

Function:	Maintains vacuum pressure for distillation tower D-102
Operation:	Continuous

Materials Handled:			**Inlet**
	Inlet Stream ID:		N/A
	Quantity (lb/hr):		87
	Composition:		
		METHANOL	79
		O2	2
		N2	6

Design Data:		
	Type:	Centrifugal Blower
	Material:	Cast Iron
	Net Work Req (HP):	1.75
	Inlet Pressure (psi):	8
	Compression Ratio:	2

C_P	$	1,600
C_{BM}	$	3,400

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011	
	Item No:	H-101		By: DC/SG/JI	
	No. Required:	1			
Function:	Increases the temperature of the feed stream to R-102				
Operation:	Continuous				

Materials Handled:		**Hot Stream**		**Cold Stream**	
Inlet Stream ID:		65	66	28	29
Quantity (lb/hr):		208,830	208,830	234851	234,851
Composition:					
	TRIGLY-1	0	0	5	5
	FAME-1	4	4	155,065	155,065
	GLYCEROL	571	571	2	2
	METHANOL	121	121	8,813	8,813
	FAME-2	1	1	43,230	43,230
	TRIGLY-2	0	0	1	1
	FAME-3	1	1	23,820	23,820
	TRIGLY-3	0	0	1	1
	OH-	0	0	300	300
	WATER	206,711	206,711	41	41
	K+	690	690	690	690
	CL-	667	667	0	0
	H3O+	22	22	0	0
	DIGLY-1	10	10	1,086	1,086
	DIGLY-2	3	3	303	303
	DIGLY-3	1	1	168	168
	MONO-1	20	20	923	923
	MONO-2	6	6	258	258
	MONO-3	3	3	144	144
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		166	150	100	140
Pressure (PSI)		30	23	47	40
Vapor Fraction		0	0	0	0

Design Data:		
	Type:	Shell and Tube Heat Exchanger
	Material:	Carbon Steel/Brass
	Heat Transfer Area (ft^2)	3635
	Heat Transfer Coefficient	25 (BTU/hr-ft^2-°F)
	Heat Duty (BTU/hr):	3,297,000
	C_P	$ 55,600
	C_{BM}	$ 176,200

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011		
	Item No:	H-102		By: DC/SG/JI		
	No. Required:	1				
Function:	Increases temperature of HCl feed stream					
Operation:	Continuous					
Materials Handled:			**Hot Stream**		**Cold Stream**	
	Inlet Stream ID:		38	39	42	43
	Quantity (lb/hr):		226,431	226,431	1754.7	1,755
	Composition:					
		FAME-1	156,846	156,846	0	0
		GLYCEROL	571	571	0	0
		METHANOL	121	121	0	0
		FAME-2	43,726	43,726	0	0
		FAME-3	24,094	24,094	0	0
		OH-	300	300	0	0
		WATER	35	35	1,057	729
		K+	690	690	0	0
		CL-	0	0	22	667
		H3O+	0	0	12	358
		HCL	0	0	664	1
		DIGLY-1	11	11	0	0
		DIGLY-2	3	3	0	0
		DIGLY-3	2	2	0	0
		MONO-1	23	23	0	0
		MONO-2	6	6	0	0
		MONO-3	4	4	0	0
			Inlet	Outlet	Inlet	Outlet
	Temperature (°F)		429	425	77	157
	Pressure (PSI)		42	35	38	31
	Vapor Fraction		0	0	0.2	0
Design Data:						
	Type:		Double-Pipe Heat Exchanger			
	Material:		Carbon Steel/Stainless Steel			
	Heat Transfer Area (ft^2)		50			
	Heat Transfer Coefficient		25 (BTU/hr-ft^2-°F)			
	Heat Duty (BTU/hr):		386,000			
	C_P	$	4,100			
	C_{BM}	$	12,900			

Heat Exchanger

Identification:	Item:	Heat Exchanger	Date: April 5, 2011
	Item No:	H-103	By: DC/SG/JI
	No. Required:	1	

Function: Increases the temperature of the process water stream entering the liq-liq extractor

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		39	40	47	48
Quantity (lb/hr):		226,431	226,431	308,779	308,779
Composition:					
	FAME-1	156,846	156,846	0	0
	GLYCEROL	571	571	0	0
	METHANOL	121	121	0	0
	FAME-2	43,726	43,726	0	0
	FAME-3	24,094	24,094	0	0
	OH-	300	300	0	0
	WATER	35	35	308,779	308,779
	K+	690	690	0	0
	DIGLY-1	10.8	10.8	0	0
	DIGLY-2	3.0	3.0	0	0
	DIGLY-3	1.7	1.7	0	0
	MONO-1	22.9	22.9	0	0
	MONO-2	6.4	6.4	0	0
	MONO-3	3.6	3.6	0	0
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		425	157	81	165
Pressure (PSI)		35	28	30	23
Vapor Fraction		0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	6960
Heat Transfer Coefficient	25 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	25,989,000

C_P	$	92,300
C_{BM}	$	292,600

Heat Exchanger

Identification:	Item:	Heat Exchanger	Date: April 5, 2011
	Item No:	H-104	By: DC/SG/JI
	No. Required:	1	

Function: Increases the temperature of the Biodiesel product stream

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		52	53	58	BIODIESE
Quantity (lb/hr):		328,115	328,115	224793	224,793
Composition:					
	FAME-1	156,843	156,843	156,843	156,843
	FAME-2	43,725	43,725	43,725	43,725
	FAME-3	24,093	24,093	24,093	24,093
	WATER	103,449	103,449	126	126
	CL-	1	1	1	0
	H3O+	0	0	0	0
	HCL	0	0	0	1
	DIGLY-1	1	1	1	1
	DIGLY-2	0.3	0.3	0.3	0
	DIGLY-3	0.2	0.2	0.2	0
	MONO-1	2	2	2	2
	MONO-2	1	1	1	1
	MONO-3	0.4	0.4	0.4	0
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		144	135	53.5639	77
Pressure (PSI)		39	32	21.7	15
Vapor Fraction		0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	1375
Heat Transfer Coefficient	15 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	1,522,000

C_P	$	30,300
C_{BM}	$	95,900

Heat Exchanger

Identification:	Item:	Heat Exchanger	Date: April 5, 2011
	Item No:	H-105	By: DC/SG/JI
	No. Required:	1	

Function: Increases the temperature of the process water recycle stream

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		51	52	61	62
Quantity (lb/hr):		328,115	328,115	103323	103,323
Composition:					
	FAME-1	156,843	156,843	0	0
	FAME-2	43,725	43,725	0	0
	FAME-3	24,093	24,093	0	0
	WATER	103,449	103,449	103323	103323
	CL-	1	1	0	0
	DIGLY-1	1	1	0	0
	DIGLY-2	0	0	0	0
	DIGLY-3	0	0	0	0
	MONO-1	2.4	2.4	0	0
	MONO-2	0.7	0.7	0	0
	MONO-3	0	0	0	0
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		166	144	53	89
Pressure (PSI)		46	39	47	40
Vapor Fraction		0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	1785
Heat Transfer Coefficient	25 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	3,715,000

C_P	$	35,100
C_{BM}	$	111,000

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011
	Item No:	H-106		By: DC/SG/JI
	No. Required:	1		

Function: Reflux condenser for distillation tower D-101

Operation: Continuous

Materials Handled:

		Hot Stream		**Cold Stream**	
Inlet Stream ID:		N/A	N/A	N/A	N/A
Quantity (lb/hr):		17,703	17,703	294,413	294,413
Composition:					
	METHANOL	17,495	17,495	0	0
	WATER	208	208	294,413	294,413
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		123	123	90	120
Pressure (PSI)		8	7	22	15
Vapor Fraction		0	0	0	0

Design Data:

Type:	Condenser
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	3,445
Heat Transfer Coefficient	200 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	8,832,000

C_P	$	53,200
C_{BM}	$	168,400

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011		
	Item No:	H-107		By: DC/SG/JI		
	No. Required:	1				

Function:	Kettle Reboiler for distillation tower D-101
Operation:	Continuous

Materials Handled:		**Cold Stream**		**Hot Stream**	
Inlet Stream ID:		N/A	N/A	N/A	N/A
Quantity (lb/hr):		23,912	23,912	15,038	15,038
Composition:					
	METHANOL	123	123	0	0
	WATER	3,927	3,927	15,038	15,038
	GLYCEROL	19,813	19,813	0	0
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		191	335	460	460
Pressure (PSI)		8	7	450	443
Vapor Fraction		0	0.5	1	0

Design Data:

Type:	Kettle Reboiler
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	225
Heat Transfer Coefficient	700 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	12,888,000

C_P	$	55,600
C_{BM}	$	176,200

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011		
	Item No:	H-108		By: DC/SG/JI		
	No. Required:	1				

Function: Reflux condenser for distillation tower D-102

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		N/A	N/A	N/A	N/A
Quantity (lb/hr):		9,262	9,262	150,881	150,881
Composition:					
	METHANOL	9,257	9,257	0	0
	WATER	5	5	150,881	150,881
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		122	122	90	120
Pressure (PSI)		8	7	22	15
Vapor Fraction		0	0	0	0

Design Data:

Type:	Condenser
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	1975
Heat Transfer Coefficient	200 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	4,526,000

C_P	$	36,900
C_{BM}	$	116,900

Heat Exchanger

Identification:	Item:	Heat Exchanger			Date: April 5, 2011	
	Item No:	H-109			By: DC/SG/JI	
	No. Required:	1				
Function:	Kettle Reboiler for distillation tower D-102					
Operation:	Continuous					
Materials Handled:			**Cold Stream**		**Hot Stream**	
	Inlet Stream ID:		N/A	N/A	N/A	N/A
	Quantity (lb/hr):		226,431	226,431	41,248	41,248
	Composition:					
		GLYCEROL	1,725	1,725	0	0
		METHANOL	1,052	1,052	41,248	41,248
		WATER	539	539	0	0
		FAME-1	147,171	147,171	0	0
		FAME-2	41,310	41,310	0	0
		FAME-3	24,784	24,784	0	0
		DIGLY-1	5	5	0	0
		DIGLY-2	1	1	0	0
		DIGLY-3	1	1	0	0
		MONO-1	18	18	0	0
		MONO-2	5	5	0	0
		MONO-3	3	3	0	0
			Inlet	Outlet	Inlet	Outlet
	Temperature (°F)		162	428	460	460
	Pressure (PSI)		9.9	9.8	450	443
	Vapor Fraction		0	0.0	1	0
Design Data:						
	Type:		Kettle Reboiler			
	Material:		Carbon Steel/Brass			
	Heat Transfer Area (ft^2)		2,760			
	Heat Transfer Coefficient		90 (BTU/hr-ft^2-°F)			
	Heat Duty (BTU/hr):		31,596,000			
	C_P	$	100,200			
	C_{BM}	$	317,400			

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011
	Item No:	C-101		By: DC/SG/JI
	No. Required:	1		

Function: Decreases the temperature of the methanol cold shot streams

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		7	8	N/A	N/A
Quantity (lb/hr):		42,632	42,632	74,076	74,076
Composition:					
	METHANOL	40,769	40,769	0	0
	OH-	254	254	0	0
	WATER	1,024	1,024	74,076	74,076
	K+	585	585	0	0
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		77	10	-30	10
Pressure (PSI)		65	58	22	15
Vapor Fraction		0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	660
Heat Transfer Coefficient	60 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	2,222,000

C_P	$	21,800
C_{BM}	$	68,900

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011		
	Item No:	C-102		By: DC/SG/JI		
	No. Required:	1				

Function: Decreases the temperature of the triglyceride cold shot streams

Operation: Continuous

Materials Handled:		**Hot Stream**		**Cold Stream**	
Inlet Stream ID:		13	14	N/A	N/A
Quantity (lb/hr):		42,632	42,632	74,076	74,076
Composition:					
	TRIGLY-1	130,085	130,085	0	0
	TRIGLY-2	36,264	36,264	0	0
	TRIGLY-3	19974.037	19974.04		
	WATER	1,024	1,024	74,076	74,076
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		77	10	-30	10
Pressure (PSI)		65	58	22	15
Vapor Fraction		0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	2,230
Heat Transfer Coefficient	35 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	4,409,000

C_P	$	40,200
C_{BM}	$	127,400

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011		
	Item No:	C-103		By: DC/SG/JI		
	No. Required:	1				
Function:	Decreases the temperature of the inlet stream to decanter S-101					
Operation:	Continuous					
Materials Handled:			**Hot Stream**		**Cold Stream**	
	Inlet Stream ID:		15	16	N/A	N/A
	Quantity (lb/hr):		274,857	274,857	34,859	34,859
	Composition:					
		TRIGLY-1	5	5	0	0
		FAME-1	155,065	155,065	0	0
		GLYCEROL	22,957	22,957	0	0
		METHANOL	24,659	24,659	0	0
		FAME-2	43,230	43,230	0	0
		TRIGLY-2	1.5	1.5	34,859	34,859
		FAME-3	23,820.1	23,820.1	0	0
		TRIGLY-3	0.8	0.8	0	0
		OH-	305.3	305.3	0	0
		WATER	1,229.5	1,229.5	0	0
		K+	701.9	701.9	0	0
		DIGLY-1	1,086.4	1,086.4	0	0
		DIGLY-2	303.0	303.0	0	0
		DIGLY-3	167.6	167.6	0	0
		MONO-1	923	923	0	0
		MONO-2	258	258	0	0
		MONO-3	144	144	0	0
			Inlet	Outlet	Inlet	Outlet
	Temperature (°F)		180	100	90	120
	Pressure (PSI)		53	46	22	15
	Vapor Fraction		0	0	0	0
Design Data:						
	Type:		Shell and Tube Heat Exchanger			
	Material:		Carbon Steel/Brass			
	Heat Transfer Area (ft^2)		13,110			
	Heat Transfer Coefficient		25 (BTU/hr-ft^2-°F)			
	Heat Duty (BTU/hr):		9,143,000			
	C_P	$	165,000			
	C_{BM}	$	523,000			

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011		
	Item No:	C-104		By: DC/SG/JI		
	No. Required:	1				
Function:	Decreases the temperature of the wastewater stream					
Operation:	Continuous					
Materials Handled:			**Hot Stream**		**Cold Stream**	
	Inlet Stream ID:		66	WASTE	N/A	N/A
	Quantity (lb/hr):		208,830	208,830	341,680	341,680
	Composition:					
		FAME-1	4	4	0	0
		GLYCEROL	571	571	0	0
		METHANOL	121	121	0	0
		FAME-2	1	1	0	0
		FAME-3	1	1	0	0
		WATER	206,711	206,711	341,680	341,680
		K+	690	690	0	0
		CL-	667	667	0	0
		H3O+	22	22	0	0
		DIGLY-1	10	10	0	0
		DIGLY-2	3	3	0	0
		DIGLY-3	1	1	0	0
		MONO-1	20	20	0	0
		MONO-2	6	6	0	0
		MONO-3	3	3	0	0
			Inlet	Outlet	Inlet	Outlet
	Temperature (°F)		150	100	90	120
	Pressure (PSI)		23	16	22	15
	Vapor Fraction		0	0	0	0
Design Data:						
	Type:		Shell and Tube Heat Exchanger			
	Material:		Carbon Steel/Brass			
	Heat Transfer Area (ft^2)		2,840			
	Heat Transfer Coefficient		200 (BTU/hr-ft^2-°F)			
	Heat Duty (BTU/hr):		10,250,000			
	C_P	$	46,600			
	C_{BM}	$	147,700			

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011
	Item No:	C-105		By: DC/SG/JI
	No. Required:	1		

Function: Decreases the temperature of the crude glcyerol product stream

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		22	GLYCEROL	N/A	N/A
Quantity (lb/hr):		23,912	23,912	116,932	116,932
Composition:					
	GLYCEROL	22,955	22,955	0	0
	METHANOL	50	50	0	0
	OH-	5	5	0	0
	WATER	890	890	116,932	116,932
	K+	12	12	0	0
	DIGLY-1	0.3	0.3	0	0
	DIGLY-2	0.2	0.2	0	0
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		335	100	90	120
Pressure (PSI)		24	17	22	15
Vapor Fraction		0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	350
Heat Transfer Coefficient	150 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	3,508,000

C_P	$	17,700
C_{BM}	$	55,900

Heat Exchanger

Identification:	Item:	Heat Exchanger	Date: April 5, 2011
	Item No:	C-106	By: DC/SG/JI
	No. Required:	1	

Function: Decreases the temperature of the crude biodiesel stream entering the decanter

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		53	54	N/A	N/A
Quantity (lb/hr):		328,115	328,115	180199	180199
Composition:					
	FAME-1	156,843	156,843	0	0
	FAME-2	43,725	43,725	0	0
	FAME-3	24,093	24,093	0	0
	WATER	103,449	103,449	0	0
	CL-	1	1	180199	180199
	DIGLY-1	1	1	0	0
	DIGLY-2	0	0	0	0
	DIGLY-3	0	0	0	0
	MONO-1	2	2	0	0
	MONO-2	1	1	0	0
	MONO-3	0	0	0	0
		Inlet	Outlet	Inlet	Outlet
Temperature (°F)		135	100	90	120
Pressure (PSI)		32	25	22	15
Vapor Fraction		0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area ft^2	4,940
Heat Transfer Coefficient	100 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	5,683,000

C_P	$	69,800
C_{BM}	$	221,200

Heat Exchanger

Identification:	Item:	Heat Exchanger		Date: April 5, 2011
	Item No:	C-107		By: DC/SG/JI
	No. Required:	1		

Function: Further decreases the temperature of the crude biodiesel stream entering the decanter S-102

Operation: Continuous

Materials Handled:

		Hot Stream		Cold Stream	
Inlet Stream ID:		54	55	N/A	N/A
Quantity (lb/hr):		329,379	329,379	144978	144977.8
Composition:					
	FAME-1	156,842	156,842	0	0
	FAME-2	43,725	43,725	0	0
	GLYCEROL	272,770	272,770	0	0
	FAME-3	24,093	24,093	0	0
	WATER	104,712	104,712	144978	144977.8
	CL-	1	1	0	0
	DIGLY-1	1	1	0	0
	DIGLY-2	0	0	0	0
	DIGLY-3	0	0	0	0
	MONO-1	3	3	0	0
	MONO-2	1	1	0	0
	MONO-3	0	0	0	0

	Inlet	Outlet	Inlet	Outlet
Temperature (°F)	100	55	40	90
Pressure (PSI)	18	11	22	15
Vapor Fraction	0	0	0	0

Design Data:

Type:	Shell and Tube Heat Exchanger
Material:	Carbon Steel/Brass
Heat Transfer Area (ft^2)	5,990
Heat Transfer Coefficient	100 (BTU/hr-ft^2-°F)
Heat Duty (BTU/hr):	7,249,000

C_P	$	81,300
C_{BM}	$	257,700

Feed Storage Tank

Identification:	Item:	Tank		Date: April 5, 2011
	Item No:	T-101		By: DC/SG/JI
	No. Required:		1	
Function:	Stores 5 days of Potassium Hydroxide Feed			
Operation:	Continuous			
Materials Handled:			**Outlet**	
	Inlet Stream ID:		KOH	
	Quantity (lb/hr):		2,237	
	Composition:			
		OH-	305	
		WATER	1,230	
		K+	702	
Design Data:				
	Type:		Floating Roof	
	Material:		Steel Alloy	
	Volume (ft^3)		3,010	
	C_P	$	78,800	
	C_{BM}	$	289,000	

Feed Storage Tank

Identification:	Item:	Tank	Date: April 5, 2011
	Item No:	T-102	By: DC/SG/JI
	No. Required:	1	
Function:	Stores 5 days of Methanol Feed		
Operation:	Continuous		
Materials Handled:		**Outlet**	
	Inlet Stream ID:	METHANOL	
	Quantity (lb/hr):	24,430	
	Composition:		
	METHANOL	24,430	
Design Data:			
	Type:	Floating Roof	
	Material:	Steel Alloy	
	Volume (ft^3)	59,900	
	C_P	$	363,000
	C_{BM}	$	1,329,000

Feed Storage Tank

Identification:	Item:	Tank		Date: April 5, 2011
	Item No:	T-103		By: DC/SG/JI
	No. Required:	2		
Function:	Stores 2 days of Triglyceride Feed			
Operation:	Continuous			
Materials Handled:			**Outlet**	
	Inlet Stream ID:		TRIGLY	
	Quantity (lb/hr):		223,678	
	Composition:			
		TRIGLY-1	156,165	
		TRIGLY-2	43,535	
		TRIGLY-3	23,978	
Design Data:				
	Type:		Floating Roof	
	Material:		Steel Alloy	
	Volume (ft^3)		133,680	
	C_P	$	1,091,000	
	C_{BM}	$	3,993,000	

Product Storage Tank

Identification:	Item:	Tank	Date: April 5, 2011
	Item No:	T-104	By: DC/SG/JI
	No. Required:	1	

Function:	Stores 5 days of Crude Glycerol Product
Operation:	Continuous

Materials Handled:			**Inlet**
	Inlet Stream ID:		GLYCEROL
	Quantity (lb/hr):		23,912
	Composition:		
		GLYCEROL	22,955
		METHANOL	50
		OH-	5
		WATER	890
		K+	12
		DIGLY-1	0.3
		DIGLY-2	0.2

Design Data:		
	Type:	Floating Roof
	Material:	Steel Alloy
	Volume (ft^3)	37,350

C_P	$	285,000
C_{BM}	$	1,043,000

Product Storage Tank

Identification:	Item: Tank		Date: April 5, 2011
	Item No: T-105		By: DC/SG/JI
	No. Required: 4		
Function:	Stores 5 days of Biodiesel Product		
Operation:	Continuous		
Materials Handled:		**Inlet**	
	Inlet Stream ID:	BIODIESE	
	Quantity (lb/hr):	224,793	
	Composition:		
	FAME-1	156,843	
	FAME-2	43,725	
	FAME-3	24,093	
	WATER	126	
	CL-	0.1	
	H3O+	0.0	
	HCL	0.7	
	DIGLY-1	1.1	
	DIGLY-2	0.3	
	DIGLY-3	0.2	
	MONO-1	2.4	
	MONO-2	0.7	
	MONO-3	0.4	
Design Data:			
	Type:	Floating Roof	
	Material:	Steel Alloy	
	Volume (ft^3)	133,680	
	C_P	$ 2,181,500	
	C_{BM}	$ 7,985,000	

Feed Storage Tank

Identification:	Item:	Tank		Date: April 5, 2011
	Item No:	T-106		By: DC/SG/JI
	No. Required:	1		
Function:	Stores 5 days of Hydrochloric Acid Feed			
Operation:	Continuous			
Materials Handled:			**Outlet**	
	Inlet Stream ID:		HCL	
	Quantity (lb/hr):		1,755	
	Composition:			
		WATER	787	
		CL-	553	
		H3O+	297	
		HCL	118	
Design Data:				
	Type:		Floating Roof	
	Material:		Steel Alloy	
	Volume (ft^3)		2,970	
	C_P	$	78,300	
	C_{BM}	$	287,000	

Surge Tank

Identification:	Item:	Tank		Date: April 5, 2011
	Item No:	T-107		By: DC/SG/JI
	No. Required:	1		
Function:	Stores Liquid-Liquid Extractor Light Phase Surge			
Operation:	Continuous			

Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		N/A	49
	Quantity (lb/hr):		328,115	328,115
	Composition:			
		FAME-1	156,843	156,843
		FAME-2	43,725	43,725
		FAME-3	24,093	24,093
		FAME-3	24,093	24,093
		HCL	1	1
		DIGLY-1	1	1
		DIGLY-2	0	0
		DIGLY-3	0.2	0.2
		MONO-1	2	2
		MONO-2	1	1
		MONO-3	0.4	0.4

Design Data:		
	Type:	Cone Roof
	Material:	Carbon Steel
	Volume (ft^3)	500

C_P	$	17,700
C_{BM}	$	54,000

Surge Tank

Identification:	Item:	Tank		Date: April 5, 2011
	Item No:	T-108		By: DC/SG/JI
	No. Required:	1		

Function: Stores Liquid-Liquid Extractor Heavy Phase Surge

Operation: Continuous

Materials Handled:

		Inlet	Outlet
Inlet Stream ID:		N/A	63
Quantity (lb/hr):		208,830	208,830
Composition:			
	FAME-1	4	4
	GLYCEROL	571	571
	METHANOL	121	121
	FAME-2	1	1
	FAME-3	1	1
	WATER	206714	206714
	K+	690	690
	CL-	661	661
	H3O+	19	19
	HCL	6	6
	DIGLY-1	10	10
	DIGLY-2	3	3
	DIGLY-3	1	1
	MONO-1	20	20
	MONO-2	6	6
	MONO-3	3	3

Design Data:

Type:	Floating Roof
Material:	Carbon Steel
Volume (ft^3)	285

C_P	$	13,300
C_{BM}	$	41,000

Product Shift Tank

Identification:	Item:	Tank	Date: April 5, 2011
	Item No:	T-109	By: DC/SG/JI
	No. Required:	1	

Function:	Stores 8 hours of biodiesel product for analysis
Operation:	Continuous

Materials Handled:			**Inlet**
	Inlet Stream ID:		BIODIESE
	Quantity (lb/hr):		224,793
	Composition:		
		FAME-1	156,843
		FAME-2	43,725
		FAME-3	24,093
		WATER	126
		CL-	0.1
		H3O+	0.0
		HCL	0.7
		DIGLY-1	1.1
		DIGLY-2	0.3
		DIGLY-3	0.2
		MONO-1	2.4
		MONO-2	0.7
		MONO-3	0.4

Design Data:		
	Type:	Floating Roof
	Material:	Steel Alloy
	Volume (ft^3)	32,840

C_P	$	553,500
C_{BM}	$	1,953,000

Product Shift Tank

Identification:	Item: Tank		Date: April 5, 2011
	Item No: T-110		By: DC/SG/JI
	No. Required: 1		
Function:	Stores 8 hours of biodiesel product for analysis		
Operation:	Continuous		
Materials Handled:		**Inlet**	
	Inlet Stream ID:	BIODIESE	
	Quantity (lb/hr):	224,793	
	Composition:		
	FAME-1	156,843	
	FAME-2	43,725	
	FAME-3	24,093	
	WATER	126	
	CL-	0.1	
	H3O+	0.0	
	HCL	0.7	
	DIGLY-1	1.1	
	DIGLY-2	0.3	
	DIGLY-3	0.2	
	MONO-1	2.4	
	MONO-2	0.7	
	MONO-3	0.4	
Design Data:			
	Type:	Floating Roof	
	Material:	Steel Alloy	
	Volume (ft^3)	32,840	
	C_P $	553,500	
	C_{BM} $	1,953,000	

Decanter

Identification:	Item:	Flash		Date: April 5, 2011
	Item No:	S-101		By: DC/SG/JI
	No. Required:	1		
Function:	Separates glycerol from reactor R-101 effluent			
Operation:	Continuous			

Materials Handled:		**Inlet**	**Outlet**	
Inlet Stream ID:		16	17	26
Quantity (lb/hr):		274,857	40,006	234,851
Composition:				
	TRIGLY-1	5	0	5
	FAME-1	155,065	0	155,065
	GLYCEROL	22,957	22,955	2
	METHANOL	24,659	15,845	8,813
	FAME-2	43,230	0	43,230
	TRIGLY-2	1	0	1
	FAME-3	23,820	0	23,820
	TRIGLY-3	1	0	1
	OH-	305	5	300
	WATER	1,230	1,188	41
	K+	702	12	690
	DIGLY-1	1,086	0	1,086
	DIGLY-2	303	0	303
	DIGLY-3	168	0	168
	MONO-1	923	0	923
	MONO-2	258	0	258
	MONO-3	144	0	144
		Inlet	Outlet	
Temperature (°F)		100	100	100
Pressure (PSI)		46	43	43
Vapor Fraction		0	0	0

Design Data:

Type:	Pressure Vessel
Material:	Carbon Steel
Length (ft):	9.75
Diameter (ft):	40

C_P	$	84,900
C_{BM}	$	258,700

Decanter

Identification:	Item:	Flash		Date: April 5, 2011
	Item No:	S-102		By: DC/SG/JI
	No. Required:		1	

Function: Separates water from liquid-liquid extractor light phase

Operation: Continuous

Materials Handled:

		Inlet	**Outlet**	
Inlet Stream ID:		55	56	59
Quantity (lb/hr):		328,115	224793	103,323
Composition:				
	FAME-1	156,843	156,843	0
	FAME-2	43,725	43,725	43,725
	FAME-3	24,093	24,093	0
	WATER	103,449	126	103,323
	CL-	1	1	0
	DIGLY-1	1.1	1	0
	DIGLY-2	0	0.3	0
	DIGLY-3	0.2	0	0
	MONO-1	2.4	2	0
	MONO-2	1	1	0
	MONO-3	0	0.4	0

	Inlet	Outlet	
Temperature (°F)	55	55	55
Pressure (PSI)	18	15	15
Vapor Fraction	0	0	0

Design Data:

Type:	Pressure Vessel
Material:	Carbon Steel
Length (ft):	9.75
Diameter (ft):	40

C_P	$	84,900
C_{BM}	$	258,700

Reflux Accumulator

Identification: Item: Flash — Date: April 5, 2011
Item No: S-103 — By: DC/SG/JI
No. Required: 1

Function: Accumulates reflux for distillation tower D-101

Operation: Continuous

Materials Handled:

		Inlet	**Outlet**	
Inlet Stream ID:		N/A	N/A	18
Quantity (lb/hr):		17,703	1,609	16,094
Composition:				
	METHANOL	17,375	1,580	15,796
	WATER	328	30	298
		Inlet	Outlet	
Temperature (°F)		123	123	123
Pressure (PSI)		8	7	7
Vapor Fraction		0	0	0

Design Data:

Type:	Pressure Vessel
Material:	Carbon Steel
Length (ft):	2.75
Diameter (ft):	5.5

C_P	$	10,800
C_{BM}	$	32,800

Reflux Accumulator

Identification:	Item:	Flash		Date: April 5, 2011
	Item No:	S-104		By: DC/SG/JI
	No. Required:	1		

Function: Accumulates reflux for distillation tower D-102

Operation: Continuous

Materials Handled:

		Inlet	**Outlet**	
Inlet Stream ID:		N/A	N/A	18
Quantity (lb/hr):		9,262	842	8,420
Composition:				
	METHANOL	9,255	841	8,414
	WATER	7	1	6

	Inlet	Outlet	
Temperature (°F)	122	122	122
Pressure (PSI)	8	7	7
Vapor Fraction	0	0	0

Design Data:

Type:	Pressure Vessel
Material:	Carbon Steel
Length (ft):	2.25
Diameter (ft):	4.5

C_P	$	9,500
C_{BM}	$	28,900

Liquid-Liquid Extractor

Identification:	Item:	Extract		Date: April 5, 2011
	Item No:	L-101		By: DC/SG/JI
	No. Required:		1	

Function: Removes potassium and hydroxide ions from crude biodiesel stream

Operation: Continuous

Materials Handled:

		Inlet		Outlet	
Inlet Stream ID:		44	48	63	49
Quantity (lb/hr):		228,186	308,779	208,830	328,115
Composition:					
	FAME-1	156,846	0	4	156,843
	GLYCEROL	571	0	571	0
	METHANOL	121	0	121	0
	FAME-2	43,726	0	1	43,725
	TRIGLY-2	0	0	0	0
	FAME-3	24,094	0	1	24,093
	WATER	1,403	308,779	206,714	103,449
	K+	690	0	690	0
	CL-	661	0	661	0
	H3O+	19	0	19	0
	HCL	7	0	6	1
	DIGLY-1	11	0	10	1
	DIGLY-2	3	0	3	0
	DIGLY-3	2	0	1	0
	MONO-1	23	0	20	2
	MONO-2	6	0	6	1
	MONO-3	4	0	3	0

	Inlet		Outlet	
Temperature (°F)	166	165	165	165
Pressure (PSI)	23	23	15	15
Vapor Fraction	0	0	0	0

Design Data:

Type:	Pressure Vessel
Material:	Carbon Steel
Diameter (ft):	13.50
Length (ft):	66

C_P	$	621,900
C_{BM}	$	2,586,800

Tubular Reactor

Identification:	Item:	Rstoic		Date: April 5, 2011		
	Item No:	R-101		By: DC/SG/JI		
	No. Requir	1				
Function:	Converts triglyceride feed to FAME product					
Operation:	Continuous					
Materials Handled:			**Inlet**			**Outlet**
	Inlet Stream ID:	12	9	14	8	15
	Quantity (lb/hr):	37,354	8,547	186,324	42,632	274,857
	Composition:					
	TRIGLY-1	26,080	0	130,085	0	5
	FAME-1	0	0	0	0	155,065
	GLYCEROL	0	0	0	0	22,957
	METHANOL	0	8,173	0	40,769	24,659
	FAME-2	0	0	0	0	43,230
	TRIGLY-2	7,270	0	36,264	0	1
	FAME-3	0	0	0	0	23,820
	TRIGLY-3	4,004	0	19,974	0	1
	OH-	0	51	0	254	305
	WATER	0	205	0	1,024	1,230
	K+	0	117	0	585	702
	DIGLY-1	0	0	0	0	1,086
	DIGLY-2	0	0	0	0	303
	DIGLY-3	0	0	0	0	168
	MONO-1	0	0	0	0	923
	MONO-2	0	0	0	0	258
	MONO-3	0	0	0	0	144
			Inlet			Outlet
	Temperature (°F)	81.440883	76	10	10	180
	Pressure (PSI)	64.7	65	57.7	58	52.7
Design Data:						
	Type:	Tubular Reactor (with cold shots of feed)				
	Material:	Stainless Steel				
	Number of Parallel Tubes	4				
	Tube Diameter (in):	8				
	Tube Length (ft):	580				
	Max. Reynolds Number:	1800				
	C_P	\$ 500,000				
	C_{BM}	\$ 1,525,000				

Stirred Tank Reactor

Identificatior	Item:	Rstoic	Date: April 5, 2011	
	Item No:	R-102	By: DC/SG/JI	
	No. Required:	1		
Function:	Converts unreacted triglycerides, diglycerides and monoglycerides to FAME product			
Operation:	Continuous			
Materials Handled:			**Inlet**	**Outlet**
	Inlet Stream ID:		29	30
	Quantity (lb/hr):		234,851	234,851
	Composition:			
		TRIGLY-1	5	0
		FAME-1	155,065	156,846
		GLYCEROL	2	571
		METHANOL	8,813	8,535
		FAME-2	43,230	43,726
		TRIGLY-2	1	0
		FAME-3	23,820	24,094
		TRIGLY-3	1	0
		OH-	300	300
		WATER	41	41
		K+	690	690
		DIGLY-1	1,086	11
		DIGLY-2	303	3
		DIGLY-3	168	2
		MONO-1	923	23
		MONO-2	258	6
		MONO-3	144	4
			Inlet	Outlet
	Temperature (°F)		140	144
	Pressure (PSI)		40	35
Design Data:				
	Type:		Constantly Stirred Tank Reactor	
	Material:		Carbon Steel	
	Stirrer Speed (rpm)		300	
	Volume (ft^3)		470	
	Heat Duty:		Adiabatic	
	C_P	$	231,100	
	C_{BM}	$	961,000	

3. Transesterification: Feasibility & Economic Analysis

What follows is a thorough economic analysis of the catalytic transesterification lipid-processing module design as it has been presented above. An overall economic analysis for the algae-to-biofuel venture that incorporates the results from Module III is discussed later in the report.

A. Fixed Capital Investment Summary

Table XXXVII summarizes the equipment costs for the transesterification process. The purchase cost represents the purchase price for the physical machinery, while the bare-module cost indicates the total costs including installation. The purchase costs were determined using the various correlations listed and explained in Chapter 22 of *Product and Process Design Principles*. The bare-module factor is specific to each type of equipment and may be found in Table 22.11 of that same text.[67]

Table XXXVII: Equipment Cost Summary for Transesterification

Designation	Equipment Description	Purchase Cost ($)	Bare Module Factor	Bare Module Cost ($)
DISTILLATION TOWERS				
D-101	Distillation Tower 1	69,700	4.16	289,900
D-102	Distillation Tower 2	92,500	4.16	384,800
PUMPS				
P-101	Centrifugal Pump 1	4,800	3.30	15,600
P-102	Centrifugal Pump 2	4,200	3.30	13,600
P-103	Centrifugal Pump 3	10,200	3.30	33,500
P-104	Centrifugal Pump 4	7,500	3.30	24,800
P-105	Centrifugal Pump 5	35,900	3.30	118,500
P-106	Centrifugal Pump 6	92,100	3.30	303,900
P-107	Centrifugal Pump 7	11,600	3.30	38,300
P-108	Centrifugal Pump 8	4,200	3.30	13,800
P-109	Centrifugal Pump 9	3,600	3.30	11,900
P-110	Centrifugal Pump 10	3,800	3.30	12,400
P-111	Centrifugal Pump 11	3,600	3.30	11,800
P-112	Centrifugal Pump 12	8,600	3.30	28,200
P-113	Centrifugal Pump 13	4,900	3.30	16,200
P-114	Centrifugal Pump 14	4,300	3.30	14,100

P-115	Centrifugal Pump 15	101,500	3.30	334,800
P-116	Centrifugal Pump 16	11,900	3.30	39,100
P-117	Centrifugal Pump 17	9,800	3.30	32,100
P-118	Centrifugal Pump 18	6,300	3.30	20,600
P-119	Centrifugal Pump 19	68,500	3.30	226,100
P-120	Centrifugal Pump 20	31,500	3.30	103,800
P-121	Centrifugal Pump 21	3,700	3.30	12,000
P-122	Centrifugal Pump 22	3,700	3.30	12,000
P-Spare	Spare Pumps	321,800	3.30	1,061,700
BLOWERS/COMPRESSORS				
B-101	Vacuum Blower 1	2,000	2.15	4,300
B-102	Vacuum Blower 2	1,600	2.15	3,400
HEAT EXCHANGERS				
H-101	Heat Exchanger 1	55,600	3.17	176,200
H-102	Heat Exchanger 2	4,100	3.17	12,900
H-103	Heat Exchanger 3	92,300	3.17	292,600
H-104	Heat Exchanger 4	30,300	3.17	95,900
H-105	Heat Exchanger 5	35,100	3.17	111,000
H-106	Tower 1 Condenser	53,200	3.17	168,400
H-107	Tower 1 Reboiler	55,600	3.17	176,200
H-108	Tower 2 Condenser	36,900	3.17	116,900
H-109	Tower 2 Reboiler	100,200	3.17	317,400
C-101	Chilled Brine HX 1	21,800	3.17	68,900
C-102	Chilled Brine HX 2	40,200	3.17	127,400
C-103	CW HX 3	165,000	3.17	523,000
C-104	CW HX 4	46,600	3.17	147,700
C-105	CW HX 5	17,700	3.17	55,900
C-106	CW HX 6	69,800	3.17	221,200
C-107	Chilled Water HX 7	81,300	3.17	257,700
STORAGE TANKS				
T-101	KOH Storage Tank	78,800	3.05	289,000
T-102	Methanol Storage Tank	363,000	3.05	1,329,000
T-103	Triglyceride Storage Tanks	1,090,800	3.05	3,993,000
T-104	Glycerol Storage Tank	285,000	3.05	1,043,000
T-105	Biodiesel Storage Tanks	2,181,500	3.05	7,985,000
T-106	HCl Storage Tank	78,300	3.05	287,000
T-107	Top L1 Hold Up Tank	17,700	3.05	54,000
T-108	Bot L1 Hold Up Tank	13,300	3.05	41,000
T-109	Day Storage Tanks	266,800	3.05	977,000
T-110	Day Storage Tanks	266,729	3.05	977,000
SEPARATORS				
S-101	Decanter 1	84,900	3.05	258,700
S-102	Decanter 2	84,900	3.05	258,700

S-103	Reflux Accumulator 1	10,800	3.05	32,800
S-104	Reflux Accumulator 2	9,500	3.05	28,900
L-101	Liq-Liq Extractor 1	621,900	4.16	2,586,800
REACTORS				
R-101	Reactor 1	500,000	3.05	1,525,000
R-102	Reactor 2	231,100	4.16	961,000

Total	**$**	**28,534,000**

The Fixed Capital Investment Summary, summarized below in table XXXVIII, details the capital investment required for the transesterification process. The direct permanent investment, C_{DPI}, is the sum of the total bare-module equipment costs, the site preparation costs, the service facilities costs and allocated costs.[67] From table XXXVII, the total Bare-Module Cost is $28,534,000, which includes the cost of spares, storage tanks and process machinery. The cost of site preparation and service facilities are estimated to cost 5.0% of the total-bare module cost.[67] Utilities are assumed to be readily available onsite and no allocated cost is expected to be incurred. The total depreciable capital, C_{TDC}, accounts for contingencies and contractor's fees, estimated at 18% of the direct permanent investment. The total permanent investment, C_{TPI}, sums non-depreciable investments with the total depreciable capital. Land costs and royalties are both assumed to be roughly 2% of the C_{TDC}, while startup costs are typically 10% of the C_{TDC}.[67]

To complete the evaluation of the total capital investment, it is necessary to determine the working capital, which is the difference between current assets and liabilities. Working capital is typically calculated by summing 30 days of cash reserves, which cover the cost of manufacture, 30 days of accounts receivable, which cover the month long delay in receipt of payment, and inventory, which is assumed to be 5 days at the sales price.[67] From that amount, the accounts payable, which allows for 30 days of payment for all feedstocks, is subtracted to give the working capital. Although working capital for the transesterification process is $90,462,000, a significant sum due to the high value products and large sales volume, this value is recuperated at the end of the plant life.[67] The working capital is added to the C_{TPI} to give a total capital investment of $132,686,000.

Table XXXVIII: Fixed Capital Investment Summary for the Transesterification Process

C_{TBM}	**Total Bare Module Cost**		
C_{PM}	Equipment Bare Module Costs	$	10,497,000
C_{spare}	Total Bare-Module Costs for Spares	$	1,063,000
$C_{storage}$	Total Bare-Module Costs for Storage Tanks	$	16,975,000
	C_{TBM}	$	28,534,000
C_{DPI}	**Direct Permanent Investment**		
C_{TBM}	Total Bare-Module Cost	$	28,534,000
C_{site}	Site Preparation	$	1,427,000
C_{serv}	Service Facilities	$	1,427,000
C_{alloc}	Allocated Costs for Utility Plants	$	0
	C_{DPI}	$	31,388,000
C_{TDC}	**Total Depreciable Capital**		
C_{DPI}	Direct Permanent Investment	$	31,388,000
C_{cont}	Contingencies and Contractor's Fees	$	5,650,000
	C_{TDC}	$	37,038,000
C_{TPI}	**Total Permanent Investment**		
C_{TDC}	Total Depreciable Capital	$	37,038,000
C_{land}	Cost of Land	$	741,000
C_{royal}	Cost of Royalties	$	741,000
$C_{startup}$	Cost of Plant Startup	$	3,704,000
	C_{TPI}	$	42,224,000
C_{WC}	**Working Capital**		
	Cash Reserves	$	5,651,000
	Inventory	$	12,683,000
	Accounts Receivable	$	76,098,000
	Accounts Payable	$	3,970,000
		$	90,462,000
C_{TCI}	**Total Capital Investment**		
C_{TPI}	Total Permanent Investment	$	42,224,000
C_{WC}	Working Capital	$	90,462,000
	C_{TCI}	**$**	**132,686,000**

B. Operating Costs

Introduction

The objective of this analysis is to determine the operating costs associated with the lipid-processing module, which are comprised of both variable and fixed costs. In contrast with the fixed capital investment summary, which concerned an initial investment that must only be made during plant startup, the operating costs are annual expenditures that must be made for the duration of the plant life. Therefore, minimizing operating costs is essential to improving the overall economics of the algae-to-biofuel venture, which was based on a designed output of 225,000 lb/hr of biodiesel fuel, equivalent to 17,560 bpd. The biodiesel fuel produced meets ASTM D6751 industrial fuel standards and may be sold as pure B100 fuel.[84] According to the Chemical News and Intelligence (ICIS) Service, the current sale price of B100 is approximately \$3.30/gal, up from \$3.00/gal one year ago.[93] Based on this price, annual revenue from the transesterification process is \$802,531,000. The crude glycerol byproduct of transesterification may also be sold to generate additional income. Based on a recommendation from Bruce M. Vrana, an industrial consultant from DuPont who proposed the 2011 CBE 459 design project entitled "Glycerol to Renewable Propylene Glycol" crude glycerol is expected to sell at \$0.22/lb.

A summary of the variable costs associated with lipid-processing, which include the purchase of raw materials and utilities in addition to other general expenses, are presented in table XXXIX below. Since a triglyceride feedstock is supplied from Module II, the raw material costs only include potassium hydroxide, hydrochloric acid and methanol and total \$51,428,000. General expenses involved selling/transfer expenses, direct research, allocated research, administrative expenses and management incentive compensation and total \$102,527,000 according to the estimations provided in Chapter 23 of *Product and Process Design Principles*.[67] Utilities cost \$10,023,000 annually, while the sale of crude glycerol not used for algal fermentation provides an extra \$38,195,000 in annual revenue. Together, the total annual variable costs are \$125,783,000.

TABLE XXXIX: Variable Costs of the Transesterification Process

			Annual Cost
General Expenses			
	Selling/Transfer Expenses	$	26,630,000
	Direct Research	$	42,608,000
	Allocated Research	$	4,439,000
	Administrative Expense	$	17,754,000
	Management Incentive Compensation	$	11,096,000
		$	**102,527,000**
Feedstock (raw materials)			
	Methanol	$	44,653,000
	Hydrochloric Acid	$	909,000
	Potassium Hydroxide	$	5,866,000
		$	**51,428,000**
Utilities			
	Electricity	$	509,000
	High Pressure Steam	$	2,385,000
	Medium Pressure Steam	$	633,000
	Cooling Water	$	113,000
	Chilled Water	$	265,000
	Process Water	$	161,000
	Chilled Brine	$	485,000
	Wastewater Treatment	$	5,472,000
		$	**10,023,000**
Byproducts			
	Excess Crude Glycerol	$	38,195,000
		$	**(38,195,000)**
Total Variable Costs		**$**	**125,783,000**

Unlike variable costs, which scale with production rate, fixed costs are expenses that are charged regardless of production rate, such as operating costs, maintenance costs, operating overhead, property taxes and insurance.[67] These fixed costs are summarized in table XXXX. These expenses may be estimated from the C_{TDC} and the number of process sections using the guidelines described in section 23.2 of *Product and Design Principles*.[64] The cost of operations was calculated based on a need for 6 operators per shift, the number required to safely operate 3 large continuous-flow process sections.[67] Maintenance costs include wages and benefits for employees that maintain equipment in proper working order as well as providing for trained supervisors, materials and services and maintenance overhead. The operating overhead includes costs not directly related to plant operating, such as medical and safety services, and may be estimated as a fraction of the combined salary, wages and benefits for maintenance and labor-related operations.[67] Together, the total fixed costs sum to $7,462,000.

TABLE XXXX: Fixed Costs of the Transesterification Process

Operations		
Direct Wages and Benefits	$	1,840,000
Direct Salaries and Benefits	$	276,000
Operating Supplies and Services	$	17,000
Technical assistance to manufacturing	$	360,000
Control laboratory	$	390,000
	$	**2,883,000**
Maintenance		
Wages and Benefits	$	1,297,000
Salaries and Benefits	$	325,000
Materials and Services	$	1,297,000
Maintenance Overhead	$	65,000
	$	**2,984,000**
Operating Overhead		
General Plant Overhead	$	266,000
Mechanical Department Services	$	90,000
Employee Relations Department	$	221,000
Business Services	$	277,000
	$	**854,000**
Property Taxes and Insurance	**$**	**741,000**
Total Fixed Costs	**$**	**7,462,000**

C. Other Important Considerations

Safety and Environmental

The transesterification process has some potential risks that must be carefully evaluated. Methanol, a key reactant that is present in both the liquid and vapor phases in the lipid-processing module, is a poisonous chemical that may be fatal or cause blindness if swallowed and cause harm to the respiratory tract if inhaled.[94] Furthermore, methanol is very flammable in both phases. Methanol vapor present in the distillation tower system is evacuated from the reflux accumulator vapor space by a centrifugal blower and is sent to a flare for safe disposal. Although minor air leaks are tolerated, a serious leak of air into the vacuum distillation towers, which are operate at moderately high temperatures, could result in an oxygen concentration high enough to permit a serious explosion. Monitors and sensors should be used to detect and address any leaks that may occur. The transesterification reactor also contains methanol at high temperature, but is under pressure, so air leaks are not as much of a concern for this vessel and the chance for explosion is reduced. Care should be taken to limit contact with methanol liquid and vapor throughout the plant.

Concentrated potassium hydroxide and hydrochloric acid are other reagents that pose serious safety hazards. Careful neutralization procedures should be well planned and rehearsed to prepare for a significant spill or acid or base. Both may cause serious burns and debilitation.[95,96]

Crude Glycerol

Glycerol is an important byproduct produced during transesterification and has many applications in the food and pharmaceutical industries.[85] As biodiesel production ramps and global glycerol supplies increase, the price commanded for crude glycerol has begun to fall. However, as enterprising chemical engineers develop novel processes that make use of this side product, the price may steady and enable biodiesel manufacturers to increase their revenues from glycerol sales. As discussed in the "Preliminary Process Synthesis," the crude glycerol byproduct from this algae-to-biodiesel venture will serve as an energy source for *C. protothecoides* during their fermentative phase. However, only 20% of the glycerol product is required for use in in Module 1 as a feedstock. The remaining 80% will be distributed to consumers, where it may be used as a crude raw material or refined into a more pure and valuable product suitable for a vaster array of applications. Crude glycerol is expected to sell at \$0.22/lb based on a recommendation from Bruce M. Vrana, an industrial consultant from DuPont who proposed the 2011 CBE 459 design project entitled "Glycerol to Renewable Propylene Glycol."

VII. OVERALL ECONOMIC ANALYSIS

A rigorous profitability analysis was conducted for the overall algae-to-biodiesel venture. The detailed and technical report may be found in Section F of the Appendix on page 379, while the key results are summarized below in table XXXXI. Based on the current market price of $3.30 per gallon for pure biodiesel, a project life of 15 years, and a 15% discount rate, the results indicate that an algae-to-biodiesel process may not only be profitable, but also a sound and reasonable investment. The projected Net Present Value (NPV) is on the order of $1.3 billion, the Return on Investment (ROI) may be as high as 32%, and the Investor's Rate of Return (IRR) was calculated to be 35%. The results in the table shown below correspond to the third year of production, during which the algae-to-biodiesel venture is assumed to operate at 90% of full capacity. Since the alternative fuels industry matures at a rapid pace, this profitability analysis was also conducted based on a project life of 10 years in case a 15 year project life was too optimistic. This decision was based on a recommendation made by Dr. Warren Seider from the University of Pennsylvania and Mr. David Kolesar, of the Dow Chemical Company. The exact results for the 10 and 15 year scenarios are only marginally different while the overall economic trends are very similar, meaning that this project is not overly sensitive to the projected lifetime.

Table XXXXI: ROI Analysis for Third Year of Production

Annual Sales	$	839,336,000
Annual Costs	$	(132,893,000)
Annual Variable Costs (excluding byproducts)	$	(349,716,000)
Annual Byproducts	$	393,798,000
Annual Fixed Costs	$	(164,555,000)
Depreciation	$	(88,988,000)
Income Tax	$	(228,459,000)
Net Earnings	$	338,997,000
Total Capital Investment	$	1,197,742,000
ROI		**32%**

Table XXXXII details the overall fixed capital investment summary for the overall algae-to-biodiesel venture and indicates that capital costs, estimated at $1.2 billion, are high, but reasonable considering the large annual sales of roughly $840 million. It is important to note that the total capital investment is not equally distributed to the three modules, but rather disproportionately stems from algal cultivation in Module I. The total bare-module cost for Module I, estimated at $600 million dollars, is approximately 20 times the total bare-module cost for Modules II and III, both of which are roughly $30 million dollars. This breakdown is unequal due to the high capital costs required to purchase the massive fermentation tanks and land required for the PFM model of algal cultivation. This high capital cost does not result in unreasonable depreciation or annual fixed costs, which are only 30% of annual revenue.

Table XXXXII: Overall Fixed Capital Investment Summary

C_{TBM}	Total Bare Module Cost	$	779,060,000
C_{DPI}	Direct Permanent Investment	$	856,966,000
C_{TDC}	Total Depreciable Capital	$	1,011,219,000
C_{TPI}	Total Permanent Investment	$	1,112,341,000
C_{TCI}	**Total Capital Investment**	**$**	**1,197,742,000**

Annual variable costs for the overall algae-to-biodiesel venture are presented below in table XXXXIII. It is key to note that although the variable costs sum to a net negative value, the utilities and feedstock costs are quite significant. However, the variable costs excluding the byproduct, on the order of $350 million, is offset by income generated from byproduct sales.

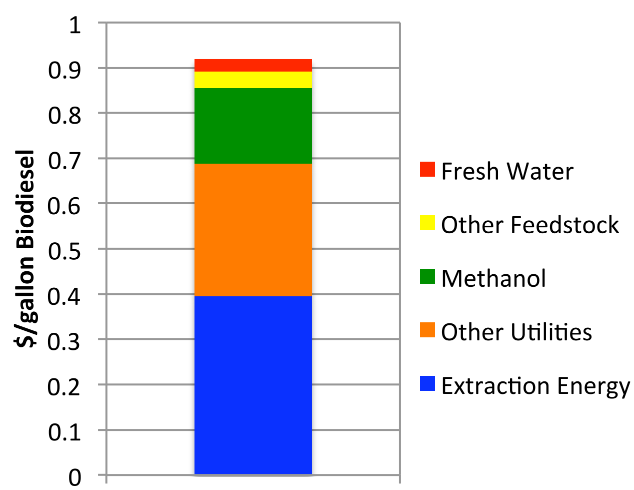


Figure 28 The variable cost breakdown for the overall algae-to-biodiesel venture

An additional layer of information may be garnered by breaking the variable cost down into its individual components, as shown in Figure 28. The graph indicates that the variable costs are most sensitive to electricity consumption, which is primarily associated with lipid-extraction. Module II accounts for more than 60% of the overall electricity consumption alone. To reduce the overall variable costs, future design teams should consider examining this source of electrical energy consumption with the intention of improving efficiency and optimizing its design.

Table XXXXIII: Annual Variable Costs for the Overall Algae to Biodiesel Process

General Expenses	102,525,000
Feedstock (raw materials)	54,801,000
Utilities	192,390,000
Byproducts	(393,798,000)
Total Variable Costs	**(44,083,000)**

As shown in figure *29*, the biomass comprises a significant portion of the revenue per gallon of biodiesel sold. Yet, the value of this byproduct is subject to wide variability. As described in Module II, algal biomass is not currently sold in large volumes as an animal feedstock. Although its value may be assessed based on a comparison with other protein sources, such as soy meal, its adoption by farmers as a safe feedstock is yet uncertain.[82] However, given that the annual algal biomass produced is only on the order of 1.5% of the demand for animal feedstock in the United States, its introduction as feed is unlikely to overwhelm or saturate the market.[83] It is therefore expected to sell readily, either as pure feed or an ingredient in a feedstock blend.

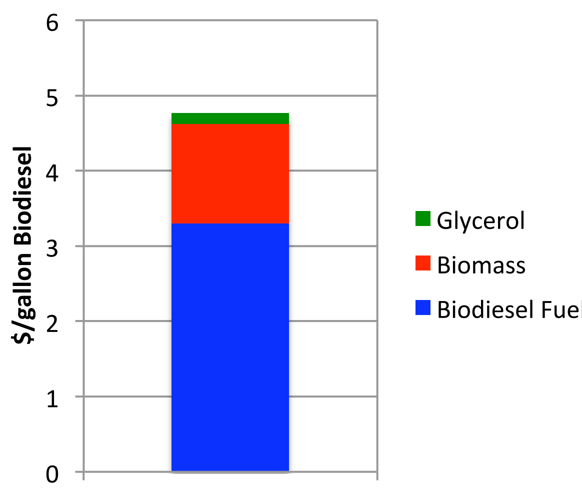


Figure 29 The revenue breakdown for the overall algae-to-biodiesel venture

Given both the uncertainty and magnitude associated with the revenue generate from algal biomass, it was suspected that the overall profitability of the algae-to-biodiesel venture would be highly sensitive to biomass sales. Therefore, a sensitivity analysis was conducted and is presented in table XXXXIV below.

Table XXXXIV: Sensitivity Analysis of Change in Biomass Price

Biomass Price ($/lb)	$0.00	$0.010	$0.039	$0.068	**$0.097**	$0.126	$0.165	$0.184	$0.194
IRR	20.0%	21.7%	26.5%	31.0%	**35.2%**	39.2%	44.3%	46.8%	48.1%
ROI	14.8%	16.6%	21.9%	27.2%	**32.5%**	37.8%	44.9%	48.4%	50.2%

As can be seen in the table above, the values of both the IRR and ROI vary significantly with the price of algal biomass; should the price of algal biomass decrease by 30%, the IRR decreases by 12%, while the ROI falls by 16%. Therefore, ensuring a high sale price for biomass is key to maximizing profitability. However, even if biomass fails to launch on the market as an animal feed and cannot be sold for an appreciable value, the IRR and ROI remain at roughly 20% and 15% respectively. This result is favorable because it indicates that the algae-to-biodiesel venture is profitable enough that it may be economically feasible regardless of the revenues that may or may not be generated from biomass sales.

Similarly, given the importance of the revenue generated from biodiesel sales, the main source of revenue for the algae-to-biodiesel venture, a second sensitivity analysis was conducted based on the biodiesel price and is presented in table XXXXV below. For the purposes of this profitability analysis, it was assumed that the price of biodiesel would increase 2.5% annually, the same rate as inflation, as recommended by *Product and Process Design*.[67]

Table XXXXV: Sensitivity Analysis of Change in Biodiesel Price

Biodiesel Price ($/gal)	$1.98	$2.31	$2.64	$2.97	**$3.30**	$3.63	$3.96	$4.29	$4.62
IRR	20.4%	24.5%	28.3%	31.8%	**35.2%**	38.4%	41.5%	44.4%	47.3%

Just as for the price of algal biomass, the IRR is also very sensitive to the biodiesel price, which is similarly uncertain. In 2010, pure biodiesel sold for $3.00/gal, while its current 2011 price is $3.23.[93] As legislation encourages the adoption of alternative fuels and the price of oil increases, it is unknown how the demand for biodiesel will shift and affect its price. If the price dramatically increases, then the IRR could become greater than 40% as shown in the table above, making the algae-to-biodiesel venture very worthy of further investigation. Yet, even if the price falls as low as $1.98, the venture will still be profitable and yield an IRR of 20.4%.

The annual fixed costs for the overall algae to biodiesel project involve costs that are not a function of the fuel production rate, such as operating costs, maintenance costs, operating overhead, property taxes and insurance, and other annual expenses and are shown in table XXXXVI below.

Table XXXXVI: Fixed Costs for the Overall Algae to Biodiesel Process

Operations	24,505,000
Maintenance	104,661,000
Operating Overhead	15,164,000
Property Taxes and Insurance	20,224,000
Total Fixed Costs	**164,555,000**

Operations costs are based on a total workforce of 21 operators for the three modules as previously discussed. Maintenance costs include wages and benefits for employees that maintain equipment in proper working order as well as providing for trained supervisors, materials and services and maintenance overhead. Is it clear from the table above that maintenance costs account for a significant portion of the total fixed costs. Yet, because maintenance costs are calculated as a fraction of the total depreciable capital as recommended by *Product and Process Design*, this result was expected.[67] The operating overhead includes costs not directly related to plant operating, such as medical and safety services, and may be estimated as a fraction of the combined salary, wages and benefits for maintenance and labor-related operations.[67] Together, the total fixed costs sum to $164,555,000.

Other Economic Considerations

Government Subsidies and Incentives

To encourage their citizens to reduce dependence on fossil fuel derived sources of transportation fuel, many governments, including that of the United States, have supported initiatives that make alternative fuel production more financially attractive. The U.S. Government passed a $1.00/gallon tax credit for FAME production in 2004, a subsidy that significantly benefits biodiesel producers.[110] This incentive has since been extended multiple times, most recently in

December, 2010. These incentives have helped accelerate the rate at which the biodiesel industry grows. In 2004, prior to the introduction of tax-based incentives, the US was producing only 25 million gallons of biodiesel per year. By 2009, the US production of biodiesel sharply increased to 545 million gallons annually, marking an increase of more than 2000%. However, the political climate is subject to significant variation, and such tax credits may not be permanent. Investors in biodiesel fuel should be wary of relying on such subsidies for the overall algae-to-biodiesel venture to be profitable.[110]

Carbon Trading Credit

Another potential stream of revenue for the algae-to-biodiesel process stems from carbon credit trading. Emissions-trading involves the assignment of a specified number of carbon credits to carbon-emitting businesses. This allows companies to emit the amount of CO_2 that corresponds to the number of carbon credits they possess without incurring any penalty.[98] However, should that credit limit be exceeded, a penalty is issued. In contrast, should a company emit less CO_2 than what they have been allotted, it may sell those unused credits at the current market price. Since algal cultivation involves the consumption of CO_2 from flue gas during photosynthesis, the algae-to-biodiesel process may be linked to the carbon-trading market.[98]

CO_2 credits are sold as contracts, each corresponding to 1000 metric tons of CO_2. As described in Module 1, the cultivation process consumes 0.142 lbs of CO_2 per second or 2030 metric tons annually. The amount of CO_2 removed from flue gas annually via cultivation is equivalent to two contracts, which may be sold to CO_2 producing business that seek additional carbon credits. The spot price of carbon credits is very dynamic, ranging from $0.05 to $2.75 per metric ton. Yet, even at $2.75 per metric ton, the revenue generated from carbon credits only amounts to $5,600 annually. Therefore, the carbon credit price is currently too low to result in the accumulation of any appreciable revenue.

Based on the proposed algal cultivation site of Thompsons, Texas, the carbon credits for process will be sold in the Chicago Climate Exchange (CCX).[97] Although the market price for carbon in the CCX is extremely low as described above, there is potential for carbon trading to become a significant source of revenue in the future. For example, the Western Climate Initiative is a carbon trading program currently under development that involves seven US states and 4

Canadian provinces. To date, it has distributed $21 billion dollars of carbon allowances at an expected credit trading price of $33/ton.[98] At this price, the annual revenue from carbon trading may approach $70,000, and could amount to a greater sum should the carbon credit trading price experience further increases in the next decade.[98]

Comparison to Catalytic Hydrotreating

A comprehensive CBE 459 design report, entitled "Algae to Alkanes" by Carlson, Lee, Oje and Xu, evaluated an algae-to-n-alkane venture that was comprised of the same three modules presented in this report.[6] Since the 2010 report involved the design and economic analysis of the catalytic hydrotreatment of algal lipids, a key difference between this process and the one presented in this report is the lipid-processing module. As discussed extensively in Module III, catalytic hydrotreating is an alternative lipid-processing module that consumes a triglyceride feedstock and outputs n-alkanes that may be sold as diesel fuel. Since catalytic hydrotreating only affects the lipid-processing fraction of the overall algae-to-biodiesel venture, the metrics associated with Module III in both processes may be directly compared. The results are presented in table XXXXVII. As is clear from table below, the Fixed Costs, Variable Costs and Total Permanent Investments required for hydrotreating and transesterification are very similar. The two lipid-processing facilities barely differ in economic scope. The clear distinction between the two alternative lipid-treatment pathways involves the products and byproducts and their associated sales.

Table XXXXVII: An Economic Comparision of Hydrotreating and Transesterification

Metric	Transesterification	Catalytic Hydrotreating
Product	Biodiesel (FAME)	Green Diesel (n-alkanes)
Mass Production Rate (lb/hr)	223,000	223,000
Vol. Production Rate (gal/hr)	33,963	32,230
Product Price ($/gal)	3.30	3.02
Annual Sales ($)	839,336,000	788,387,000
Module III Total Permanent Inv. ($)	42,224,000	41,402,000
Module III Var Costs (Excl. Bypdts) ($)	163,978,000	164,941,000
Module III Fixed Costs ($)	7,462,000	7,582,000
Byproducts	Glycerol	Fuel Gas, H_2, Propane
Byproduct Sales ($)	38,195,000	63,340,000

While biodiesel currently sells for $3.30, Carlson et al. (2010) considered the profitability of n-alkanes at a price of only $3.00.[6] In contrast, the sale of glycerol yields only $38 million while the sale of hydrotreating byproducts reportedly adds an additional $63 million in revenues.[6] Because the fixed costs, variable costs and total permanent investments for these two processes are similar and would be expected to change in the same fashion with market variation, the prices that transportation fuel producers are able to command for biodiesel, n-alkanes, glycerol, propane, fuel gas and hydrogen will be key in determining the economic feasibility of these two processes. Should the price of diesel rise, hydrotreaters will likely gain the economic edge, but if government subsidies and tax incentives continue to reduce expenses for FAME production, then biodiesel companies may be able to operate with more favorable finances.

VIII. CONCLUSION & RECOMMENDATIONS

This report evaluated the technical and economic potential of a variety of recently proposed methodologies for the culture and processing of algae to produce biofuel. Although several uncertainties in the market for biodiesel and its associated byproducts, thoroughly discussed in the preceding report, the overall economic analysis indicates that the algae-to-biodiesel venture may be profitable with a return of investment (ROI) on the order of 32% and an Investor's Rate of Return (IRR) at roughly 35%. What follows are the major conclusions drawn from the analysis conducted for each of the three modules, and a series of recommendations for future investigation.

The process design and economic analysis performed by the design team in 2010 suggested that a significant obstacle to the commercialization of the algae-to-biofuel venture was the staggering cost of land required for cultivation, which they estimated at $2.2 billion (Carlson et al., 2010). In contrast, the results presented in Module I indicate that the cost of land is on the order of $129 million. This is a significant departure from previous results and indicates that both adopting the novel autotrophic-heterotrophic growth model and increasing the diameter of the cultivation tubing allowed for a critical improvement in the optimization of land usage and field acreage. Despite these advances, the current design requires the consumption of vast volumes of fresh water, which is both scarce and expensive throughout the world, especially in Texas. This high fresh water usage, however, is an unavoidable consequence of growing *Chlorella protothecoides*, which is a fresh water algae. Future designs could make significant reductions in the operating costs alone by cultivating algae strains are tolerant of saline or even waste water.

Since OriginOil™ is just beginning to commercialize and license its Single-Step Extraction process, most of the technical data that characterizes their technologies are unavailable to the public. Therefore, it remains difficult to fully evaluate the claims that the process developed by OriginOil™ results in energy savings. However, based on the available data, it was determined that the cost of lipid-extraction via Single-Step Extraction technology was on the order of $0.06/lb, an appreciable difference from the $0.56/lb that OriginOil™ estimates as the cost for the conventional solvent or mechanical extraction processes. As more information regarding the Single-Step Extraction process becomes publically available, a more thorough process design and economic analysis may be conducted. Furthermore, the results presented in this report assume that the biomass byproduct sells as animal feedstock at a price equivalent to soybean meal, generating a large amount of additional revenue. However, given the uncertainty of this revenue stream as discussed previously, further analysis of the viability of this byproduct in the animal feed market must be performed.

In recent years, the science of transesterification has been well-studied and there are significant data available for the transformation of various vegetable oils into methyl and ethyl esters. However, there is little public information regarding oil derived from algal sources. Although it is possible to generate models, such as the Aspen PLUS simulation thoroughly discussed in this report, the kinetics used to model the lipid-processing module were based on data collected for palm oil at similar conditions. Until additional kinetic data that characterizes the transesterification of algal oil is published, the steady-state conversions reported remain very close approximations. Furthermore, estimating the properties of the triglycerides, diglycerides, monoglycerides and fatty acid methyl esters consumed and produced during transesterification using group contribution models in Aspen PLUS was challenging. Improvements in the accuracy of these predictions are necessary to improve confidence in the simulation results.

An overall economic analysis performed for the three modules of the algae-to-biodiesel venture indicated that the ROI and IRR are on the order of 32% and 35% respectively. Sensitivity analyses suggest that the process may still be economically feasible with a positive ROI and IRR even if algal biomass or biodiesel fuel decrease in price. A comparison of transesterification and catalytic hydrotreating lipid-processing modules was also performed and revealed that given their similarity in operating costs and required capital investment, the more economically favorable process is dictated by the current market prices for their respective products and byproducts .

Despite the fact that the preceding economic analysis portrayed the algae-to-biodiesel venture as a potentially attractive investment, in order to convince investors to support such a venture, it is critical to move beyond theoretical modeling. Pilot studies must be conducted to determine whether the science and economics discussed in this report can be verified with experimental data. Once the three modules have been experimentally analyzed and their associated models refined, a more accurate economic analysis may be conducted, thereby pushing algae one step closer to becoming a significant source of alternative energy in the 21st century.

IX. ACKNOWLEDGEMENTS

We would like to thank our professors at the University of Pennsylvania and industrial consultants for the essential assistance they provided us throughout this semester with our project. We would like to especially thank Mr. John Wismer of Arkema, Inc., for originally recommending this project; as well as Mr. Bruce Vrana, of Dupont Energy Technology; Mr. Gary Sawyer, of Lyondell Chemical Company; and Mr. David Kolesar, from the Dow Chemical Company.

Dr. Wen K. Shieh was an invaluable resource that spent countless hours donating his wisdom and experience regarding the cultivation of algae. His feedback was critical for the design of key equipment in both Modules I and II.

Professor Warren Seider, who guided us throughout the semester, ensured that we achieved our project goals, met with us repeatedly to offer help, and consistently provided insightful advice. Professor Seider possesses a wealth of information regarding technologies applicable to the algae-to-biofuel venture and directed us to valuable technical papers relevant to our project. His enduring support and patience with us was critical to our successful performance.

Professor Stuart Churchill generously hosted us for a productive afternoon at his home, where we had an opportunity to discuss our concerns, particularly with regards to the fluid dynamic principles crucial for the design of the algal cultivation module and the tubular transesterification reactor.

Professor Leonard Fabiano spent hours of his time helping us refine our Aspen PLUS simulation of the tranesterification process, and his advice was particularly crucial for successful property estimation. Without him, the simulation we have presented in Module III would not be possible. He also greatly assisted us greatly with equipment sizing and costing. His weekly reassurance that we were on the right track was much appreciated.

VII. APPENDIX

A. Problem Statement

4. Algae to Biodiesel
(recommended by Warren D. Seider, UPenn, Kimberly Ogden, UArizona)

During the Spring 2010, a design project [1] focused on the growth of algae using the SIMGAE cultivation process [2-4], (2) OriginOil's process for lipid extraction [9-11], and (3) a process to convert the lipids to alkanes (green diesel) [15-16]. A profitable overall process was designed, although the investment costs were very high. Since then, a research project to convert algae to biofuels by the National Alliance for Advanced Biofuels and Bioproducts (NAABB) was funded by DOE ($49 million). The Univ. Arizona and Penn are collaborating on a portion of this project. Furthermore, some of the major oil companies (e.g., ExxonMobil) have announced significant development efforts. The advantages of algae are numerous: (1) its cultivation does not encroach on the food sector, (2) its biomass productivity per acre far exceeds that of any agricultural commodity, (3) it produces lipids that can be converted easily to biodiesel or fuel-range hydrocarbons.

In this design project, the focus will be on exploring the processing technologies described below, and likely other promising technologies that are proposed in the next few months.

Algae Cultivation

Recently, the *heteroboost* photosynthesis-fermentation process was proposed to generate lipids for biodiesel production [5]. First, *chlorella prototecoides* algae are grown autotrophically to fix CO_2. Then these are metabolized heterotrophically using glucose to significantly increase the lipid yield.

Yet, another cultivation process was recently proposed by Teymour and coauthors [6] involving a farm of artificial algae trees whose trunk and branches are filled with algae solution and exposed to solar energy.

More recently, a kinetic model has been proposed to estimate the rate of algae growth as a function of the light intensity and the concentration of nitrogen nutrients [7]. This model can be used to better estimate the size and cost of various conversion processes, including photobioreactors. Also, data taken in "raceways" at the Univ. of Arizona are showing the impact of temperature on the algae growth rate [8].

Based upon the data in [5-8], better estimates of the installation and operating costs of cultivation processes, as compared with SIMGAE process costs, should be obtained. It is anticipated that lower-cost processes will be designed.

Lipid Extraction

The OriginOil process involves the generation of sonic waves and microbubbles at high frequency [9-11]. Energy costs are claimed to be 10% of conventional processing costs. An objective of this design project will be to gain better estimates of these costs – although this aspect of the design is likely to be deemphasized.
Lipid to Biodiesel

The conversion of the lipid fraction to useful products has several options [12-14]. One approach is to convert it to biodiesel [15-16]. Recently, two papers discuss the ASPEN PLUS simulation of a potential process to produce biodiesel [17-18]. Your group will seek to improve upon these designs and will attempt to carry out design optimization. This will include sizing and costing the trans-esterification reactor and the remainder of the processing equipment.

Life-Cycle Analysis

When estimating the profitability of your entire process to convert algae to biodiesel, your group will carry out a life-cycle analysis (LCA) [19]. To achieve a sustainable design, it is important that no net CO_2 be produced, that waste water be recycled, and the like. You will attempt to optimize the profitability of your design, as well as its sustainability.

References

Algae Cultivation

1) Carlson, L.S., Lee, M.Y., Oje, C.A.E., and A. Xu, *Algae to Alkanes*, SEAS Library, Univ. of Pennsylvania, 2010.

2) Hassania, J. (Diversified Energy) "Simgae – Low Cost, Broad Application – Algal Biomass Production System" , NETL 2008 Conference Proceedings, Sept. 18, 2008 (Presentation on Simgae technology with economics) at www.netl.doe.gov/publications/proceedings/08/.../index.html

3) Editor, "Simgae – Low Cost Algae Prod. Sys.", Aug, 2007, http//www.eneverve.com (contains optimistic claims for project economics)

4) Cloud, G. et al., US2008/0311649 A1, "Pressurized Flexible Tubing System for Producing Algae", Dec 18, 2008, XL Renewables

5) Xiong, W., Gao, C., Yan, D., Wu, C., and Q. Wu, "Bouble CO2 Fixation in Photosynthesis-Fermentation Model Enhances Algal Lipid Synthesis for Biodiesel Production, *Bioresource Tech.*, **101**, 2287-2293 (2010).

6) Eltayeb, A.E., Khalil, O., Al-Hallaj, S., and F. Teymour, "Design and Modeling of Optical Modules for Use in the Emerald Forest Algae Photobioreactors," *Comput. Chem. Eng.*, **34** 1323-1340 (2010).

7) Packer, A., Li, Y. Andersen, T., Hu, Q., Kuang, Y., and M. Sommerfeld, " Growth and Neutral Lipid Synthesis in Green Microalgae: A Mathematical Model," *Bioresource Tech.*, in print.

8) Waller, P., "Raceway Developments and Temperature Effects for Arid Climates,"

POWER POINT presentation, July 2010.

Lipid Isolation

9) Chementator News Item, "A One-Step Process for Extracting Oil from Algae", *Chem. Eng.*, June 2009

10) Eckelberry, N. and T.R, "Algae Growth System for Oil Production", US2009/0029445 A1, Jan. 29, 2009 (describes OriginOil's first-generation extraction technology)

11) OriginOil.com – Website contains video of lipid extraction process

Lipid Processing

12) Huber, G.W., O'Connor, P., and A. Corma, "Processing Biomass in Conventional Oil Refineries," *Applied Catalysis A*, **329**, 120-129 (2007).

13) Stiefel, S., and G. Dassori, "Simulation of Biodiesel Production through Transesterification of Vegetable Oils," *Ind. Eng. Chem. Res.*, **48**, 1068-1071 (2009).

14) Moser, B.R., "Biodiesel Production, Properties, and Feedstocks," *In Vitro Cell Dev. Biol.*, **45**, 229-266 (2009).

15) Machacek, M. T. et al., "Continuous Algal Biodiesel Production Facility", US2009/0071064 A1, March 19, 2009 (describes ASPEN PLUS Simulation of lipids to biodiesel)

16) Doviak , S., Long, M., Peck, S., and J. Yuen, *Green Diesel Fuel: A Biofuel Process*, SEAS Library, Univ. of Pennsylvania, 2008.

17) Chung, A.-F., and Y.A. Lui, "Integrated Process Modeling and Product Design of Biodiesel Manufacturing," *Ind. Eng. Chem. Res.*, **49**, 1197-1213 (2010).

18) Pokoo-Aikins, G., Nadim, A., El-Halwagi, M.M., and V. Mahalec, "Design and Analysis of Biodiesel Production from Algae Grown Through Carbon Sequestration," *Clean Techn. Environ. Policy*, **12**, 239-254 (2010).

LCA

19) Aresta, M. et al., "Utilization of macro-algae for enhanced CO_2 fixation and biofuels production: Development of a computing software for an LCA Study, "*Fuel Processing Technology*" 86 (2005), 1679-1693.

B. Module I Calculations

Table A1: The Recipe & Cost of the Guillard f/2 Formula (Adapted from Carlson et al., 2010)

Trace Metal Solution	Amt. per L Water	Stock Concentration	Purchase Price ($/g)	Cost ($)
FeCL3.6H2O	1.3 g	--	0.106 $/g	$0.137
Na2EDTA.2H2O	8.7 g	--	0.104 $/g	$0.906
CuSO4.5H2O	1 mL	9.8 mg/ml dH2O	0.098 $/g	$0.001
Na2MoO4.2H2O	1 mL	6.3 mg/ml dH2O	0.200 $/g	$0.001
ZnSO4.7H2O	1 mL	0.002 g/ml dH2O	0.070 $/g	$0.002
CoCl2.6H2O	1 mL	0.01 g/ml dH2O	0.464 $/g	$0.005
MnCl2.4H2O	1 mL	0.18 g/ml dH2O	0.118 $/g	$0.021
			TOTAL	**$1.073 per Liter**
Vitamin Solution	**Amt. per L Water**	**Stock Concentration**	**Purchase Price ($/g)**	**Cost ($)**
Vitamin B12	1 mL	1 mg/ml dH2O	40.200 $/g	$0.0400
Biotin	1 mL	1 mg/ml dH2O	35.960 $/g	$0.0360
Thiamine HCl	200 mg	--	0.3070 $/g	$0.0610
			TOTAL	**$0.1376 per Liter**
Component	**Amt per L Water**	**Stock Concentration**	**Purchase Price ($/g)**	**Cost ($)**
NaNO3	1 mL	75 g/L dH2O	0.05600 $/g	$0.00400
NaH2PO4.H2O	1 mL	5 g/L dH2O	0.12200 $/g	$0.00100
Na2.SiO3.9H2O	1 mL	30 g/L dH2O	0.13000 $/g	$0.00400
F/2 Trace Metal Solution	1 mL		0.00100 $/mL	$0.00100
f/2 Vitamin Solution	0.5 mL		0.00006 $/mL	$0.00007
			TOTAL COST	**$0.00980 per Liter**

Simgae™ Reactor Tubes

In order to determine the effect of changing the length and diameter of the tubes on the various parameters of the cultivation fields, the following analysis was carried out. Four tube diameters were proposed: 6, 8, 10 and 12 inches. The corresponding tube lengths were calculated such that the same initial volume of 245 ft^3 per tube was maintained, as shown in table A2. The reactor bed-width is determined by the number of tubes in the bed and the tube diameters.

Table A2: Analysis of the Dimensions of the Four Proposed Tube Diameters

Tube Diameter (in)	Tube Radius (in)	Tube Length (ft)	Volume Per Tube (ft^3)	% Increase In Diameter	% Decrease In Length	% Length Decrease / % Diameter Increase
6	3	1250	245	0	0	0
8	4	703	245	33	44	1.3
10	5	450	245	67	64	0.96
12	6	313	245	100	75	0.75

The reactor bed width was kept constant at 8 ft. Therefore, in order to determine how many tubes would fit per reactor bed, the reactor bed width was divided by each of the tube diameters as shown below in table A3. Given the width and length of the reactor beds, the reactor area could be calculated. By accounting for the excess lengths and widths for path space, the total bed area was calculated as shown below.

Table A3: Dimensions of the Reactor Bed for the Four Proposed Tube Diameters

Tube Diameter (in)	Reactor Width (ft)	Tubes per Reactor Bed	Excess Width* (ft)	Total Width (ft)	Tube Length (ft)	Excess Length** (ft)	Total Length (ft)	Net Reactor Area (acres)	Total Bed Area (acres)
6	8	16	3	11	1250	70	1320	0.230	0.33
8	8	12	3	11	703	70	773	0.129	0.20
10	8	9.6	3	11	450	70	520	0.083	0.13
12	8	8	3	11	313	70	383	0.057	0.10

*Excess width is due to 1.5 ft of path space on both sides of the reactor bed.
**Excess length is due to PM and FM media storage tanks, fermentation tanks and clarifiers.

The ratio of the desired number of tubes to the number of tubes that can fit in a field (*ratio T*) can be found in table A4. To find the actual field area required to maintain the same total volume per field at the different tube diameters, the original field area is multiplied by the *ratio T*. The capital costs for installation including estimates for the cost of land, harvesting, and product storage for the Simgae™ cultivation system is within the range of \$45k to \$60k per acre. To remain conservative, the upper limit of this range is used for estimating the capital cost per field. As shown below, since the area per field decreases as the tube diameter increases, the capital cost per field also decreases. The costs per field for each possible tube diameter are shown below in table A4, and as can be seen, and increase in the diameter from 6 to 12 inches, results in a 42% decrease in the cost per field. Based on the significant reduction in capital costs, and the aforementioned benefits of utilizing tubes of larger diameters, a decision was made to design the fields based on 12-inch diameter tubes. However, it is important to note that while the land area required per field decreases as the diameter increases, the pumping power requirement increases as both the flow velocity and volumetric flow rates increase.

Table A4: A Summary of the Actual Field Areas and the Cost per Field

Tube Diameter (in)	Field area (acres)	*ratio T* (T_{want}/T_{fit})	Actual Net Reactor Area* (acres)	Actual Total Bed Area** (acres)	Actual Field Area*** (acres)	Cost per Acre \$	Cost Per Field \$
6	33.33	1	0.230	0.333	33.3	6.0E+04	2.00E+06
8	19.52	1.3	0.172	0.260	26.0	6.0E+04	1.56E+06
10	13.13	1.7	0.138	0.219	21.9	6.0E+04	1.31E+06
12	9.66	2	0.115	0.193	19.3	6.0E+04	1.16E+06

*Actual Net Reactor Area = Net Reactor Area x *ratio T*
** Actual Total Bed Area = Total Bed Area x *ratio T*
*** Actual Field Area = Actual Total Bed Area x 100

Based on the results discussed above, the proposed cultivation system will be comprised of a series of fields, each with the following parameters: a gross area of 23.2 acres (383 ft. x 2640 ft.), a bed area of 19.3 acres (100 x 383 ft. x 2200 ft.), a net reactor area of 11.5 acres (100 x16 ft. x 313 ft.), and a total volume capacity of 393,000 ft^3. Each field will contain 100 reactor beds, each of which will have 16 tubes, for a total of 1600 tubes per field.

Carbon Dioxide Concentration Profile along the Tube (PM)

Each Simgae^{TM} reactor bed tube can be modeled as a plug flow reactor (PFR), where the change in the concentration of a species *i* can be modeled by solving the following differential equation, subject to the appropriate boundary conditions[99].

$$\frac{\partial c_i}{\partial t} = -\frac{\partial (c_i Q)}{\partial V} + R_j \qquad EQN.A1$$

Where,

c_i = the concentration of species of interest, *i.* (mol.m^{-3})
Q = the volumetric flow rate (m^3.s^{-1})
V = the volume of the reactor (m3)
R_j = productivity (mol.m^{-3}.s^{-1})

However, under steady state conditions, and for a constant cross sectional area of the tube, the equation can be simplified to the following form[100]:

$$\frac{\partial (Qc_i)}{\partial V} = R_j \qquad EQN.A2$$

Where *V* is just the product of the cross sectional area of a single tube, *A*, and the length *z*. Given the initial concentration of CO_2, a specific growth (0.005 hr^{-1}), and the stoichiometric amount of CO_2 required for growth (1.83 g_{CO2}/g_{Algae}), equation A2 can be solved to obtain the CO_2 concentration profile down the tube as a function of position. The solution to the differential equation shown above has the following form:

$$c_i(z) = c_o e^{\frac{-k.Az}{Q}} \qquad EQN.A3$$

Where c_o is the required initial concentration of CO_2 in the reactor tubes.

In order to calculate the amount of heat lost to the surroundings or gained from the surroundings through the walls of the fermentation tank, the following calculation was performed[101]:

$$Q_{conduction} = \frac{k.A_s\Delta T}{t} \qquad EQN.A4$$

Where,
k= thermal conductivity of foam glass ($Wm^{-1}K^{-1}$)
A_s = surface area of a cylindrical fermentation tank (m^2)
ΔT = temperature difference across the wall of the fermentation tank (°K)
t= thickness of the wall of the fermentation tank (m)

The cultivation fields will be located in Texas, where temperatures range from 10°C to 16°C between November and March, and from 16°C to 27°C between the months of April and October.[102] Average temperatures of 13°C and 22°C will be used to determine the amount of heating required during the winter and summer seasons respectively. It is anticipated that a significant amount of heat loss will occur through the walls of the fermentation tanks. It is therefore suggested that the tanks be insulated with foam glass, which not only has a very low thermal conductivity, but also is very cheap and adds negligible cost to the current design. The calculation below indicates the change in the temperature that would occur during the wintertime without auxilliary heating or cooling due to conductive losses through the tank walls alone.

November – March: T_{avg} = 13°C

$$\frac{(0.045\ \mathrm{W.m^{-1}K^{-1}}\times 584\ \mathrm{m^2}\times(13-28)°\mathrm{C})}{0.625\ in\times 0.0254\ \frac{m}{in}\times\frac{hr}{3{,}600\ s}}\times\frac{1\ kJ}{1{,}000\ J} = 1909m^3\times\frac{1{,}000kg}{m^3}\times\frac{4.184\square J}{kg.°C}\times\Delta T\left(\frac{°C}{hr}\right)$$

$\Delta T = -0.011°C.hr^{-1}$

April - October: T_{avg} = 22°C

Similarly, the calculation below indicates the change in the temperature that would occur during the summertime without auxilliary heating or cooling due to conductive losses through the tank walls alone.

$$\frac{(0.045\ \mathrm{W.m^{-1}K^{-1}} \times 584\ \mathrm{m^2} \times (22-28)\mathrm{°C})}{0.625\ in \times 0.0254\ \frac{m}{in} \times \frac{hr}{3{,}600\ s}} \times \frac{1\ kJ}{1{,}000\ J} = 1909 m^3 \times \frac{1{,}000 kg}{m^3} \times \frac{4.184 kJ}{kg.°C} \times \Delta T\left(\frac{°C}{hr}\right)$$

$\Delta T = -0.0045°C.hr^{-1}$

C. Module II Conventional Energy Requirements

Figure A1: Conventional Energy Requirements for Module II

Conventional Energy Requirements

OriginOil

Harvest Concentration:
1 gram/L dry weight

Centrifuge for 1 MGD sludge processing	1,059		kWh
Centrifuge for processing 10,000,000 L (2.64 MG)	2,798		kWh
Sludge solid content		27	%
Sludge moisture content		73	%
Total biomass in 10,000,000 L		**10,000**	kg
Total moisture (water) content	27,037		kg
Energy requirement for water evaporation	16,770		kWh
Total energy requirement for dewatering	19,568		kWh
Cost for dewatering 10,000,000 L of algae culture	1,370		$
Energy cost for oil extraction	1,113		$
Total energy cost of crude oil	2,483		$
Energy cost per kg of crude oil		1.24	$ /kg

A BREAKTHROUGH TECHNOLOGY TO TRANSFORM ALGAE INTO OIL

(OTCBB: OOIL)

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D. Module III Calculations and Simulation Details

1. MATLAB Transesterification Simulation Results

As discussed in Section B (Process Description), a MATLAB Model was used to determine the steady state concentration for each of the six species involved in the transesterification reacton network. The following differential equations were written to model the changing concentrations of each species as a function of time. (Note: Notation is as follows – TG, Triglyceride; DG, Diglyceride; MG, Monoglyceride; GL, Glycerol; MeOH, Methanol; FAME, Fatty acid methyl esters)

(1) $$\frac{d[TG]}{dt} = -k_{1f}[TG][MeOH] + k_{1r}[DG][FAME]$$

(2) $$\frac{d[DG]}{dt} = k_{1f}[TG][MeOH] - k_{1r}[DG][FAME] - k_{2f}[DG][MeOH] + k_{2r}[MG][FAME]$$

(3) $$\frac{d[MG]}{dt} = k_{2f}[DG][MeOH] - k_{2r}[MG][FAME] - k_{3f}[MG][MeOH] + k_{3r}[GL][FAME]$$

(4) $$\frac{d[GL]}{dt} = k_{3f}[MG][MeOH] - k_{3r}[GL][FAME]$$

(5) $$\frac{d[FAME]}{dt} = k_{1f}[TG][MeOH] - k_{1r}[DG][FAME] + k_{2f}[DG][MeOH] - k_{2r}[MG][FAME] + k_{3f}[MG][MeOH] - k_{3r}[GL][FAME]$$

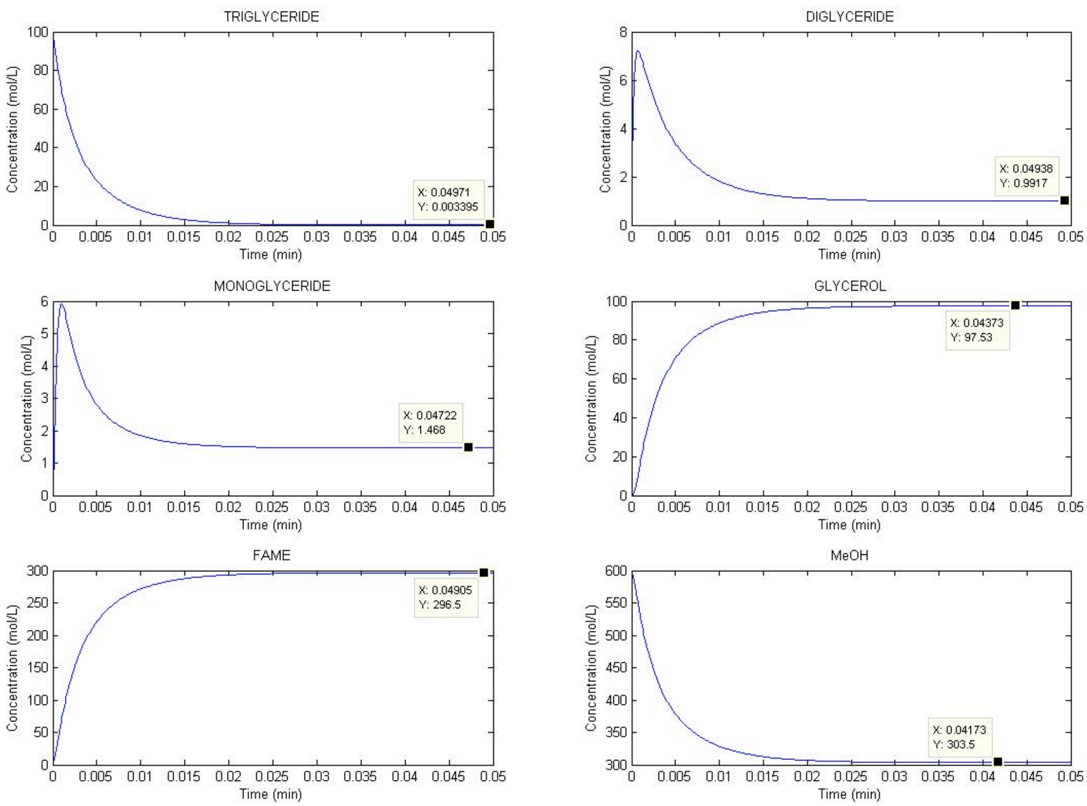
(6) $$\frac{d[MeOH]}{dt} = -\frac{d[FAME]}{dt}$$

These reactions were represented by the following MATLAB code below. A basis of 100 mols/L was used for the Triglyceride starting concentration. Please see Figure A2 below for the MATLAB simulation results.

```
function [dC] = tri(t,C)
k1f = 0.634; k1r = 0.000; k2f = 7.104; k2r = 4.912; k3f = 7.860; k3r = 0.121; % (L/mol-min)
dC = zeros(6,1);
% C(1) = [TG] % C(2) = [DG] % C(3) = [MG] % C(4) = [GL] % C(5) = [FAME] % C(6) = [MeOH]
 dC(1)= -(k1f)*C(1)*C(6)+(k1r)*C(2)*C(5); % dTG
dC(2)= (k1f)*C(1)*C(6)-(k1r)*C(2)*C(5) (k2f)*C(2)*C(6)+(k2r)*C(3)*C(5); % dDG
dC(3)= (k2f)*C(2)*C(6)-(k2r)*C(3)*C(5)-(k3f)*C(3)*C(6)+(k3r)*C(4)*C(5); % dMG dC(4)=
(k3f)*C(3)*C(6)-(k3r)*C(4)*C(5); % dGL
```

```
dC(5)= (k1f)*C(1)*C(6)-(k1r)*C(2)*C(5)+(k2f)*C(2)*C(6)-(k2r)*C(3)*C(5)+(k3f)*C(3)*C(6)-
(k3r)*C(4)*C(5); % dFAME
dC(6)= -dC(5); % dMeOH
end
```

Figure A2: Dynamic Modeling of the Transesterification Reaction Network



From Figure A2, the steady-state concentration of each species in the Transesterificaiton Reaction Network may be determined. Since the initial reactant and product concentrations are already known, these results may be used to compute the steady state conversion for each of the three reactions. The calculated conversions are shown below in table A5 below

Table A5. The Steady-State Transesterification Reaction Conversions

Reaction	[Initial or Intrmd] (mol/L)	[Final] (mol/L)	Fractional Conversion Outlet 1
TG --> DG	100	0.003395	0.99996605
DG --> MG	99.996605	0.9917	0.990082663
MG --> GL	99.004905	1.468	0.985172452
Overall			**0.97536905**

2. Tubular Reactor Design: Temperature and Concentration Profiles

As discussed in Module III, Section D, "Equipment List and Unit Descriptions," the first reactor is tubular and is feed with a series of cold shots. The scheme for injecting the reactor with portions of chilled feed is detailed below. The resulting temperature profile and concentration profile in the reactor are shown in Figures A3 and A4. It is important to note that the saw-tooth temperature profile gives an average reactor temperature of 140ºF, shown as a red dashed line.

First Shot:

$T_{Initial}$: 77ºF; T_{Final}: 186.9ºF;
Portion of Total Feed: 1/6;
Segment Distance: 32.4 ft

Second Shot:

$T_{Initial}$: 10ºF; T_{Final}: 174.6ºF;
Portion of Total Feed: 1/4;
Segment Distance: 98.6 ft

Third Shot:

$T_{Initial}$: 10ºF; T_{Final}: 175.9ºF;
Portion of Total Feed: 1/4;
Segment Distance: 98.6 ft

Fourth Shot:

$T_{Initial}$: 10ºF; T_{Final}: 180.0ºF;
Portion of Total Feed: 1/3;
Segment Distance: 350 ft

Overall
Temp Initial: 77ºF
Temp Final: 180.0ºF
Total Distance: 580 ft

Figure A3: Reactor R-101 Temperature Profile

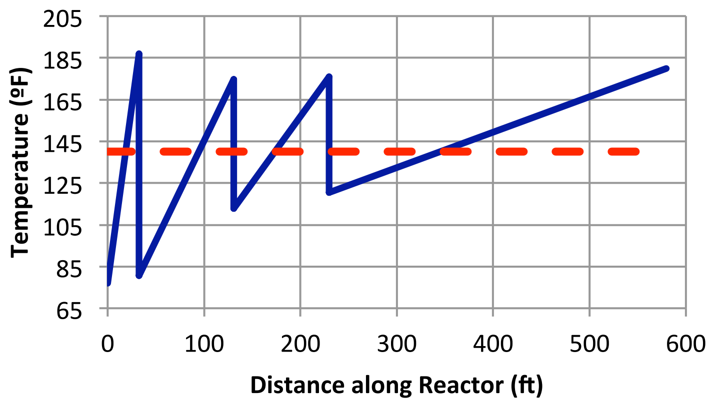
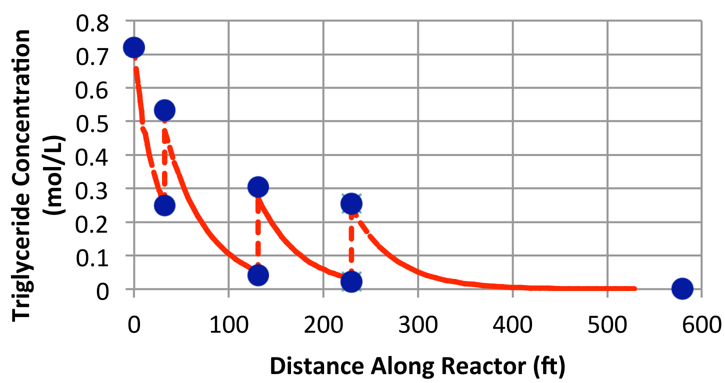
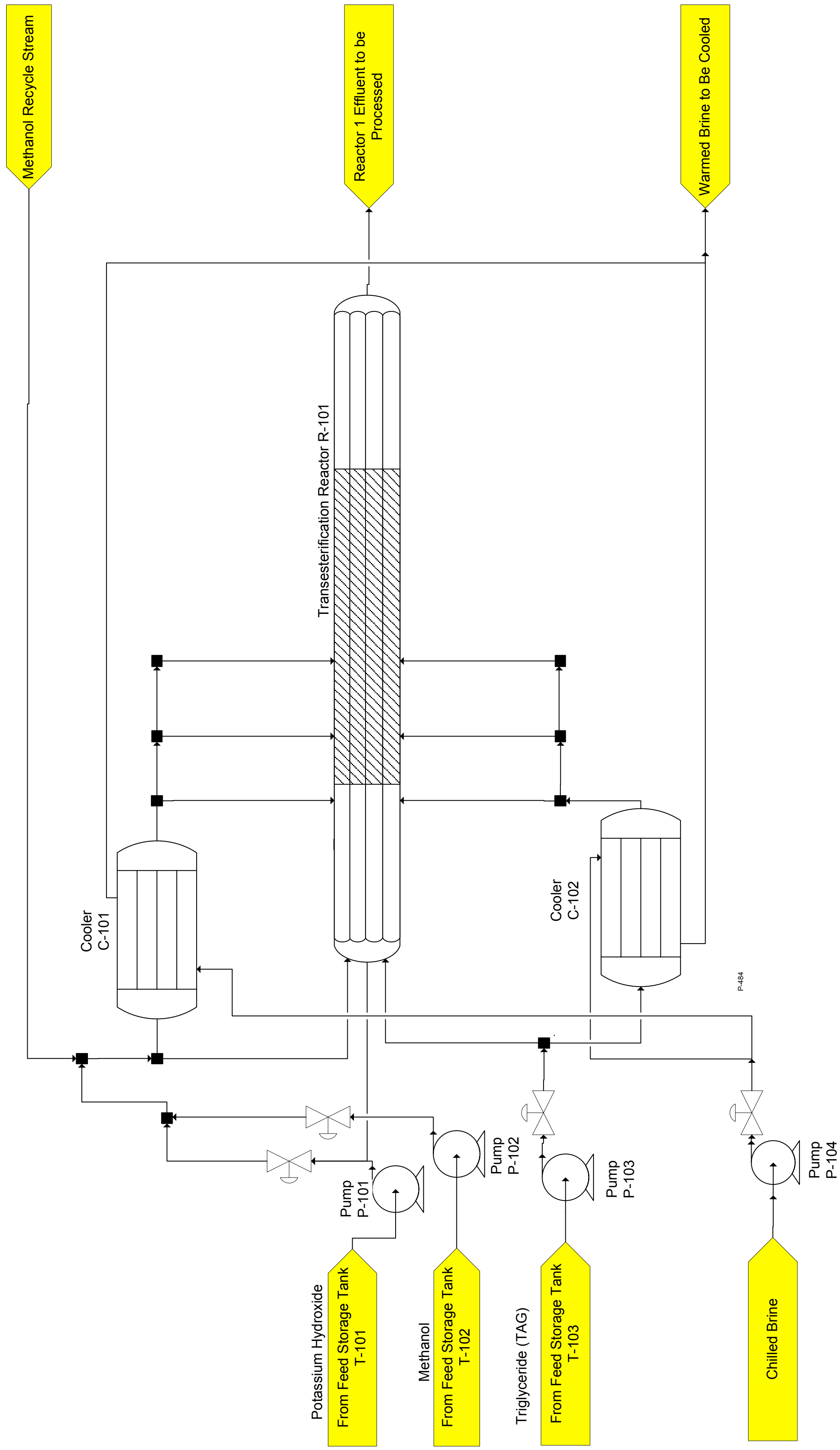


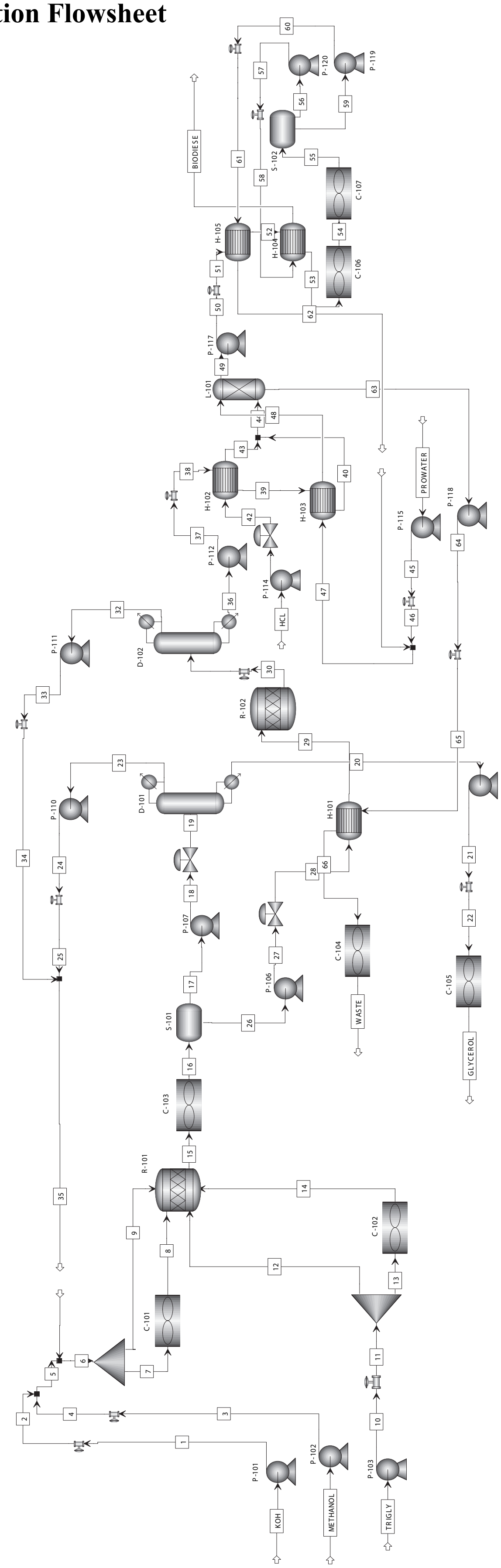
Figure A4: Reactor R-101 Concentration Profile



3. Tubular Reactor Design: Visio Flow Sheet



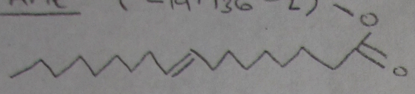
4. Aspen PLUS Simulation Flowsheet



5. Aspen PLUS Group Contribution Property Estimation Method

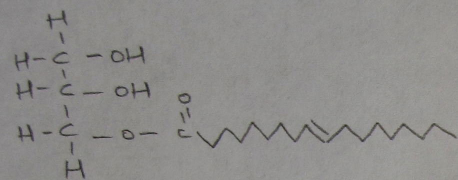
9- OCTADECANOIC ACID METHYL ESTER (65.75%)

FAME ($C_{19}H_{36}O_2$)



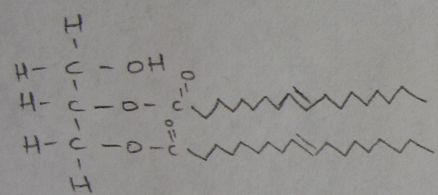
UNIFAC	JOBACK
CH_3 (1015): 2	CH_3 (100): 2
HC=CH (1065): 1	=CH- (105): 2
COO (3300): 1	COO (127): 1
CH_2 (1010): 14	CH_2 (101): 14

MONOGLYCERIDE ($C_{21}O_4H_{40}$)



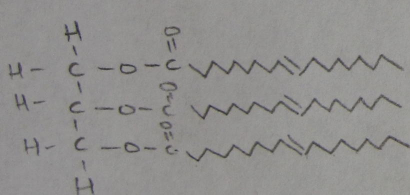
UNIFAC	JOBACK
CH_3 (1015): 1	CH_3 (100): 1
CH (1005): 1	CH (102): 1
CH_2COO (1500): 1	COO (127): 1
CH_2 (1010): 15	CH_2 (101): 16
OH (1200): 2	OH (119): 2
HC=CH (1065): 1	=CH- (105): 2

DIGLYCERIDE ($C_{39}O_5H_{72}$)



UNIFAC	JOBACK
CH_3 (1015): 2	CH_3 (100): 2
CH (1005): 1	CH (102): 1
CH_2COO (1500): 1	COO (127): 2
COO (3300): 1	CH_2 (101): 30
CH_2 (1010): 29	OH (119): 1
	=CH- (105): 4

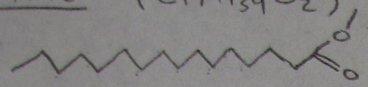
TRIGLYCERIDE ($C_{57}O_6H_{104}$)



UNIFAC	JOBACK
CH_3 (1015): 3	CH_3 (100): 3
CH (1005): 1	CH (102): 1
CH_2COO (1500): 2	COO (127): 3
COO (3300): 1	CH_2 (101): 44
CH_2 (1010): 42	=CH- (105): 6
HC=CH (1065): 3	

HEXADECENOIC ACID METHYL ESTER (10.1%)

FAME ($C_{17}H_{34}O_2$)



ORRICK-E

c(100): 16

coo(114): 1

UNIFAC

CH_3 (1015): 2

CH_2 (1010): 14

coo (3300): 1

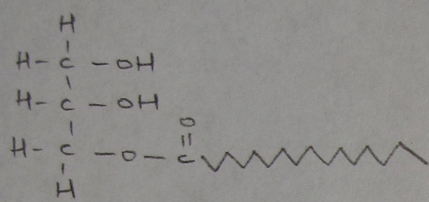
JOBACK

CH_3 (100): 2

CH_2 (101): 14

coo (127): 1

MONOGLYCERIDE ($C_{19}H_{38}O_4$)



UNIFAC

CH_3 (1015): 1

CH_2COO (1500): 1

CH_2 (1010): 15

CH (1005): 1

OH (1200): 2

JOBACK

CH_3 (100): 1

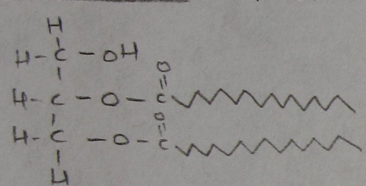
CH_2 (101): 16

coo (127): 1

CH (102): 1

OH (119): 2

DIGLYCERIDE ($C_{35}H_{68}O_5$)



UNIFAC

CH (1005): 1

OH (1200): 1

CH_3 (1015): 2

CH_2 (1010): 29

COO (3300): 1

CH_2COO (1500): 1

JOBACK

CH_3 (100): 2

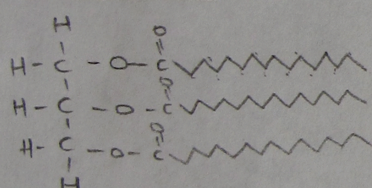
CH_2 (101): 30

COO (127): 2

CH (102): 1

OH (119): 1

TRIGLYCERIDE ($C_{51}H_{98}O_6$)



UNIFAC

CH (1005): 1

CH_3 (1015): 3

CH_2 (1010): 42

COO (3300): 1

CH_2COO (1500): 2

ORRICK-E

c (100): 48

coo (114): 3

JOBACK

CH_3 (100): 3

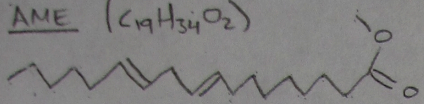
CH_2 (101): 44

coo (127): 3

CH (102): 1

9,12 OCTADECADIENOIC ACID METHYL ESTER (18.33%)

AME ($C_{19}H_{34}O_2$)

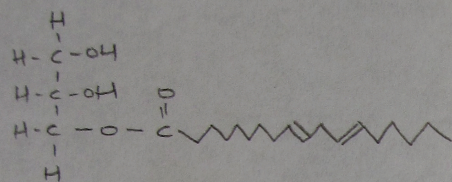


ORRICK-E
100 (14)(C)
114 (COO): 1
104 (C=C): 4

UNIFAC
CH_3 (1015): 2
HC=CH (1065): 2
COO (3300): 1
CH_2 (1010): 12

JOBACK
CH_3 (100): 2
-CH= (105): 4
COO (127): 1
CH_2 (101): 12

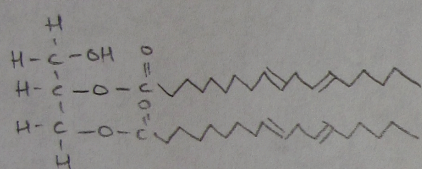
ONOGLYCERIDE ($C_{21}O_4H_{38}$)



UNIFAC
CH_3 (1015): 1
CH (1005): 1
CH_2COO (1500): 1
CH_2 (1010): 13
OH (1200): 2
HC=CH (1065): 2

JOBACK
CH_3 (100): 1
CH_2 (102): 1
COO (127): 1
CH_2 (101): 14
OH (119): 2
-CH= (105): 4

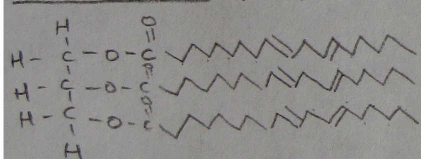
IGLYCERIDE ($C_{39}O_5H_{68}$)



UNIFAC
CH_3 (1015): 2
CH (1005): 1
CH_2COO (1500): 1
COO (3300): 1
CH_2 (1010): 25
OH (1200): 1
HC=CH (1065): 4

JOBACK
CH_3 (100): 2
CH (102): 1
COO (127): 2
CH_2 (101): 26
OH (119): 1
=CH- (105): 8

RIGLYCERIDE ($C_{57}O_6H_{98}$)



UNIFAC
CH_3 (1015): 3
CH (1005): 1
CH_2COO (1500): 2
COO (3300): 1
CH_2 (1010): 34
HC=CH (1065): 6

JOBACK
CH_3 (100): 3
CH (102): 1
COO (127): 3
CH_2 (101): 36
=CH- (105): 12

ORRICK-E
C (100): 42
C=C (104): 12
COO (114): 3

6. Aspen PLUS Simulation Results

```
 BLOCK:  C-101    MODEL: HEATER
 ------------------------------
   INLET STREAM:          7
   OUTLET STREAM:         8
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN               OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           1359.12         1359.12
-0.167294E-15
       MASS(LB/HR   )           42632.5         42632.5
0.170667E-15
       ENTHALPY(BTU/HR  )      -0.140960E+09   -0.143182E+09
0.155207E-01

                          ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE               F                       10.00000
   PRESSURE DROP                      PSI                       7.00000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.000100000



                           ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                  10.000
   OUTLET PRESSURE       PSIA                               57.700
   HEAT DUTY             BTU/HR                           -0.22223E+07



   PRESSURE-DROP CORRELATION PARAMETER                     0.13791E+07



 BLOCK:  C-102    MODEL: HEATER
 ------------------------------
   INLET STREAM:          13
   OUTLET STREAM:         14
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
```

```
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          212.893         212.893         0.00000
       MASS(LB/HR   )          186324.         186324.
0.156200E-15
       ENTHALPY(BTU/HR  )      -0.162026E+09   -0.166434E+09
0.264886E-01

                        ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE               F                       10.00000
   PRESSURE DROP                       PSI                      7.00000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.000100000



                         ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                  10.000
   OUTLET PRESSURE       PSIA                               57.700
   HEAT DUTY             BTU/HR                           -0.44086E+07



   PRESSURE-DROP CORRELATION PARAMETER                      4641.7



  BLOCK:  C-103    MODEL: HEATER
  ------------------------------
   INLET STREAM:          15
   OUTLET STREAM:         16
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
```

```
      MOLE(LBMOL/HR)           1887.17         1887.17         0.00000
      MASS(LB/HR   )           274857.         274857.
0.211774E-15
      ENTHALPY(BTU/HR  )      -0.377740E+09   -0.386883E+09
0.236329E-01

                          ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE              F                      100.000
   PRESSURE DROP                      PSI                      7.00000
   MAXIMUM NO. ITERATIONS                                     30
   CONVERGENCE TOLERANCE
0.000100000



                           ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                 100.00
   OUTLET PRESSURE       PSIA                              45.700
   HEAT DUTY             BTU/HR                           -0.91432E+07



   PRESSURE-DROP CORRELATION PARAMETER                     30761.



 BLOCK:  C-104    MODEL: HEATER
 ------------------------------
   INLET STREAM:          66
   OUTLET STREAM:         WASTE
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                      ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
      MOLE(LBMOL/HR)           11521.9         11521.9
0.157872E-15
      MASS(LB/HR   )           208830.         208830.         0.00000
      ENTHALPY(BTU/HR  )      -0.140153E+10   -0.141178E+10
0.726064E-02

                          ***  INPUT DATA  ***
```

```
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE               F                       100.000
   PRESSURE DROP                       PSI                       7.00000
   MAXIMUM NO. ITERATIONS                                       30
   CONVERGENCE TOLERANCE
0.000100000


                          ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                    100.00
   OUTLET PRESSURE       PSIA                                 15.996
   HEAT DUTY             BTU/HR                             -0.10250E+08



   PRESSURE-DROP CORRELATION PARAMETER                        69080.



 BLOCK:  C-105    MODEL: HEATER
 ------------------------------
   INLET STREAM:          22
   OUTLET STREAM:         GLYCEROL
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          300.820        300.820        0.00000
       MASS(LB/HR   )          23912.2        23912.2        0.00000
       ENTHALPY(BTU/HR  )     -0.748642E+08   -0.783722E+08
0.447603E-01

                         ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE               F                       100.000
   PRESSURE DROP                       PSI                       7.00000
   MAXIMUM NO. ITERATIONS                                       30
   CONVERGENCE TOLERANCE
0.000100000
```

```
                     ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                   100.00
   OUTLET PRESSURE       PSIA                                16.880
   HEAT DUTY             BTU/HR                            -0.35080E+07


   PRESSURE-DROP CORRELATION PARAMETER                      0.61658E+07


 BLOCK:  C-106    MODEL: HEATER
 ------------------------------
   INLET STREAM:          53
   OUTLET STREAM:         54
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                  ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          6508.89         6508.89
-0.413185E-12
       MASS(LB/HR   )          328115.         328115.
-0.147419E-12
       ENTHALPY(BTU/HR  )     -0.923359E+09   -0.929043E+09
0.611738E-02

                      ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE              F                       100.000
   PRESSURE DROP                      PSI                       7.00000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.000100000


                     ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                   100.00
   OUTLET PRESSURE       PSIA                                24.696
   HEAT DUTY             BTU/HR                            -0.56833E+07
```

```
  PRESSURE-DROP CORRELATION PARAMETER                        25394.


 BLOCK:  C-107    MODEL: HEATER
 ------------------------------
   INLET STREAM:          54
   OUTLET STREAM:         55
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           6508.89         6508.89         0.00000
       MASS(LB/HR   )           328115.         328115.
-0.354800E-15
       ENTHALPY(BTU/HR  )      -0.929043E+09   -0.936292E+09
0.774213E-02

                        ***  INPUT DATA  ***
   TWO    PHASE   TP  FLASH
   SPECIFIED TEMPERATURE                F                         54.5000
   PRESSURE DROP                        PSI                        7.00000
   MAXIMUM NO. ITERATIONS                                         30
   CONVERGENCE TOLERANCE
0.000100000



                         ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                     54.500
   OUTLET PRESSURE       PSIA                                  17.696
   HEAT DUTY             BTU/HR                               -0.72489E+07
   OUTLET VAPOR FRACTION                                       0.0000
   PRESSURE-DROP CORRELATION PARAMETER                         25736.


   V-L PHASE EQUILIBRIUM :

      COMP                F(I)           X(I)           Y(I)           K(I)
```

```
   MOLE(LBMOL/HR)         810.330       810.330      0.269227E-15
0.140297E-15
   MASS(LB/HR   )         40005.9       40005.9
-0.181872E-15
   ENTHALPY(BTU/HR  )   -0.130954E+09 -0.127013E+09
-0.300940E-01
```

```
                         **********************
                         ****  INPUT DATA  ****
                         **********************

   ****   INPUT PARAMETERS   ****

    NUMBER OF STAGES                                              10
    ALGORITHM OPTION                                            STANDARD
    INITIALIZATION OPTION                                       STANDARD
    HYDRAULIC PARAMETER CALCULATIONS                            NO
    INSIDE LOOP CONVERGENCE METHOD                              NEWTON
    DESIGN SPECIFICATION METHOD                                 NESTED
    MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS                       100
    MAXIMUM NO. OF INSIDE LOOP ITERATIONS                         10
    MAXIMUM NUMBER OF FLASH ITERATIONS                            30
    FLASH TOLERANCE                                                0.000100000
    OUTSIDE LOOP CONVERGENCE TOLERANCE                             0.000100000

   ****   COL-SPECS   ****

    MOLAR VAPOR DIST / TOTAL DIST                                  0.0
    MOLAR REFLUX RATIO                                             0.100000
    MASS DISTILLATE RATE           LB/HR                 16,093.7

   **** REAC-STAGES SPECIFICATIONS ****

    STAGE  TO  STAGE            REACTIONS/CHEMISTRY ID
      1        10                      GLOBAL


        *****  CHEMISTRY PARAGRAPH GLOBAL  *****


          ****  REACTION PARAMETERS  ****

   RXN NO. TYPE         PHASE      CONC.    TEMP APP TO EQUIL    CONVERSION
                                   BASIS            F
```

1	EQUILIBRIUM	LIQUID	MOLE-GAMMA	0.0000
2	EQUILIBRIUM	LIQUID	MOLE-GAMMA	0.0000

** STOICHIOMETRIC COEFFICIENTS **

RXN NO.	TRIGLY-1	GLYCEROL	METHANOL	TRIGLY-2	TRIGLY-3
1	0.000	0.000	0.000	0.000	0.000
2	0.000	0.000	0.000	0.000	0.000

RXN NO.	OH-	WATER	K+	H3O+	DIGLY-1
1	1.000	-2.000	0.000	1.000	0.000
2	0.000	-1.000	0.000	1.000	0.000

RXN NO.	DIGLY-2	DIGLY-3
1	0.000	0.000
2	0.000	0.000

** COEFFICIENTS OF EQUILIBRIUM CONSTANT EXPRESSION **

RXN NO.	A	B	C	D
1	132.90	-13446.	-22.477	0.0000
2	9.2058	0.0000	0.0000	0.0000

**** PROFILES ****

P-SPEC STAGE 1 PRES, PSIA 8.00000

****
**** RESULTS ****
****

*** COMPONENT SPLIT FRACTIONS ***

	OUTLET STREAMS	
	23	20
COMPONENT:		
TRIGLY-1	1.0000	0.0000
GLYCEROL	0.0000	1.0000
METHANOL	.99686	.31380E-02
TRIGLY-2	1.0000	.33149E-11
TRIGLY-3	1.0000	.24875E-12
OH-	.27468E-07	1.0000
WATER	.25086	.74914
K+	0.0000	1.0000

H3O+	.99992	.79844E-04
DIGLY-1	0.0000	1.0000
DIGLY-2	0.0000	1.0000
DIGLY-3	0.0000	1.0000

*** SUMMARY OF KEY RESULTS ***

TOP STAGE TEMPERATURE	F	123.200
BOTTOM STAGE TEMPERATURE	F	334.718
TOP STAGE LIQUID FLOW	LBMOL/HR	50.9510
BOTTOM STAGE LIQUID FLOW	LBMOL/HR	300.820
TOP STAGE VAPOR FLOW	LBMOL/HR	0.0
BOILUP VAPOR FLOW	LBMOL/HR	576.272
MOLAR REFLUX RATIO		0.100000
MOLAR BOILUP RATIO		1.91567
CONDENSER DUTY (W/O SUBCOOL)	BTU/HR	-8,832,380.
REBOILER DUTY	BTU/HR	0.128879+08

**** MAXIMUM FINAL RELATIVE ERRORS ****

BUBBLE POINT	0.12856E-04	STAGE= 9	
COMPONENT MASS BALANCE	0.31466E-08	STAGE= 7	COMP=WATER
ENERGY BALANCE	0.52718E-02	STAGE= 5	

**** PROFILES ****

NOTE REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS
FROM THE STAGE INCLUDING ANY SIDE PRODUCT.

STAGE	TEMPERATURE F	PRESSURE PSIA	ENTHALPY BTU/LBMOL LIQUID	ENTHALPY BTU/LBMOL VAPOR	HEAT DUTY BTU/HR
1	123.20	8.0000	-0.10234E+06	-86281.	-.88324+07
2	129.56	9.0000	-0.10329E+06	-86580.	
3	130.43	9.1100	-0.10353E+06	-86659.	
4	131.02	9.2200	-0.10359E+06	-86673.	
5	146.53	9.3300	-0.15209E+06	-86495.	
6	148.48	9.4400	-0.15294E+06	-86771.	
7	153.60	9.5500	-0.15541E+06	-87663.	
9	191.44	9.7700	-0.16653E+06	-95742.	
10	334.72	9.8800	-0.24889E+06	-0.10118E+06	.12888+08

STAGE FLOW RATE FEED RATE PRODUCT RATE

	LBMOL/HR		LBMOL/HR			LBMOL/HR	
	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	LIQUID	VAPOR
1	560.5	0.000				509.5097	
2	50.71	560.5					
3	50.45	560.2					
4	44.45	560.0					
5	943.8	554.0	810.3301				
6	940.4	643.0					
7	930.3	639.6					
9	877.1	615.2					
10	300.8	576.3				300.8204	

**** MASS FLOW PROFILES ****

STAGE	FLOW RATE		FEED RATE			PRODUCT RATE	
	LB/HR		LB/HR			LB/HR	
	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	LIQUID	VAPOR
1	0.1770E+05	0.000				.16094+05	
2	1562.	0.1770E+05					
3	1545.	0.1766E+05					
4	1360.	0.1764E+05					
5	0.4402E+05	0.1745E+05	.40006+05				
6	0.4343E+05	0.2011E+05					
7	0.4175E+05	0.1952E+05					
9	0.3580E+05	0.1466E+05					
10	0.2391E+05	0.1188E+05				.23912+05	

**** MOLE-X-PROFILE ****

STAGE	TRIGLY-1	GLYCEROL	METHANOL	TRIGLY-2	TRIGLY-3
1	0.38966E-09	0.22342E-17	0.96753	0.16699E-10	0.60065E-11
2	0.19609E-11	0.33707E-12	0.91222	0.26639E-11	0.70233E-13
3	0.17732E-11	0.45183E-08	0.89907	0.24265E-11	0.63737E-13
4	0.17680E-11	0.59455E-04	0.89607	0.24142E-11	0.63860E-13
5	0.12169E-11	0.26409	0.64544	0.97342E-12	0.39956E-13
6	0.59369E-14	0.26505	0.60791	0.84791E-13	0.41214E-15
7	0.27234E-16	0.26794	0.49977	0.62504E-14	0.38602E-17
9	0.32667E-21	0.29149	0.85304E-01	0.71380E-17	0.10449E-21

```
     10    0.35327E-26   0.82858       0.51586E-02   0.93756E-22
0.25306E-23

                         ****   MOLE-X-PROFILE     ****
   STAGE     OH-           WATER         K+            H3O+
DIGLY-1
      1    0.16524E-10   0.32469E-01   0.10000E-29   0.16524E-10
0.11013E-22
      2    0.78280E-10   0.87778E-01   0.10000E-29   0.78280E-10
0.11389E-17
      3    0.10172E-09   0.10093       0.10000E-29   0.10172E-09
0.10571E-13
      4    0.10822E-09   0.10387       0.10000E-29   0.10822E-09
0.96929E-10
      5    0.32476E-03   0.89817E-01   0.32476E-03   0.51137E-17
0.47840E-06
      6    0.32594E-03   0.12639       0.32594E-03   0.21975E-16
0.48014E-06
      7    0.32949E-03   0.23163       0.32949E-03   0.64833E-15
0.48538E-06
      9    0.34947E-03   0.62250       0.34947E-03   0.72438E-11
0.51539E-06
     10    0.10189E-02   0.16422       0.10189E-02   0.22349E-14
0.15010E-05

                         ****   MOLE-X-PROFILE     ****
   STAGE     DIGLY-2       DIGLY-3
      1    0.10000E-29   0.10000E-29
      2    0.31406E-20   0.13798E-22
      3    0.17177E-15   0.23201E-20
      4    0.93846E-11   0.22101E-15
      5    0.25987E-06   0.10360E-10
      6    0.26081E-06   0.10397E-10
      7    0.26366E-06   0.10511E-10
      9    0.27974E-06   0.11151E-10
     10    0.81535E-06   0.32504E-10

                         ****   MOLE-Y-PROFILE     ****
   STAGE     TRIGLY-1      GLYCEROL      METHANOL      TRIGLY-2
TRIGLY-3
      1    0.75152E-07   0.12805E-22   0.98827       0.98938E-10
0.51928E-09
      2    0.38966E-09   0.22342E-17   0.96753       0.16699E-10
0.60065E-11
      3    0.35457E-09   0.30512E-13   0.96253       0.15429E-10
0.54692E-11
```

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      4    0.35471E-09   0.40711E-09   0.96136        0.15413E-10
0.54711E-11
      5    0.35854E-09   0.47704E-05   0.96180        0.15553E-10
0.55297E-11
      6    0.17862E-11   0.50659E-05   0.94499        0.14288E-11
0.58649E-13
      7    0.87293E-14   0.60375E-05   0.89140        0.12467E-12
0.60599E-15
      9    0.15988E-18   0.28311E-04   0.41473        0.45007E-15
0.41078E-19
     10    0.49719E-21   0.11131E-01   0.12714        0.10306E-16
0.46750E-18

                         ****   MOLE-Y-PROFILE     ****
   STAGE     OH-           WATER         K+            H3O+
DIGLY-1
      1    0.44922-119   0.11734E-01   0.27185-138   0.44922-119
0.98665E-28
      2    0.21316-118   0.32469E-01   0.27230-138   0.21316-118
0.10999E-22
      3    0.27704-118   0.37475E-01   0.27234-138   0.27704-118
0.10309E-18
      4    0.29478-118   0.38637E-01   0.27240-138   0.29478-118
0.95243E-15
      5    0.88363-112   0.38198E-01   0.88363-112   0.13914-125
0.77771E-11
      6    0.88685-112   0.55006E-01   0.88685-112   0.59791-125
0.80828E-11
      7    0.89622-112   0.10859       0.89622-112   0.17635-123
0.90601E-11
      9    0.94719-112   0.58524       0.94719-112   0.19633-119
0.23424E-10
     10    0.27427-111   0.86173       0.27427-111   0.60157-123
0.90884E-09

                         ****   MOLE-Y-PROFILE     ****
   STAGE     DIGLY-2       DIGLY-3
      1    0.14612E-35   0.82553E-36
      2    0.50634E-26   0.12660E-28
      3    0.28056E-21   0.21596E-26
      4    0.15476E-16   0.20780E-21
      5    0.75297E-12   0.17733E-16
      6    0.78822E-12   0.18634E-16
      7    0.90054E-12   0.21533E-16
      9    0.26577E-11   0.69031E-16
     10    0.15172E-09   0.49417E-14
```

**** K-VALUES ****

STAGE	TRIGLY-1	GLYCEROL	METHANOL	TRIGLY-2	TRIGLY-3
1	192.87	0.57313E-05	1.0214	5.9248	86.452
2	198.71	0.66282E-05	1.0606	6.2685	85.522
3	199.96	0.67529E-05	1.0706	6.3583	85.809
4	200.63	0.68473E-05	1.0729	6.3844	85.673
5	294.63	0.18064E-04	1.4901	15.977	138.39
6	300.84	0.19113E-04	1.5545	16.851	142.30
7	320.52	0.22533E-04	1.7836	19.946	156.98
9	489.41	0.97127E-04	4.8617	63.053	393.11
10	0.14076E+06	0.13434E-01	24.646	0.10994E+06	0.18476E+06

**** K-VALUES ****

STAGE	OH-	WATER	K+	H3O+	DIGLY-1
1	0.0000	0.36140	0.0000	0.0000	0.89592E-05
2	0.0000	0.36990	0.0000	0.0000	0.96577E-05
3	0.0000	0.37129	0.0000	0.0000	0.97528E-05
4	0.0000	0.37198	0.0000	0.0000	0.98261E-05
5	0.0000	0.42528	0.0000	0.0000	0.16256E-04
6	0.0000	0.43521	0.0000	0.0000	0.16834E-04
7	0.0000	0.46880	0.0000	0.0000	0.18666E-04
9	0.0000	0.94013	0.0000	0.0000	0.45449E-04
10	0.0000	5.2472	0.0000	0.0000	0.60550E-03

**** K-VALUES ****

STAGE	DIGLY-2	DIGLY-3
1	0.14612E-05	0.82553E-06
2	0.16122E-05	0.91756E-06
3	0.16334E-05	0.93079E-06
4	0.16491E-05	0.94023E-06
5	0.28975E-05	0.17117E-05
6	0.30221E-05	0.17922E-05
7	0.34156E-05	0.20487E-05

9	0.95004E-05	0.61905E-05
10	0.18608E-03	0.15204E-03

**** RATES OF GENERATION ****

LBMOL/HR

STAGE	TRIGLY-1	GLYCEROL	METHANOL	TRIGLY-2	TRIGLY-3	OH-
1	0.000	0.000	0.000	0.000	0.000	0.9261E-08
2	0.000	0.000	0.000	0.000	0.000	0.3127E-08
3	0.000	0.000	0.000	0.000	0.000	0.1163E-08
4	0.000	0.000	0.000	0.000	0.000	-.3225E-09
5	0.000	0.000	0.000	0.000	0.000	-.4810E-08
6	0.000	0.000	0.000	0.000	0.000	0.1584E-13
7	0.000	0.000	0.000	0.000	0.000	0.5832E-12
9	0.000	0.000	0.000	0.000	0.000	0.6199E-08
10	0.000	0.000	0.000	0.000	0.000	-.6353E-08

**** RATES OF GENERATION ****

LBMOL/HR

STAGE	WATER	K+	H3O+	DIGLY-1	DIGLY-2	DIGLY-3
1	-.1852E-07	0.000	0.9261E-08	0.000	0.000	0.000
2	-.6255E-08	0.000	0.3127E-08	0.000	0.000	0.000
3	-.2326E-08	0.000	0.1163E-08	0.000	0.000	0.000
4	0.6449E-09	0.000	-.3225E-09	0.000	0.000	0.000
5	0.9620E-08	0.000	-.4810E-08	0.000	0.000	0.000
6	-.3169E-13	0.000	0.1584E-13	0.000	0.000	0.000
7	-.1166E-11	0.000	0.5832E-12	0.000	0.000	0.000
9	-.1240E-07	0.000	0.6199E-08	0.000	0.000	0.000

10	0.1271E-07	0.000	-.6353E-08	0.000	0.000	0.000

**** MASS-X-PROFILE ****

STAGE	TRIGLY-1	GLYCEROL	METHANOL	TRIGLY-2	TRIGLY-3
1	0.10923E-07	0.65140E-17	0.98148	0.46491E-09	0.15352E-09
2	0.56353E-10	0.10075E-11	0.94868	0.76034E-10	0.18403E-11
3	0.51266E-10	0.13587E-07	0.94063	0.69674E-10	0.16802E-11
4	0.51179E-10	0.17900E-03	0.93865	0.69405E-10	0.16855E-11
5	0.23102E-10	0.52148	0.44343	0.18354E-10	0.69165E-12
6	0.11382E-12	0.52853	0.42176	0.16145E-11	0.72046E-14
7	0.53728E-15	0.54980	0.35680	0.12247E-12	0.69439E-16
9	0.70871E-20	0.65776	0.66972E-01	0.15380E-15	0.20670E-20
10	0.39351E-25	0.95996	0.20794E-02	0.10372E-20	0.25702E-22

**** MASS-X-PROFILE ****

STAGE	OH-	WATER	K+	H3O+	DIGLY-1
1	0.88976E-11	0.18518E-01	0.12378E-29	0.99516E-11	0.21651E-21
2	0.43211E-10	0.51324E-01	0.12690E-29	0.48330E-10	0.22954E-16
3	0.56491E-10	0.59371E-01	0.12766E-29	0.63183E-10	0.21434E-12
4	0.60171E-10	0.61174E-01	0.12782E-29	0.67299E-10	0.19678E-08
5	0.11843E-03	0.34693E-01	0.27224E-03	0.20857E-17	0.63698E-05
6	0.12003E-03	0.49301E-01	0.27593E-03	0.90511E-17	0.64560E-05
7	0.12486E-03	0.92976E-01	0.28703E-03	0.27479E-15	0.67158E-05
9	0.14563E-03	0.27478	0.33478E-03	0.33763E-11	0.78420E-05
10	0.21801E-03	0.37219E-01	0.50117E-03	0.53482E-15	0.11726E-04

**** MASS-X-PROFILE ****

STAGE	DIGLY-2	DIGLY-3
1	0.19532E-28	0.18011E-28
2	0.62888E-19	0.25478E-21
3	0.34602E-14	0.43099E-19
4	0.18928E-09	0.41106E-14
5	0.34377E-05	0.12637E-09
6	0.34842E-05	0.12808E-09
7	0.36244E-05	0.13323E-09
9	0.42288E-05	0.15544E-09
10	0.63284E-05	0.23263E-09

**** MASS-Y-PROFILE ****

STAGE	TRIGLY-1	GLYCEROL	METHANOL	TRIGLY-2	TRIGLY-3
1	0.20875E-05	0.36993E-22	0.99337	0.27294E-08	0.13151E-07
2	0.10923E-07	0.65140E-17	0.98148	0.46491E-09	0.15352E-09
3	0.99615E-08	0.89158E-13	0.97858	0.43050E-09	0.14010E-09
4	0.99707E-08	0.11902E-08	0.97790	0.43029E-09	0.14022E-09
5	0.10076E-07	0.13944E-04	0.97814	0.43410E-09	0.14169E-09
6	0.50576E-10	0.14919E-04	0.96830	0.40181E-10	0.15142E-11
7	0.25326E-12	0.18219E-04	0.93588	0.35924E-11	0.16030E-13
9	0.59395E-17	0.10939E-03	0.55754	0.16606E-13	0.13914E-17
10	0.21346E-19	0.49706E-01	0.19754	0.43947E-15	0.18301E-16

**** MASS-Y-PROFILE ****

STAGE	OH-	WATER	K+	H3O+	DIGLY-1
1	0.23968-119	0.66314E-02	0.33342-138	0.26807-119	0.19221E-26
2	0.11477-118	0.18518E-01	0.33705-138	0.12837-118	0.21624E-21
3	0.14950-118	0.21421E-01	0.33785-138	0.16721-118	0.20314E-17
4	0.15916-118	0.22097E-01	0.33810-138	0.17801-118	0.18776E-13

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      5    0.47700-112   0.21841E-01   0.10965-111   0.84006-126
0.15329E-09
      6    0.48235-112   0.31689E-01   0.11088-111   0.36372-125
0.16051E-09
      7    0.49945-112   0.64100E-01   0.11481-111   0.10992-123
0.18435E-09
      9    0.67589-112   0.44235       0.15537-111   0.15670-119
0.61030E-09
     10    0.22619-111   0.75276       0.51997-111   0.55488-123
0.27367E-07

                         ****   MASS-Y-PROFILE     ****
   STAGE     DIGLY-2       DIGLY-3
      1    0.28281E-34   0.14733E-34
      2    0.98900E-25   0.22803E-27
      3    0.54923E-20   0.38984E-25
      4    0.30312E-15   0.37531E-20
      5    0.14745E-10   0.32020E-15
      6    0.15551E-10   0.33901E-15
      7    0.18205E-10   0.40140E-15
      9    0.68794E-10   0.16477E-14
     10    0.45388E-08   0.13632E-12

 BLOCK:  D-102    MODEL: RADFRAC
 -------------------------------
    INLETS   - 31       STAGE   5
    OUTLETS  - 32       STAGE   1
               36       STAGE  10
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                      ***  MASS AND ENERGY BALANCE  ***
                             IN           OUT       GENERATION   RELATIVE
DIFF.
   TOTAL BALANCE
   MOLE(LBMOL/HR)         1076.84       1076.84     -0.525308E-18   0.00000
   MASS(LB/HR   )         234851.       234851.
0.371773E-15
   ENTHALPY(BTU/HR  )   -0.252573E+09 -0.225504E+09              -0.107176


                         **********************
                         ****  INPUT DATA  ****
                         **********************
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**** INPUT PARAMETERS ****

NUMBER OF STAGES	10
ALGORITHM OPTION	STANDARD
INITIALIZATION OPTION	STANDARD
HYDRAULIC PARAMETER CALCULATIONS	NO
INSIDE LOOP CONVERGENCE METHOD	NEWTON
DESIGN SPECIFICATION METHOD	NESTED
MAXIMUM NO. OF OUTSIDE LOOP ITERATIONS	100
MAXIMUM NO. OF INSIDE LOOP ITERATIONS	10
MAXIMUM NUMBER OF FLASH ITERATIONS	30
FLASH TOLERANCE	0.000100000
OUTSIDE LOOP CONVERGENCE TOLERANCE	0.000100000

**** COL-SPECS ****

MOLAR VAPOR DIST / TOTAL DIST		0.0
MOLAR REFLUX RATIO		0.100000
MASS DISTILLATE RATE	LB/HR	8,420.00

**** REAC-STAGES SPECIFICATIONS ****

STAGE	TO STAGE	REACTIONS/CHEMISTRY ID
1	10	GLOBAL

***** CHEMISTRY PARAGRAPH GLOBAL *****

**** REACTION PARAMETERS ****

RXN NO.	TYPE	PHASE	CONC. BASIS	TEMP APP TO EQUIL F	CONVERSION
1	EQUILIBRIUM	LIQUID	MOLE-GAMMA	0.0000	
2	EQUILIBRIUM	LIQUID	MOLE-GAMMA	0.0000	

** STOICHIOMETRIC COEFFICIENTS **

RXN NO.	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2
1	0.000	0.000	0.000	0.000	0.000
2	0.000	0.000	0.000	0.000	0.000

RXN NO.	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.000	0.000	0.000	1.000	-2.000
2	0.000	0.000	0.000	0.000	-1.000

RXN NO.	K+	H3O+	DIGLY-1	DIGLY-2	DIGLY-3
1	0.000	1.000	0.000	0.000	0.000
2	0.000	1.000	0.000	0.000	0.000

RXN NO.	MONO-1	MONO-2	MONO-3
1	0.000	0.000	0.000
2	0.000	0.000	0.000

** COEFFICIENTS OF EQUILIBRIUM CONSTANT EXPRESSION **

RXN NO.	A	B	C	D
1	132.90	-13446.	-22.477	0.0000
2	9.2058	0.0000	0.0000	0.0000

**** PROFILES ****

P-SPEC STAGE 1 PRES, PSIA 8.00000

****
**** RESULTS ****
****

*** COMPONENT SPLIT FRACTIONS ***

	OUTLET STREAMS	
	32	36
COMPONENT:		
TRIGLY-1	.21511	.78489
FAME-1	0.0000	1.0000
GLYCEROL	0.0000	1.0000
METHANOL	.98581	.14195E-01
FAME-2	0.0000	1.0000
TRIGLY-2	.13252	.86748
FAME-3	.18646E-13	1.0000
TRIGLY-3	.15093	.84907
OH-	.32584E-14	1.0000
WATER	.15931	.84069
K+	0.0000	1.0000
H3O+	MISSING	MISSING
DIGLY-1	0.0000	1.0000
DIGLY-2	0.0000	1.0000
DIGLY-3	0.0000	1.0000
MONO-1	0.0000	1.0000

MONO-2	0.0000	1.0000
MONO-3	0.0000	1.0000

*** SUMMARY OF KEY RESULTS ***

TOP STAGE TEMPERATURE	F	122.377
BOTTOM STAGE TEMPERATURE	F	428.222
TOP STAGE LIQUID FLOW	LBMOL/HR	26.2939
BOTTOM STAGE LIQUID FLOW	LBMOL/HR	813.903
TOP STAGE VAPOR FLOW	LBMOL/HR	0.0
BOILUP VAPOR FLOW	LBMOL/HR	282.678
MOLAR REFLUX RATIO		0.100000
MOLAR BOILUP RATIO		0.34731
CONDENSER DUTY (W/O SUBCOOL)	BTU/HR	-4,526,420.
REBOILER DUTY	BTU/HR	0.315962+08

**** MAXIMUM FINAL RELATIVE ERRORS ****

BUBBLE POINT	0.26055E-09	STAGE= 6
COMPONENT MASS BALANCE	0.11099E-14	STAGE= 6 COMP=GLYCEROL
ENERGY BALANCE	0.14677E-09	STAGE= 5

**** PROFILES ****

NOTE REPORTED VALUES FOR STAGE LIQUID AND VAPOR RATES ARE THE FLOWS FROM THE STAGE INCLUDING ANY SIDE PRODUCT.

STAGE	TEMPERATURE F	PRESSURE PSIA	ENTHALPY BTU/LBMOL LIQUID	ENTHALPY BTU/LBMOL VAPOR	HEAT DUTY BTU/HR
1	122.38	8.0000	-0.10171E+06	-86093.	-.45264+07
2	127.28	9.0000	-0.10162E+06	-86058.	
3	127.80	9.1100	-0.10162E+06	-86056.	
4	128.30	9.2200	-0.10162E+06	-86052.	
5	146.32	9.3300	-0.22988E+06	-85851.	
6	146.75	9.4400	-0.22960E+06	-85852.	
7	147.37	9.5500	-0.22929E+06	-85876.	
8	149.04	9.6600	-0.22901E+06	-86008.	
9	162.39	9.7700	-0.23445E+06	-86593.	
10	428.22	9.8800	-0.24421E+06	-94578.	.31596+08

STAGE	FLOW RATE LBMOL/HR	FEED RATE LBMOL/HR	PRODUCT RATE LBMOL/HR

	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	LIQUID	VAPOR
1	289.2	0.000				262.9394	
2	26.42	289.2					
3	26.34	289.4					
4	22.66	289.3					
5	1114.	285.6	1076.8426				
6	1116.	300.3					
7	1119.	302.6					
8	1123.	305.2					
9	1097.	308.7					
10	813.9	282.7				813.9031	

**** MASS FLOW PROFILES ****

STAGE	FLOW RATE LB/HR		FEED RATE LB/HR			PRODUCT RATE LB/HR	
	LIQUID	VAPOR	LIQUID	VAPOR	MIXED	LIQUID	VAPOR
1	9262.	0.000				8420.0000	
2	845.0	9262.					
3	842.3	9265.					
4	724.8	9262.					
5	0.2360E+06	9145.	.23485+06				
6	0.2361E+06	9615.					
7	0.2362E+06	9681.					
8	0.2361E+06	9728.					
9	0.2362E+06	9660.					
10	0.2264E+06	9741.				.22643+06	

**** MOLE-X-PROFILE ****

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2
1	0.16679E-09	0.51854E-16	0.34046E-18	0.99860	0.56174E-17
2	0.85794E-12	0.18340E-11	0.53228E-13	0.99618	0.25823E-12
3	0.77798E-12	0.58647E-08	0.75005E-09	0.99558	0.10730E-08
4	0.77804E-12	0.18539E-04	0.10408E-04	0.99538	0.44063E-05
5	0.18546E-09	0.47477	0.55663E-02	0.27240	0.13326
6	0.18610E-09	0.47381	0.55550E-02	0.27341	0.13299
7	0.18757E-09	0.47269	0.55420E-02	0.27278	0.13268
8	0.19107E-09	0.47123	0.55258E-02	0.26360	0.13227
9	0.19387E-09	0.48400	0.16486E-01	0.19261	0.13581
10	0.19661E-09	0.64996	0.76200E-02	0.46453E-02	0.18244

**** MOLE-X-PROFILE ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.28843E-10	0.63174E-14	0.19707E-10	0.21866E-15	0.13955E-02
2	0.49840E-11	0.44232E-10	0.23946E-12	0.67044E-15	0.38232E-02
3	0.46007E-11	0.28071E-07	0.21850E-12	0.78826E-15	0.44241E-02
4	0.45929E-11	0.17615E-04	0.21963E-12	0.82357E-15	0.45658E-02
5	0.51822E-10	0.79952E-01	0.31068E-10	0.15836E-01	0.22701E-02
6	0.51836E-10	0.79790E-01	0.31091E-10	0.15804E-01	0.27277E-02
7	0.52091E-10	0.79603E-01	0.31280E-10	0.15767E-01	0.50562E-02
8	0.53488E-10	0.79356E-01	0.32148E-10	0.15718E-01	0.16478E-01
9	0.57307E-10	0.82149E-01	0.34416E-10	0.16091E-01	0.56657E-01
10	0.60994E-10	0.10945	0.35818E-10	0.21680E-01	0.23791E-02

**** MOLE-X-PROFILE ****

STAGE	K+	H3O+	DIGLY-1	DIGLY-2	DIGLY-3
1	0.10000E-29	0.21866E-15	0.23076E-21	0.52459E-25	0.33600E-26
2	0.10000E-29	0.67044E-15	0.24314E-16	0.33419E-19	0.37932E-20
3	0.10000E-29	0.78826E-15	0.23228E-12	0.19267E-14	0.38642E-15
4	0.10000E-29	0.82357E-15	0.21986E-08	0.10987E-09	0.38925E-10
5	0.15836E-01	0.49617E-25	0.15625E-04	0.43848E-05	0.26320E-05
6	0.15804E-01	0.80761E-25	0.15594E-04	0.43759E-05	0.26267E-05
7	0.15767E-01	0.56281E-24	0.15557E-04	0.43656E-05	0.26205E-05
8	0.15718E-01	0.49724E-23	0.15509E-04	0.43521E-05	0.26124E-05
9	0.16091E-01	0.35167E-21	0.15887E-04	0.44564E-05	0.26749E-05
10	0.21680E-01	0.55961E-22	0.21391E-04	0.60028E-05	0.36032E-05

**** MOLE-X-PROFILE ****

STAGE	MONO-1	MONO-2	MONO-3
1	0.99313E-29	0.99789E-29	0.10000E-29
2	0.10000E-29	0.21052E-27	0.17524E-27
3	0.94805E-22	0.62019E-23	0.62563E-22
4	0.11646E-12	0.16030E-13	0.38886E-13
5	0.57577E-04	0.16160E-04	0.96971E-05
6	0.57461E-04	0.16127E-04	0.96775E-05
7	0.57326E-04	0.16089E-04	0.96548E-05
8	0.57148E-04	0.16039E-04	0.96248E-05
9	0.58505E-04	0.16420E-04	0.98534E-05
10	0.78823E-04	0.22123E-04	0.13275E-04

**** MOLE-Y-PROFILE ****

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2
1	0.31906E-07	0.13395E-20	0.19263E-23	0.99950	0.11140E-21
2	0.16679E-09	0.51854E-16	0.34046E-18	0.99860	0.56174E-17
3	0.15164E-09	0.16747E-12	0.48595E-14	0.99838	0.23579E-13
4	0.15167E-09	0.53399E-09	0.68294E-10	0.99833	0.97696E-10
5	0.15362E-09	0.14709E-05	0.82573E-06	0.99835	0.34959E-06
6	0.15525E-09	0.14803E-05	0.82427E-06	0.99802	0.35187E-06
7	0.15785E-09	0.15041E-05	0.82639E-06	0.99633	0.35752E-06
8	0.16347E-09	0.16074E-05	0.85333E-06	0.98780	0.38193E-06
9	0.17647E-09	0.29385E-05	0.46373E-05	0.94634	0.69838E-06
10	0.18600E-09	0.61671E-02	0.42015E-01	0.73380	0.15573E-02

**** MOLE-Y-PROFILE ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.16599E-09	0.83570E-18	0.16803E-08	0.59449-124	0.50271E-03
2	0.28843E-10	0.63174E-14	0.19707E-10	0.18261-123	0.13955E-02
3	0.26665E-10	0.40437E-11	0.17930E-10	0.21475-123	0.16171E-02
4	0.26635E-10	0.25560E-08	0.17933E-10	0.22441-123	0.16713E-02

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      5    0.26919E-10   0.13976E-05   0.18161E-10   0.43091-110
0.16470E-02
      6    0.26966E-10   0.14048E-05   0.18198E-10   0.43012-110
0.19747E-02
      7    0.27201E-10   0.14253E-05   0.18377E-10   0.42918-110
0.36652E-02
      8    0.28349E-10   0.15188E-05   0.19181E-10   0.42787-110
0.12195E-01
      9    0.33699E-10   0.27296E-05   0.22475E-10   0.43758-110
0.53649E-01
     10    0.46690E-10   0.35284E-02   0.30381E-10   0.58279-110   0.21294

                         ****   MOLE-Y-PROFILE     ****
   STAGE     K+            H3O+          DIGLY-1       DIGLY-2
DIGLY-3
      1    0.27188-138   0.59449-124   0.20545E-26   0.75916E-31
0.27412E-32
      2    0.27238-138   0.18261-123   0.23076E-21   0.52457E-25
0.33685E-26
      3    0.27243-138   0.21475-123   0.22198E-17   0.30508E-20
0.34629E-21
      4    0.27249-138   0.22441-123   0.21150E-13   0.17543E-15
0.35185E-16
      5    0.43091-110   0.13501-133   0.17444E-09   0.87172E-11
0.30883E-11
      6    0.43012-110   0.21980-133   0.17453E-09   0.87334E-11
0.30944E-11
      7    0.42918-110   0.15320-132   0.17540E-09   0.87953E-11
0.31182E-11
      8    0.42787-110   0.13536-131   0.18079E-09   0.91209E-11
0.32436E-11
      9    0.43758-110   0.95634-130   0.25332E-09   0.13419E-10
0.49140E-11
     10    0.58279-110   0.15043-130   0.39210E-07   0.40611E-08
0.22023E-08

                         ****   MOLE-Y-PROFILE     ****
   STAGE     MONO-1        MONO-2        MONO-3
      1    0.57496E-39   0.27170E-39   0.11455E-39
      2    0.71049E-40   0.70958E-38   0.24612E-37
      3    0.68832E-32   0.21381E-33   0.89781E-32
      4    0.86323E-23   0.56470E-24   0.56966E-23
      5    0.92396E-14   0.12718E-14   0.30852E-14
      6    0.93437E-14   0.12870E-14   0.31196E-14
      7    0.95541E-14   0.13174E-14   0.31896E-14
      8    0.10383E-13   0.14357E-14   0.34669E-14
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9	0.22487E-13	0.31782E-14	0.75254E-14
10	0.12933E-08	0.24653E-09	0.44548E-09

**** K-VALUES ****

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2
1	191.30	0.25832E-04	0.56579E-05	1.0009	0.19831E-04
2	194.40	0.28273E-04	0.63963E-05	1.0024	0.21753E-04
3	194.91	0.28556E-04	0.64789E-05	1.0028	0.21976E-04
4	194.94	0.28804E-04	0.65619E-05	1.0030	0.22172E-04
5	0.82830	0.30981E-05	0.14835E-03	3.6650	0.26233E-05
6	0.83421	0.31244E-05	0.14838E-03	3.6502	0.26458E-05
7	0.84154	0.31819E-05	0.14911E-03	3.6524	0.26946E-05
8	0.85553	0.34110E-05	0.15443E-03	3.7473	0.28875E-05
9	0.91022	0.60712E-05	0.28128E-03	4.9133	0.51423E-05
10	0.94603	0.94884E-02	5.5138	157.97	0.85361E-02

**** K-VALUES ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	5.7550	0.13229E-03	85.264	0.0000	0.36023
2	5.7870	0.14282E-03	82.301	0.0000	0.36501
3	5.7958	0.14405E-03	82.060	0.0000	0.36553
4	5.7993	0.14510E-03	81.649	0.0000	0.36604
5	0.51944	0.17480E-04	0.58456	0.0000	0.72554
6	0.52023	0.17607E-04	0.58531	0.0000	0.72395
7	0.52219	0.17905E-04	0.58749	0.0000	0.72490
8	0.53000	0.19139E-04	0.59664	0.0000	0.74008
9	0.58803	0.33227E-04	0.65303	0.0000	0.94691
10	0.76549	0.32237E-01	0.84821	0.0000	89.503

**** K-VALUES ****

STAGE	K+	H3O+	DIGLY-1	DIGLY-2	DIGLY-3
1	0.0000	0.0000	0.89034E-05	0.14472E-05	0.81583E-06
2	0.0000	0.0000	0.94907E-05	0.15697E-05	0.88803E-06

3	0.0000	0.0000	0.95566E-05	0.15835E-05	0.89613E-06
4	0.0000	0.0000	0.96196E-05	0.15967E-05	0.90391E-06
5	0.0000	0.0000	0.11164E-04	0.19880E-05	0.11734E-05
6	0.0000	0.0000	0.11192E-04	0.19958E-05	0.11781E-05
7	0.0000	0.0000	0.11274E-04	0.20147E-05	0.11899E-05
8	0.0000	0.0000	0.11657E-04	0.20958E-05	0.12416E-05
9	0.0000	0.0000	0.15946E-04	0.30111E-05	0.18371E-05
10	0.0000	0.0000	0.18330E-02	0.67653E-03	0.61120E-03

**** K-VALUES ****

STAGE	MONO-1	MONO-2	MONO-3
1	0.57894E-10	0.27228E-10	0.11455E-09
2	0.71049E-10	0.33705E-10	0.14045E-09
3	0.72603E-10	0.34474E-10	0.14350E-09
4	0.74124E-10	0.35227E-10	0.14649E-09
5	0.16047E-09	0.78702E-10	0.31816E-09
6	0.16261E-09	0.79805E-10	0.32236E-09
7	0.16666E-09	0.81879E-10	0.33037E-09
8	0.18169E-09	0.89511E-10	0.36021E-09
9	0.38436E-09	0.19355E-09	0.76374E-09
10	0.16408E-04	0.11144E-04	0.33556E-04

**** RATES OF GENERATION ****

LBMOL/HR

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2	TRIGLY-2
1	0.000	0.000	0.000	0.000	0.000	0.000
2	0.000	0.000	0.000	0.000	0.000	0.000
3	0.000	0.000	0.000	0.000	0.000	0.000
4	0.000	0.000	0.000	0.000	0.000	0.000
5	0.000	0.000	0.000	0.000	0.000	0.000
6	0.000	0.000	0.000	0.000	0.000	0.000

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      7    0.000       0.000       0.000       0.000       0.000
0.000
      8    0.000       0.000       0.000       0.000       0.000
0.000
      9    0.000       0.000       0.000       0.000       0.000
0.000
     10    0.000       0.000       0.000       0.000       0.000
0.000

                     ****     RATES OF GENERATION      ****
                                LBMOL/HR
   STAGE     FAME-3      TRIGLY-3    OH-         WATER       K+
H3O+
      1    0.000       0.000      0.6324E-13  -.1265E-12   0.000
0.6324E-13
      2    0.000       0.000      0.1196E-13  -.2392E-13   0.000
0.1196E-13
      3    0.000       0.000      0.3053E-14  -.6105E-14   0.000
0.3053E-14
      4    0.000       0.000      -.2101E-14  0.4202E-14   0.000      -.
2101E-14
      5    0.000       0.000      -.1866E-13  0.3732E-13   0.000      -.
1866E-13
      6    0.000       0.000      0.4582E-22  -.9165E-22   0.000
0.4582E-22
      7    0.000       0.000      0.1937E-21  -.3873E-21   0.000
0.1937E-21
      8    0.000       0.000      0.5264E-20  -.1053E-19   0.000
0.5264E-20
      9    0.000       0.000      0.3803E-18  -.7607E-18   0.000
0.3803E-18
     10    0.000       0.000      -.3403E-18  0.6806E-18   0.000      -.
3403E-18

                     ****     RATES OF GENERATION      ****
                                LBMOL/HR
   STAGE     DIGLY-1     DIGLY-2     DIGLY-3     MONO-1      MONO-2
MONO-3
      1    0.000       0.000       0.000       0.000       0.000
0.000
      2    0.000       0.000       0.000       0.000       0.000
0.000
      3    0.000       0.000       0.000       0.000       0.000
0.000
      4    0.000       0.000       0.000       0.000       0.000
0.000
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5	0.000	0.000	0.000	0.000	0.000	0.000
6	0.000	0.000	0.000	0.000	0.000	0.000
7	0.000	0.000	0.000	0.000	0.000	0.000
8	0.000	0.000	0.000	0.000	0.000	0.000
9	0.000	0.000	0.000	0.000	0.000	0.000
10	0.000	0.000	0.000	0.000	0.000	0.000

**** MASS-X-PROFILE ****

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2
1	0.46118E-08	0.48011E-15	0.97913E-18	0.99921	0.51657E-16
2	0.23748E-10	0.16999E-10	0.15324E-12	0.99785	0.23772E-11
3	0.21540E-10	0.54373E-07	0.21599E-08	0.99751	0.98800E-08
4	0.21536E-10	0.17183E-03	0.29963E-04	0.99704	0.40562E-04
5	0.77516E-09	0.66447	0.24198E-02	0.41202E-01	0.18524
6	0.77921E-09	0.66429	0.24191E-02	0.41427E-01	0.18519
7	0.78704E-09	0.66415	0.24187E-02	0.41421E-01	0.18515
8	0.80446E-09	0.66435	0.24198E-02	0.40163E-01	0.18521
9	0.79706E-09	0.66631	0.70497E-02	0.28655E-01	0.18569
10	0.62575E-09	0.69269	0.25225E-02	0.53502E-03	0.19311

**** MASS-X-PROFILE ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.79208E-09	0.53355E-13	0.49685E-09	0.11614E-15	0.78508E-03
2	0.13702E-09	0.37397E-09	0.60434E-11	0.35646E-15	0.21532E-02
3	0.12651E-09	0.23740E-06	0.55160E-11	0.41922E-15	0.24922E-02
4	0.12626E-09	0.14893E-03	0.55431E-11	0.43788E-15	0.25713E-02
5	0.21512E-09	0.10207	0.11840E-09	0.12714E-02	0.19305E-03
6	0.21555E-09	0.10204	0.11869E-09	0.12710E-02	0.23237E-03
7	0.21708E-09	0.10202	0.11967E-09	0.12708E-02	0.43166E-03

8	0.22366E-09	0.10205	0.12341E-09	0.12711E-02	0.14115E-02
9	0.23400E-09	0.10316	0.12901E-09	0.12707E-02	0.47392E-02
10	0.19280E-09	0.10641	0.10394E-09	0.13254E-02	0.15406E-03

**** MASS-X-PROFILE ****

STAGE	K+	H3O+	DIGLY-1	DIGLY-2	DIGLY-3
1	0.12209E-29	0.12989E-15	0.44750E-20	0.10107E-23	0.59694E-25
2	0.12222E-29	0.39869E-15	0.47201E-15	0.64455E-18	0.67463E-19
3	0.12226E-29	0.46888E-15	0.45105E-11	0.37170E-13	0.68744E-14
4	0.12222E-29	0.48975E-15	0.42682E-07	0.21191E-08	0.69228E-09
5	0.29227E-02	0.44554E-26	0.45803E-04	0.12770E-04	0.70683E-05
6	0.29219E-02	0.72646E-26	0.45791E-04	0.12766E-04	0.70664E-05
7	0.29213E-02	0.50735E-25	0.45781E-04	0.12764E-04	0.70649E-05
8	0.29221E-02	0.44977E-24	0.45795E-04	0.12768E-04	0.70670E-05
9	0.29211E-02	0.31061E-22	0.45808E-04	0.12766E-04	0.70661E-05
10	0.30468E-02	0.38264E-23	0.47748E-04	0.13312E-04	0.73685E-05

**** MASS-X-PROFILE ****

STAGE	MONO-1	MONO-2	MONO-3
1	0.11058E-27	0.11048E-27	0.10321E-28
2	0.11146E-28	0.23332E-26	0.18106E-26
3	0.10570E-20	0.68754E-22	0.64658E-21
4	0.12980E-11	0.17766E-12	0.40177E-12
5	0.96905E-04	0.27044E-04	0.15129E-04
6	0.96878E-04	0.27037E-04	0.15125E-04
7	0.96859E-04	0.27031E-04	0.15122E-04
8	0.96887E-04	0.27039E-04	0.15126E-04
9	0.96854E-04	0.27030E-04	0.15121E-04
10	0.10102E-03	0.28192E-04	0.15771E-04

**** MASS-Y-PROFILE ****

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2

1	0.88187E-06	0.12397E-19	0.55377E-23	0.99972	0.10240E-20
2	0.46118E-08	0.48011E-15	0.97913E-18	0.99921	0.51657E-16
3	0.41934E-08	0.15508E-11	0.13977E-13	0.99909	0.21685E-12
4	0.41944E-08	0.49448E-08	0.19643E-09	0.99906	0.89851E-09
5	0.42480E-08	0.13620E-04	0.23750E-05	0.99904	0.32151E-05
6	0.42937E-08	0.13709E-04	0.23711E-05	0.99886	0.32365E-05
7	0.43688E-08	0.13940E-04	0.23789E-05	0.99790	0.32909E-05
8	0.45413E-08	0.14953E-04	0.24657E-05	0.99307	0.35288E-05
9	0.49934E-08	0.27843E-04	0.13648E-04	0.96904	0.65723E-05
10	0.47793E-08	0.53063E-01	0.11229	0.68233	0.13308E-01

**** MASS-Y-PROFILE ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.45566E-08	0.70554E-17	0.42347E-07	0.31562-124	0.28270E-03
2	0.79208E-09	0.53355E-13	0.49685E-09	0.96990-124	0.78508E-03
3	0.73233E-09	0.34155E-10	0.45209E-09	0.11407-123	0.90985E-03
4	0.73155E-09	0.21590E-07	0.45217E-09	0.11921-123	0.94033E-03
5	0.73930E-09	0.11804E-04	0.45791E-09	0.22888-110	0.92667E-03
6	0.74072E-09	0.11868E-04	0.45890E-09	0.22850-110	0.11112E-02
7	0.74772E-09	0.12049E-04	0.46376E-09	0.22817-110	0.20640E-02
8	0.78219E-09	0.12888E-04	0.48587E-09	0.22833-110	0.68930E-02
9	0.94705E-09	0.23592E-04	0.57986E-09	0.23784-110	0.30887E-01
10	0.11915E-08	0.27693E-01	0.71178E-09	0.28764-110	0.11132

**** MASS-Y-PROFILE ****

STAGE	K+	H3O+	DIGLY-1	DIGLY-2	DIGLY-3

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      1    0.33182-138   0.35301-124   0.39827E-25   0.14621E-29
0.48681E-31
      2    0.33256-138   0.10848-123   0.44750E-20   0.10107E-23
0.59845E-25
      3    0.33266-138   0.12758-123   0.43052E-16   0.58785E-19
0.61528E-20
      4    0.33273-138   0.13333-123   0.41020E-12   0.33804E-14
0.62518E-15
      5    0.52616-110   0.80208-134   0.33831E-08   0.16796E-09
0.54872E-10
      6    0.52527-110   0.13060-133   0.33853E-08   0.16830E-09
0.54989E-10
      7    0.52451-110   0.91093-133   0.34047E-08   0.16962E-09
0.55452E-10
      8    0.52487-110   0.80788-132   0.35225E-08   0.17656E-09
0.57899E-10
      9    0.54674-110   0.58137-130   0.50274E-08   0.26457E-09
0.89344E-10
     10    0.66123-110   0.83043-131   0.70662E-06   0.72710E-07
0.36359E-07

                       ****   MASS-Y-PROFILE     ****
   STAGE     MONO-1        MONO-2        MONO-3
      1    0.63993E-38   0.30069E-38   0.11818E-38
      2    0.79107E-39   0.78560E-37   0.25402E-36
      3    0.76646E-31   0.23673E-32   0.92673E-31
      4    0.96125E-22   0.62526E-23   0.58802E-22
      5    0.10288E-12   0.14082E-13   0.31845E-13
      6    0.10406E-12   0.14252E-13   0.32205E-13
      7    0.10648E-12   0.14599E-13   0.32952E-13
      8    0.11616E-12   0.15970E-13   0.35951E-13
      9    0.25622E-12   0.36008E-13   0.79485E-13
     10    0.13382E-07   0.25364E-08   0.42727E-08

 BLOCK:  H-101    MODEL: HEATX
 -----------------------------
   HOT SIDE:
   ---------
   INLET STREAM:          65
   OUTLET STREAM:         66
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES
   COLD SIDE:
   ----------
   INLET STREAM:          28
```

```
   OUTLET STREAM:          29
   PROPERTY OPTION SET:    ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:         GLOBAL
   CHEMISTRY ID:           GLOBAL   - TRUE SPECIES

                    ***  MASS AND ENERGY BALANCE  ***
                                  IN               OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          12598.7        12598.7        0.00000
       MASS(LB/HR   )          443682.        443682.        0.00000
       ENTHALPY(BTU/HR  )     -0.165410E+10   -0.165410E+10
0.119134E-08

                        ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                  30
   CONVERGENCE TOLERANCE                                    0.000100000

   FLASH SPECS FOR COLD SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                  30
   CONVERGENCE TOLERANCE                                    0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED COLD OUTLET TEMP
     SPECIFIED VALUE               F                      140.0000
     LMTD CORRECTION FACTOR                                 1.00000

   PRESSURE SPECIFICATION:
     HOT  SIDE PRESSURE DROP       PSI                      7.0000
     COLD SIDE PRESSURE DROP       PSI                      7.0000

   HEAT TRANSFER COEFFICIENT SPECIFICATION:
     HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R         149.6937
```

```
                    ***  OVERALL RESULTS  ***

   STREAMS:
                   --------------------------------------
                   |                                    |
   65        ----->|               HOT                  |-----> 66
   T=  1.6558D+02  |                                    |       T=  1.4965D
+02
   P=  2.9996D+01  |                                    |       P=  2.2996D
+01
   V=  0.0000D+00  |                                    |       V=  0.0000D
+00
                   |                                    |
   29        <-----|               COLD                 |<----- 28
   T=  1.4000D+02  |                                    |       T=  1.0000D
+02
   P=  3.9700D+01  |                                    |       P=  4.6700D
+01
   V=  0.0000D+00  |                                    |       V=  0.0000D
+00
                   --------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY            BTU/HR              3296980.6677
     CALCULATED (REQUIRED) AREA      SQFT                    606.8277
     ACTUAL EXCHANGER AREA           SQFT                    606.8277
     PER CENT OVER-DESIGN                                      0.0000

   HEAT TRANSFER COEFFICIENT:
     AVERAGE COEFFICIENT (DIRTY)     BTU/HR-SQFT-R           149.6937
     UA (DIRTY)                      BTU/HR-R              90838.3101

   LOG-MEAN TEMPERATURE DIFFERENCE:
     LMTD CORRECTION FACTOR                                    1.0000
     LMTD (CORRECTED)                F                        36.2950
     NUMBER OF SHELLS IN SERIES                                 1

   PRESSURE DROP:
     HOTSIDE, TOTAL                  PSI                       7.0000
     COLDSIDE, TOTAL                 PSI                       7.0000

   PRESSURE DROP PARAMETER:
     HOT SIDE:                                             68455.
     COLD SIDE:                                            44456.
```

```
                    ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:

                                    HOT
         -------------------------------------------------------------
         |                                                           |
 65      |                          LIQ                              | 66
 ------> |
|------>
  165.6  |                                                           |
149.6
         |                                                           |
 29      |                          LIQ                              | 28
 <------ |                                                           |
<------
  140.0  |                                                           |
100.0
         |                                                           |
         -------------------------------------------------------------
                                    COLD

   ZONE HEAT TRANSFER AND AREA:

   ZONE       HEAT DUTY       AREA       LMTD       AVERAGE U       UA
              BTU/HR          SQFT        F         BTU/HR-SQFT-R   BTU/HR-
R
     1    3296978.697       606.8277      36.2950   149.6937
90838.2558

 BLOCK:  H-102    MODEL: HEATX
 -----------------------------
   HOT SIDE:
   ---------
   INLET STREAM:          38
   OUTLET STREAM:         39
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES
   COLD SIDE:
   ----------
   INLET STREAM:          42
   OUTLET STREAM:         43
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES
```

```
                    ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT         RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          892.027         892.027         0.00000
       MASS(LB/HR   )          228186.         228186.
0.255089E-15
       ENTHALPY(BTU/HR  )    -0.206748E+09   -0.207519E+09
0.371597E-02

                        ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                  100
   CONVERGENCE TOLERANCE                                     0.000100000

   FLASH SPECS FOR COLD SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                  100
   CONVERGENCE TOLERANCE                                     0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED COLD OUTLET TEMP
     SPECIFIED VALUE                F                      156.8300
     LMTD CORRECTION FACTOR                                  1.00000

   PRESSURE SPECIFICATION:
     HOT  SIDE PRESSURE DROP        PSI                      7.0000
     COLD SIDE PRESSURE DROP        PSI                      7.0000

   HEAT TRANSFER COEFFICIENT SPECIFICATION:
     HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R          149.6937
     HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R          149.6937
     HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R          149.6937
     HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R          149.6937
     HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R          149.6937
     HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R          149.6937
     HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R          149.6937
     HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R          149.6937
     HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R          149.6937

                      ***  OVERALL RESULTS  ***
```

```
   STREAMS:
                    --------------------------------------
                    |                                    |
   38        ----->|               HOT                  |-----> 39
   T=  4.2883D+02  |                                    |       T=  4.2546D
+02
   P=  4.1730D+01  |                                    |       P=  3.4730D
+01
   V=  0.0000D+00  |                                    |       V=  0.0000D
+00
                    |                                    |
   43        <-----|               COLD                 |<----- 42
   T=  1.5683D+02  |                                    |       T=  7.6730D
+01
   P=  3.0700D+01  |                                    |       P=  3.7700D
+01
   V=  0.0000D+00  |                                    |       V=
2.3608D-01
                    --------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY            BTU/HR               385566.5620
     CALCULATED (REQUIRED) AREA      SQFT                      8.3416
     ACTUAL EXCHANGER AREA           SQFT                      8.3416
     PER CENT OVER-DESIGN                                      0.0000

   HEAT TRANSFER COEFFICIENT:
     AVERAGE COEFFICIENT (DIRTY)     BTU/HR-SQFT-R           149.6937
     UA (DIRTY)                      BTU/HR-R               1248.6858

   LOG-MEAN TEMPERATURE DIFFERENCE:
     LMTD CORRECTION FACTOR                                    1.0000
     LMTD (CORRECTED)                F                       308.7779
     NUMBER OF SHELLS IN SERIES                                 1

   PRESSURE DROP:
     HOTSIDE, TOTAL                  PSI                       7.0000
     COLDSIDE, TOTAL                 PSI                       7.0000

   PRESSURE DROP PARAMETER:
     HOT SIDE:                                               44483.
     COLD SIDE:                                             0.19695E+08

                         ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:
```

```
                            HOT
         -------------------------------------------------------------
         |                                                           |
  38     |                  LIQ                                      | 39
  ------> |
|------>
   428.8  |                                                          |
425.5
         |                                                           |
  43     |                  LIQ                                      | 42
  <------ |                                                          |
<------
   156.8  |                                                          |
76.7
         |                                                           |
         -------------------------------------------------------------
                            COLD

   ZONE HEAT TRANSFER AND AREA:

   ZONE       HEAT DUTY       AREA       LMTD       AVERAGE U       UA
              BTU/HR          SQFT        F         BTU/HR-SQFT-R   BTU/HR-
R
     1     385566.373         8.3416     308.7779   149.6937
1248.6852

  BLOCK:  H-103    MODEL: HEATX
  -----------------------------
   HOT SIDE:
   ---------
   INLET STREAM:           39
   OUTLET STREAM:          40
   PROPERTY OPTION SET:    ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:         GLOBAL
   CHEMISTRY ID:           GLOBAL   - TRUE SPECIES
   COLD SIDE:
   ----------
   INLET STREAM:           47
   OUTLET STREAM:          48
   PROPERTY OPTION SET:    ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:         GLOBAL
   CHEMISTRY ID:           GLOBAL   - TRUE SPECIES

                        ***  MASS AND ENERGY BALANCE  ***
                                  IN                OUT        RELATIVE
```

```
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           17953.8         17953.8         0.00000
       MASS(LB/HR   )           535211.         535211.         0.00000
       ENTHALPY(BTU/HR  )      -0.230536E+10   -0.230536E+10
0.137298E-10

                          ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                   100
   CONVERGENCE TOLERANCE                                      0.000100000

   FLASH SPECS FOR COLD SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                   100
   CONVERGENCE TOLERANCE                                      0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED HOT OUTLET TEMP
     SPECIFIED VALUE               F                        156.8300
     LMTD CORRECTION FACTOR                                   1.00000

   PRESSURE SPECIFICATION:
     HOT  SIDE PRESSURE DROP       PSI                        7.0000
     COLD SIDE PRESSURE DROP       PSI                        7.0000

   HEAT TRANSFER COEFFICIENT SPECIFICATION:
     HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R           149.6937
     HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R           149.6937
     HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R           149.6937
     HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R           149.6937
     HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R           149.6937
     HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R           149.6937
     HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R           149.6937
     HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R           149.6937
     HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R           149.6937

                        ***  OVERALL RESULTS  ***

   STREAMS:
                   --------------------------------------
                   |                                    |
   39        ----->|               HOT                  |-----> 40
```

```
   T=  4.2546D+02  |                                     |       T=  1.5683D
+02
   P=  3.4730D+01  |                                     |       P=  2.7730D
+01
   V=  0.0000D+00  |                                     |       V=  0.0000D
+00
                   |                                     |
   48        <-----|               COLD                  |<----- 47
   T=  1.6529D+02  |                                     |       T=  8.1074D
+01
   P=  2.2696D+01  |                                     |       P=  2.9696D
+01
   V=  0.0000D+00  |                                     |       V=  0.0000D
+00
                   ---------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY           BTU/HR           25989376.7149
     CALCULATED (REQUIRED) AREA     SQFT                 1161.5714
     ACTUAL EXCHANGER AREA          SQFT                 1161.5714
     PER CENT OVER-DESIGN                                   0.0000

   HEAT TRANSFER COEFFICIENT:
     AVERAGE COEFFICIENT (DIRTY)    BTU/HR-SQFT-R         149.6937
     UA (DIRTY)                     BTU/HR-R           173879.8661

   LOG-MEAN TEMPERATURE DIFFERENCE:
     LMTD CORRECTION FACTOR                                 1.0000
     LMTD (CORRECTED)               F                     149.4674
     NUMBER OF SHELLS IN SERIES                             1

   PRESSURE DROP:
     HOTSIDE, TOTAL                 PSI                     7.0000
     COLDSIDE, TOTAL                PSI                     7.0000

   PRESSURE DROP PARAMETER:
     HOT SIDE:                                            47280.
     COLD SIDE:                                           31435.

                        ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:

                                  HOT
         -------------------------------------------------------------
         |                                                           |
```

```
       ENTHALPY(BTU/HR  )      -0.114921E+10   -0.114921E+10
-0.173611E-08

                          ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                 100
   CONVERGENCE TOLERANCE                                    0.000100000

   FLASH SPECS FOR COLD SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                 100
   CONVERGENCE TOLERANCE                                    0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED COLD OUTLET TEMP
     SPECIFIED VALUE                F                      76.7300
     LMTD CORRECTION FACTOR                                 1.00000

   PRESSURE SPECIFICATION:
     HOT  SIDE PRESSURE DROP        PSI                     7.0000
     COLD SIDE PRESSURE DROP        PSI                     7.0000

   HEAT TRANSFER COEFFICIENT SPECIFICATION:
     HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R         149.6937
     HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R         149.6937
     HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R         149.6937
     HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R         149.6937
     HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R         149.6937

                        ***  OVERALL RESULTS  ***

   STREAMS:
                   --------------------------------------
                   |                                    |
   52        ----->|                HOT                 |-----> 53
   T=  1.4380D+02  |                                    |       T=  1.3476D
+02
   P=  3.8696D+01  |                                    |       P=  3.1696D
+01
```

```
   V=  0.0000D+00  |                                    |       V=  0.0000D
+00
                   |                                    |
   BIODIESE  <-----|               COLD                 |<----- 58
   T=  7.6730D+01  |                                    |       T=  5.3564D
+01
   P=  1.4700D+01  |                                    |       P=  2.1700D
+01
   V=  0.0000D+00  |                                    |       V=  0.0000D
+00
                   --------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY            BTU/HR             1522201.7808
     CALCULATED (REQUIRED) AREA      SQFT                   137.5875
     ACTUAL EXCHANGER AREA           SQFT                   137.5875
     PER CENT OVER-DESIGN                                     0.0000

   HEAT TRANSFER COEFFICIENT:
     AVERAGE COEFFICIENT (DIRTY)     BTU/HR-SQFT-R          149.6937
     UA (DIRTY)                      BTU/HR-R             20595.9793

   LOG-MEAN TEMPERATURE DIFFERENCE:
     LMTD CORRECTION FACTOR                                   1.0000
     LMTD (CORRECTED)                F                       73.9077
     NUMBER OF SHELLS IN SERIES                                 1

   PRESSURE DROP:
     HOTSIDE, TOTAL                  PSI                      7.0000
     COLDSIDE, TOTAL                 PSI                      7.0000

   PRESSURE DROP PARAMETER:
     HOT SIDE:                                              25196.
     COLD SIDE:                                             53014.

                        ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:

                                                HOT
         -------------------------------------------------------------
         |                                                           |
 52      |                                      LIQ                  | 53
 ------> |
|------>
  143.8  |                                                           |
```

```
134.8
         |                                                           |
 BIODIESE|                          LIQ                              | 58
 <------ |                                                           |
<------
   76.7  |                                                           |
53.6
         |                                                           |
         -------------------------------------------------------------
                                    COLD

   ZONE HEAT TRANSFER AND AREA:

   ZONE       HEAT DUTY       AREA       LMTD       AVERAGE U       UA
              BTU/HR          SQFT        F         BTU/HR-SQFT-R   BTU/HR-
R
     1    1522201.781       137.5875      73.9077   149.6937
20595.9793

 BLOCK:  H-105    MODEL: HEATX
 -----------------------------
   HOT SIDE:
   ---------
   INLET STREAM:            51
   OUTLET STREAM:           52
   PROPERTY OPTION SET:     ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:          GLOBAL
   CHEMISTRY ID:            GLOBAL   - TRUE SPECIES
   COLD SIDE:
   ----------
   INLET STREAM:            61
   OUTLET STREAM:           62
   PROPERTY OPTION SET:     ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:          GLOBAL
   CHEMISTRY ID:            GLOBAL   - TRUE SPECIES

                       ***  MASS AND ENERGY BALANCE  ***
                                 IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           12244.2         12244.2         0.00000
       MASS(LB/HR   )           431438.         431438.
0.134915E-15
       ENTHALPY(BTU/HR  )      -0.162578E+10   -0.162578E+10
0.871919E-08
```

```
                         ***  INPUT DATA  ***

   FLASH SPECS FOR HOT SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                    30
   CONVERGENCE TOLERANCE                                      0.000100000

   FLASH SPECS FOR COLD SIDE:
   ONE    PHASE      FLASH   SPECIFIED PHASE IS  LIQUID
   MAXIMUM NO. ITERATIONS                                    30
   CONVERGENCE TOLERANCE                                      0.000100000

   FLOW DIRECTION AND SPECIFICATION:
     COUNTERCURRENT   HEAT EXCHANGER
     SPECIFIED COLD OUTLET TEMP
     SPECIFIED VALUE                F                        89.3300
     LMTD CORRECTION FACTOR                                   1.00000

   PRESSURE SPECIFICATION:
     HOT  SIDE PRESSURE DROP        PSI                       7.0000
     COLD SIDE PRESSURE DROP        PSI                       7.0000

   HEAT TRANSFER COEFFICIENT SPECIFICATION:
     HOT LIQUID    COLD LIQUID      BTU/HR-SQFT-R          149.6937
     HOT 2-PHASE   COLD LIQUID      BTU/HR-SQFT-R          149.6937
     HOT VAPOR     COLD LIQUID      BTU/HR-SQFT-R          149.6937
     HOT LIQUID    COLD 2-PHASE     BTU/HR-SQFT-R          149.6937
     HOT 2-PHASE   COLD 2-PHASE     BTU/HR-SQFT-R          149.6937
     HOT VAPOR     COLD 2-PHASE     BTU/HR-SQFT-R          149.6937
     HOT LIQUID    COLD VAPOR       BTU/HR-SQFT-R          149.6937
     HOT 2-PHASE   COLD VAPOR       BTU/HR-SQFT-R          149.6937
     HOT VAPOR     COLD VAPOR       BTU/HR-SQFT-R          149.6937

                       ***  OVERALL RESULTS  ***

   STREAMS:
                   --------------------------------------
                   |                                    |
   51        ----->|               HOT                  |-----> 52
   T=  1.6578D+02  |                                    |       T=  1.4380D
+02
   P=  4.5696D+01  |                                    |       P=  3.8696D
+01
   V=  0.0000D+00  |                                    |       V=  0.0000D
+00
                   |                                    |
```

```
   62        <-----|                COLD                |<----- 61
   T=  8.9330D+01  |                                    |       T=  5.3330D
+01
   P=  3.9700D+01  |                                    |       P=  4.6700D
+01
   V=  0.0000D+00  |                                    |       V=  0.0000D
+00
                   --------------------------------------

   DUTY AND AREA:
     CALCULATED HEAT DUTY            BTU/HR             3714950.0856
     CALCULATED (REQUIRED) AREA      SQFT                   298.0487
     ACTUAL EXCHANGER AREA           SQFT                   298.0487
     PER CENT OVER-DESIGN                                     0.0000

   HEAT TRANSFER COEFFICIENT:
     AVERAGE COEFFICIENT (DIRTY)     BTU/HR-SQFT-R          149.6942
     UA (DIRTY)                      BTU/HR-R             44616.1707

   LOG-MEAN TEMPERATURE DIFFERENCE:
     LMTD CORRECTION FACTOR                                   1.0000
     LMTD (CORRECTED)                F                       83.2647
     NUMBER OF SHELLS IN SERIES                                 1

   PRESSURE DROP:
     HOTSIDE, TOTAL                  PSI                      7.0000
     COLDSIDE, TOTAL                 PSI                      7.0000

   PRESSURE DROP PARAMETER:
     HOT SIDE:                                              25047.
     COLD SIDE:                                           0.28406E+06

                        ***  ZONE RESULTS  ***

   TEMPERATURE LEAVING EACH ZONE:

                                   HOT
         -------------------------------------------------------------
         |                                                           |
 51      |                         LIQ                               | 52
 ------> |
|------>
  165.8  |                                                           |
143.8
         |                                                           |
 62      |                         LIQ                               | 61
```

```
 <------ |                                                            |
<------
   89.3  |                                                            |
53.3
         |                                                            |
         --------------------------------------------------------------
                                    COLD

   ZONE HEAT TRANSFER AND AREA:

   ZONE       HEAT DUTY       AREA       LMTD       AVERAGE U       UA
              BTU/HR          SQFT        F         BTU/HR-SQFT-R   BTU/HR-
R
     1    3714935.910       298.0487      83.2647   149.6937
44616.0005

 BLOCK:  L-101    MODEL: EXTRACT
 -------------------------------

    INLETS   - 48       STAGE   1
               44       STAGE  15
    OUTLETS  - 49       STAGE   1
               63       STAGE  15
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL

                       ***  MASS AND ENERGY BALANCE  ***
                                     IN                OUT        RELATIVE
DIFF.
    CONVENTIONAL COMPONENTS (LBMOL/HR)
       TRIGLY-1                 0.160019E-06    0.159988E-06
0.199588E-03
       FAME-1                    529.004         529.004
-0.360038E-08
       GLYCEROL                  6.20191         6.20191
0.119151E-12
       METHANOL                  3.78081         3.78081
0.190032E-10
       NAOH                      0.00000         0.00000         0.00000
       FAME-2                    148.487         148.487
-0.357826E-08
       TRIGLY-2                 0.496435E-07    0.496350E-07
0.171978E-03
       FAME-3                    89.0855         89.0855
-0.263467E-08
       TRIGLY-3                 0.291521E-07    0.291472E-07
```

```
0.169861E-03
       DODECANE                  0.00000         0.00000         0.00000
       OH-                     0.135178E-03    0.135168E-03
0.736335E-04
       WATER                     17217.8         17216.7
0.635701E-04
       K+                        17.6452         17.6452
0.129207E-08
       H+                        0.00000         0.00000         0.00000
       CL-                       18.6355         18.6355
0.168234E-08
       KCL                       0.00000         0.00000         0.00000
       H3O+                     0.990466        0.990466
0.141642E-07
       HCL                      0.191740        0.191680
0.310294E-03
       DIGLY-1                 0.174101E-01    0.174047E-01
0.311637E-03
       DIGLY-2                 0.488570E-02    0.488418E-02
0.311637E-03
       DIGLY-3                 0.293266E-02    0.293174E-02
0.311637E-03
       MONO-1                  0.641546E-01    0.641346E-01
0.311637E-03
       MONO-2                  0.180059E-01    0.180003E-01
0.311637E-03
       MONO-3                  0.108049E-01    0.108015E-01
0.311637E-03
    TOTAL BALANCE
       MOLE(LBMOL/HR)            18031.9         18030.8
0.607051E-04
       MASS(LB/HR   )            536966.         536946.
0.367524E-04
       ENTHALPY(BTU/HR  )      -0.231380E+10   -0.231647E+10
0.115270E-02


                         **********************
                         ****  INPUT DATA  ****
                         **********************

   ****   INPUT PARAMETERS   ****

    NUMBER OF THEORETICAL STAGES                                15
    MAXIMUM NO. OF OUTSIDE LOOPS                               100
```

```
 MAXIMUM NO. OF INSIDE LOOPS PER OUTSIDE LOOP            10
 OUTSIDE LOOP CONVERGENCE TOLERANCE                       0.0100000
 QMIN FOR BOUNDED WEGSTEIN IN OUTSIDE LOOP                0.0
 QMAX FOR BOUNDED WEGSTEIN IN OUTSIDE LOOP                0.50000
 QMIN FOR BOUNDED WEGSTEIN IN INSIDE LOOP                 0.0
 QMAX FOR BOUNDED WEGSTEIN IN INSIDE LOOP                 0.50000
 OUTSIDE LOOP ERROR THRESHOLD FOR BROYDEN METHOD          0.100000


****   KEY COMPONENT SPECIFICATIONS   ****

 KEY COMPONENTS FOR LIQUID1                   WATER
                                              K+
                                              CL-
                                              H3O+

 KEY COMPONENTS FOR LIQUID2                   FAME-1
                                              FAME-2
                                              FAME-3


****    PROFILES   ****

 TEMP-SPEC       STAGE   1  TEMP  F                     165.400
                         2                              165.400
                         3                              165.400
                         4                              165.400
                         5                              165.400
                         6                              165.400

 P-SPEC          STAGE   1  PRES  PSIA                   14.6959

                       *******************
                       ****  RESULTS  ****
                       *******************


***   COMPONENT SPLIT FRACTIONS   ***

                          OUTLET STREAMS
                          --------------
              49            63
 COMPONENT:
 TRIGLY-1    .36317       .63683
 FAME-1      .99998       .23112E-04
 GLYCEROL    .26597E-10   1.0000
```

1	165.40	14.696	-0.12137E+06	-0.14107E+06	.22590+05
2	165.40	14.696	-0.12137E+06	-0.14107E+06	102.8382
3	165.40	14.696	-0.12137E+06	-0.14107E+06	93.3372
4	165.40	14.696	-0.12137E+06	-0.14107E+06	82.7201
5	165.40	14.696	-0.12137E+06	-0.14107E+06	66.6039
6	165.40	14.696	-0.12137E+06	-0.14107E+06	31.6060
7	165.40	14.696	-0.12137E+06	-0.14106E+06	-52.9569
8	165.40	14.696	-0.12137E+06	-0.14106E+06	-223.6155
9	165.40	14.696	-0.12137E+06	-0.14106E+06	-469.9226
10	165.40	14.696	-0.12137E+06	-0.14107E+06	-1159.0020
11	165.40	14.696	-0.12137E+06	-0.14107E+06	-6589.6407
12	165.40	14.696	-0.12137E+06	-0.14110E+06	-.12953+05
13	165.40	14.696	-0.12136E+06	-0.14115E+06	-.21326+05
14	165.40	14.696	-0.12135E+06	-0.14120E+06	-.87738+05
15	165.40	14.696	-0.12136E+06	-0.14158E+06	-.26956+07

STAGE	FLOW RATE LBMOL/HR		FEED RATE LBMOL/HR	PRODUCT RATE LBMOL/HR	
	LIQUID1	LIQUID2		LIQUID1	LIQUID2
1	0.1714E+05	6509.	.17140+05		6508.8902
2	0.1714E+05	6509.			
3	0.1714E+05	6510.			
4	0.1714E+05	6510.			
5	0.1714E+05	6511.			
6	0.1714E+05	6511.			
7	0.1714E+05	6511.			
8	0.1714E+05	6511.			
9	0.1714E+05	6511.			
10	0.1714E+05	6511.			
11	0.1713E+05	6510.			
12	0.1711E+05	6499.			
13	0.1709E+05	6479.			
14	0.1695E+05	6455.			
15	0.1152E+05	6322.	892.0268	.11522+05	

**** LIQUID1 PROFILE ****

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2
1	0.0000	0.62941E-08	0.48916E-13	0.32447E-12	0.19009E-08
2	0.0000	0.63166E-08	0.29752E-12	0.17093E-11	0.19076E-08
3	0.30229E-14	0.63381E-08	0.15609E-11	0.76193E-11	0.19139E-08

4	0.12329E-13	0.63588E-08	0.79808E-11	0.32839E-10	0.19201E-08
5	0.31140E-13	0.63796E-08	0.40602E-10	0.14045E-09	0.19262E-08
6	0.63040E-13	0.64022E-08	0.20635E-09	0.59958E-09	0.19329E-08
7	0.11927E-12	0.64313E-08	0.10485E-08	0.25584E-08	0.19415E-08
8	0.23722E-12	0.64789E-08	0.53276E-08	0.10916E-07	0.19556E-08
9	0.52252E-12	0.65776E-08	0.27074E-07	0.46581E-07	0.19848E-08
10	0.11686E-11	0.68161E-08	0.13763E-06	0.19885E-06	0.20553E-08
11	0.22978E-11	0.74540E-08	0.70036E-06	0.84974E-06	0.22436E-08
12	0.38047E-11	0.93807E-08	0.35770E-05	0.36455E-05	0.28109E-08
13	0.54575E-11	0.16721E-07	0.18424E-04	0.15783E-04	0.49566E-08
14	0.70744E-11	0.65202E-07	0.96673E-04	0.69713E-04	0.18894E-07
15	0.88437E-11	0.10611E-05	0.53822E-03	0.32811E-03	0.29700E-06

**** LIQUID1 PROFILE ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.0000	0.92508E-08	0.0000	0.78868E-08	1.0000
2	0.0000	0.92682E-08	0.0000	0.78868E-08	1.0000
3	0.0000	0.92847E-08	0.0000	0.78868E-08	0.99999
4	0.0000	0.93008E-08	0.0000	0.78868E-08	0.99999
5	0.0000	0.93168E-08	0.0000	0.78867E-08	0.99999
6	0.0000	0.93342E-08	0.0000	0.78867E-08	0.99999
7	0.0000	0.93565E-08	0.0000	0.78865E-08	0.99999
8	0.0000	0.93932E-08	0.0000	0.78861E-08	0.99998
9	0.36950E-14	0.94691E-08	0.0000	0.78850E-08	0.99998
10	0.32604E-13	0.96508E-08	0.11696E-13	0.78821E-08	0.99996
11	0.17880E-12	0.10124E-07	0.84060E-13	0.78743E-08	0.99992
12	0.50740E-12	0.11462E-07	0.28768E-12	0.78542E-08	0.99980
13	0.93071E-12	0.15766E-07	0.61450E-12	0.78014E-08	0.99949
14	0.13432E-11	0.34733E-07	0.10193E-11	0.76590E-08	0.99861
15	0.17782E-11	0.21276E-06	0.15046E-11	0.72752E-08	0.99587

**** LIQUID1 PROFILE ****

STAGE	K+	CL-	H3O+	HCL	DIGLY-1

```
      1    0.12049E-08   0.12725E-08   0.79544E-08   0.10971E-05
0.10010E-06
      2    0.43768E-08   0.46224E-08   0.81324E-08   0.21352E-05
0.19483E-06
      3    0.12727E-07   0.13441E-07   0.86010E-08   0.31175E-05
0.28445E-06
      4    0.34706E-07   0.36653E-07   0.98346E-08   0.40468E-05
0.36924E-06
      5    0.92559E-07   0.97754E-07   0.13082E-07   0.49261E-05
0.44947E-06
      6    0.24484E-06   0.25858E-06   0.21628E-07   0.57580E-05
0.52538E-06
      7    0.64565E-06   0.68188E-06   0.44123E-07   0.65454E-05
0.59722E-06
      8    0.17006E-05   0.17960E-05   0.10333E-06   0.72913E-05
0.66528E-06
      9    0.44774E-05   0.47287E-05   0.25918E-06   0.79997E-05
0.72992E-06
     10    0.11787E-04   0.12448E-04   0.66940E-06   0.86770E-05
0.79171E-06
     11    0.31029E-04   0.32771E-04   0.17494E-05   0.93368E-05
0.85191E-06
     12    0.81774E-04   0.86364E-04   0.45974E-05   0.10032E-04
0.91530E-06
     13    0.21591E-03   0.22803E-03   0.12126E-04   0.10867E-04
0.99133E-06
     14    0.57122E-03   0.60328E-03   0.32067E-04   0.12100E-04
0.11029E-05
     15    0.15313E-02   0.16172E-02   0.85951E-04   0.14915E-04
0.13536E-05

                         ****   LIQUID1 PROFILE    ****
   STAGE     DIGLY-2       DIGLY-3       MONO-1        MONO-2        MONO-3
      1    0.28092E-07   0.16862E-07   0.36888E-06   0.10353E-06
0.62126E-07
      2    0.54673E-07   0.32817E-07   0.71791E-06   0.20149E-06
0.12091E-06
      3    0.79823E-07   0.47914E-07   0.10482E-05   0.29418E-06
0.17653E-06
      4    0.10362E-06   0.62197E-07   0.13606E-05   0.38188E-06
0.22916E-06
      5    0.12613E-06   0.75711E-07   0.16562E-05   0.46485E-06
0.27894E-06
      6    0.14743E-06   0.88497E-07   0.19360E-05   0.54336E-06
0.32605E-06
      7    0.16759E-06   0.10060E-06   0.22007E-05   0.61766E-06
```

					0.37064E-06
8	0.18669E-06	0.11206E-06	0.24515E-05	0.68804E-06	0.41288E-06
9	0.20483E-06	0.12295E-06	0.26897E-05	0.75489E-06	0.45299E-06
10	0.22217E-06	0.13336E-06	0.29174E-05	0.81880E-06	0.49134E-06
11	0.23907E-06	0.14350E-06	0.31392E-05	0.88106E-06	0.52870E-06
12	0.25685E-06	0.15418E-06	0.33728E-05	0.94662E-06	0.56804E-06
13	0.27819E-06	0.16699E-06	0.36530E-05	0.10253E-05	0.61523E-06
14	0.30950E-06	0.18578E-06	0.40641E-05	0.11406E-05	0.68447E-06
15	0.37984E-06	0.22800E-06	0.49877E-05	0.13999E-05	0.84003E-06

**** LIQUID2 PROFILE ****

STAGE	TRIGLY-1	FAME-1	GLYCEROL	METHANOL	FAME-2
1	0.89284E-11	0.81272E-01	0.25343E-13	0.20017E-12	0.22812E-01
2	0.89277E-11	0.81266E-01	0.15414E-12	0.10546E-11	0.22811E-01
3	0.89277E-11	0.81260E-01	0.80872E-12	0.47009E-11	0.22809E-01
4	0.89345E-11	0.81255E-01	0.41351E-11	0.20261E-10	0.22808E-01
5	0.89584E-11	0.81250E-01	0.21038E-10	0.86660E-10	0.22806E-01
6	0.90075E-11	0.81246E-01	0.10692E-09	0.36997E-09	0.22805E-01
7	0.90911E-11	0.81242E-01	0.54329E-09	0.15787E-08	0.22804E-01
8	0.92389E-11	0.81241E-01	0.27604E-08	0.67357E-08	0.22803E-01
9	0.95496E-11	0.81242E-01	0.14026E-07	0.28739E-07	0.22804E-01
10	0.10301E-10	0.81247E-01	0.71279E-07	0.12264E-06	0.22805E-01
11	0.12004E-10	0.81262E-01	0.36240E-06	0.52358E-06	0.22809E-01
12	0.14999E-10	0.81398E-01	0.18460E-05	0.22398E-05	0.22848E-01
13	0.19016E-10	0.81642E-01	0.94459E-05	0.96269E-05	0.22916E-01

14	0.23448E-10	0.81947E-01	0.48766E-04	0.41775E-04	0.23002E-01
15	0.28164E-10	0.83679E-01	0.25925E-03	0.18695E-03	0.23488E-01

**** LIQUID2 PROFILE ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.44790E-11	0.13686E-01	0.18152E-11	0.78885E-08	0.88222
2	0.44787E-11	0.13685E-01	0.18150E-11	0.78885E-08	0.88223
3	0.44783E-11	0.13684E-01	0.18149E-11	0.78885E-08	0.88223
4	0.44781E-11	0.13683E-01	0.18148E-11	0.78885E-08	0.88224
5	0.44778E-11	0.13683E-01	0.18147E-11	0.78885E-08	0.88224
6	0.44776E-11	0.13682E-01	0.18146E-11	0.78884E-08	0.88224
7	0.44774E-11	0.13681E-01	0.18145E-11	0.78882E-08	0.88224
8	0.44775E-11	0.13681E-01	0.18145E-11	0.78878E-08	0.88224
9	0.44786E-11	0.13681E-01	0.18148E-11	0.78867E-08	0.88223
10	0.44874E-11	0.13682E-01	0.18172E-11	0.78838E-08	0.88220
11	0.45643E-11	0.13685E-01	0.18457E-11	0.78762E-08	0.88214
12	0.49572E-11	0.13708E-01	0.20395E-11	0.78556E-08	0.88183
13	0.58393E-11	0.13749E-01	0.25831E-11	0.78026E-08	0.88117
14	0.69797E-11	0.13800E-01	0.34567E-11	0.76627E-08	0.87990
15	0.82137E-11	0.14092E-01	0.46025E-11	0.72777E-08	0.87500

**** LIQUID2 PROFILE ****

STAGE	K+	CL-	H3O+	HCL	DIGLY-1
1	0.12052E-08	0.12728E-08	0.79561E-08	0.30530E-05	0.27856E-06
2	0.43778E-08	0.46235E-08	0.81342E-08	0.59416E-05	0.54213E-06
3	0.12729E-07	0.13444E-07	0.86029E-08	0.86747E-05	0.79151E-06
4	0.34713E-07	0.36662E-07	0.98368E-08	0.11260E-04	0.10274E-05
5	0.92580E-07	0.97776E-07	0.13084E-07	0.13707E-04	0.12506E-05
6	0.24489E-06	0.25864E-06	0.21633E-07	0.16021E-04	0.14618E-05
7	0.64579E-06	0.68203E-06	0.44133E-07	0.18211E-04	0.16616E-05
8	0.17010E-05	0.17964E-05	0.10335E-06	0.20284E-04	0.18507E-05
9	0.44784E-05	0.47297E-05	0.25923E-06	0.22248E-04	0.20299E-05
10	0.11789E-04	0.12451E-04	0.66955E-06	0.24114E-04	0.22002E-05

1	0.12442E+08	0.12910E+08	0.51796	0.61675	0.11999E+08
2	36692.	0.12864E+08	0.51797	0.61678	0.11956E+08
3	2989.9	0.12819E+08	0.51798	0.61681	0.11916E+08
4	731.37	0.12776E+08	0.51800	0.61684	0.11877E+08
5	289.71	0.12734E+08	0.51801	0.61686	0.11838E+08
6	143.66	0.12688E+08	0.51802	0.61688	0.11797E+08
7	76.543	0.12631E+08	0.51802	0.61689	0.11744E+08
8	39.066	0.12538E+08	0.51800	0.61688	0.11659E+08
9	18.312	0.12350E+08	0.51794	0.61681	0.11488E+08
10	8.8243	0.11919E+08	0.51777	0.61659	0.11095E+08
11	5.2256	0.10902E+08	0.51730	0.61599	0.10166E+08
12	3.9421	0.86794E+07	0.51598	0.61426	0.81304E+07
13	3.4840	0.48856E+07	0.51260	0.60985	0.46262E+07
14	3.3125	0.12573E+07	0.50416	0.59890	0.12178E+07
15	3.1834	78875.	0.48149	0.56955	79095.

**** K-VALUES ****

STAGE	TRIGLY-2	FAME-3	TRIGLY-3	OH-	WATER
1	0.40272E+13	0.14794E+07	0.86539E+13	1.0000	0.88203
2	0.50627E+10	0.14765E+07	0.15808E+11	1.0000	0.88203
3	0.10112E+09	0.14737E+07	0.32686E+09	1.0000	0.88204
4	0.75459E+07	0.14711E+07	0.23044E+08	1.0000	0.88204
5	0.11296E+07	0.14685E+07	0.31729E+07	1.0000	0.88204
6	0.23843E+06	0.14657E+07	0.60867E+06	1.0000	0.88205
7	53603.	0.14621E+07	0.12195E+06	1.0000	0.88205
8	9732.6	0.14564E+07	18859.	1.0000	0.88205
9	1219.5	0.14448E+07	1862.8	1.0000	0.88205
10	138.13	0.14177E+07	156.04	1.0000	0.88203
11	25.566	0.13517E+07	21.999	1.0000	0.88199
12	9.7742	0.11960E+07	7.0942	1.0000	0.88184
13	6.2736	0.87237E+06	4.2039	1.0000	0.88148
14	5.1923	0.39741E+06	3.3892	1.0000	0.88069

15	4.6170	66237.	3.0578	1.0000	0.87832

**** K-VALUES ****

STAGE	K+	CL-	H3O+	HCL	DIGLY-1
1	1.0000	1.0000	1.0000	2.7816	2.7816
2	1.0000	1.0000	1.0000	2.7816	2.7816
3	1.0000	1.0000	1.0000	2.7815	2.7815
4	1.0000	1.0000	1.0000	2.7815	2.7815
5	1.0000	1.0000	1.0000	2.7814	2.7814
6	1.0000	1.0000	1.0000	2.7814	2.7814
7	1.0000	1.0000	1.0000	2.7812	2.7812
8	1.0000	1.0000	1.0000	2.7808	2.7808
9	1.0000	1.0000	1.0000	2.7800	2.7800
10	1.0000	1.0000	1.0000	2.7780	2.7780
11	1.0000	1.0000	1.0000	2.7728	2.7729
12	1.0000	1.0000	1.0000	2.7571	2.7573
13	1.0000	1.0000	1.0000	2.7191	2.7196
14	1.0000	1.0000	1.0000	2.6294	2.6317
15	1.0000	1.0000	1.0000	2.3852	2.3958

**** K-VALUES ****

STAGE	DIGLY-2	DIGLY-3	MONO-1	MONO-2	MONO-3
1	2.7816	2.7816	2.7816	2.7816	2.7816
2	2.7816	2.7816	2.7816	2.7816	2.7816
3	2.7815	2.7815	2.7815	2.7815	2.7815
4	2.7815	2.7815	2.7815	2.7815	2.7815
5	2.7814	2.7814	2.7814	2.7814	2.7814
6	2.7814	2.7814	2.7814	2.7814	2.7814
7	2.7812	2.7812	2.7812	2.7812	2.7812
8	2.7808	2.7808	2.7808	2.7808	2.7808
9	2.7800	2.7800	2.7800	2.7800	2.7800
10	2.7780	2.7780	2.7780	2.7780	2.7780
11	2.7729	2.7729	2.7729	2.7729	2.7729
12	2.7573	2.7573	2.7573	2.7573	2.7573
13	2.7196	2.7196	2.7196	2.7196	2.7196
14	2.6317	2.6317	2.6317	2.6317	2.6317
15	2.3958	2.3958	2.3958	2.3958	2.3958

```
BLOCK:  P-101    MODEL: PUMP
----------------------------
  INLET STREAM:           KOH
  OUTLET STREAM:          1
  PROPERTY OPTION SET:    ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
  HENRY-COMPS ID:         GLOBAL
  CHEMISTRY ID:           GLOBAL   - TRUE SPECIES
```

```
                    ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           104.152         104.152         0.00000
       MASS(LB/HR   )           2236.71         2236.71         0.00000
       ENTHALPY(BTU/HR  )     -0.120443E+08   -0.120429E+08
-0.116129E-03

                        ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                       90.0000
    DRIVER EFFICIENCY                                           1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                                30
    TOLERANCE                                                   0.000100000

                         ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                              24.8438
    PRESSURE CHANGE  PSI                                       90.0000
    NPSH AVAILABLE   FT-LBF/LB                                 23.2646
    FLUID POWER  HP                                             0.16261
    BRAKE POWER  HP                                             0.55001
    ELECTRICITY  KW                                             0.41014
    PUMP EFFICIENCY USED                                        0.29566
    NET WORK REQUIRED  HP                                       0.55001
    HEAD DEVELOPED FT-LBF/LB                                  143.951

 BLOCK:  P-102    MODEL: PUMP
 ----------------------------
   INLET STREAM:          METHANOL
   OUTLET STREAM:         3
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                    ***  MASS AND ENERGY BALANCE  ***
                                   IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           762.418         762.418         0.00000
       MASS(LB/HR   )           24429.5         24429.5         0.00000
       ENTHALPY(BTU/HR  )     -0.784648E+08   -0.784472E+08
```

```
-0.224625E-03

                          ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                    90.0000
    DRIVER EFFICIENCY                                        1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                             30
    TOLERANCE                                                0.000100000

                           ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                          498.521
    PRESSURE CHANGE  PSI                                    90.0000
    NPSH AVAILABLE   FT-LBF/LB                              36.3117
    FLUID POWER  HP                                          3.26305
    BRAKE POWER  HP                                          6.92695
    ELECTRICITY  KW                                          5.16543
    PUMP EFFICIENCY USED                                     0.47107
    NET WORK REQUIRED  HP                                    6.92695
    HEAD DEVELOPED FT-LBF/LB                               264.469

 BLOCK:  P-103    MODEL: HEATER
 ------------------------------
   INLET STREAM:          TRIGLY
   OUTLET STREAM:         10
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                      ***  MASS AND ENERGY BALANCE  ***
                                    IN               OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            255.573        255.573         0.00000
       MASS(LB/HR   )            223678.        223678.         0.00000
       ENTHALPY(BTU/HR  )      -0.195464E+09   -0.194508E+09
-0.488898E-02

                          ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE              F                     76.7300
   SPECIFIED PRESSURE                 PSIA                  94.7000
   MAXIMUM NO. ITERATIONS                                   50
   CONVERGENCE TOLERANCE
```

```
0.000100000


                          ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                    76.730
   OUTLET PRESSURE       PSIA                                 94.700
   HEAT DUTY             BTU/HR                              0.95562E+06



   PRESSURE-DROP CORRELATION PARAMETER                       -36757.



 BLOCK:  P-106    MODEL: PUMP
 ----------------------------
   INLET STREAM:          26
   OUTLET STREAM:         27
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN               OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           1076.84        1076.84         0.00000
       MASS(LB/HR   )           234851.        234851.
-0.123924E-15
       ENTHALPY(BTU/HR  )      -0.263933E+09   -0.263900E+09
-0.124784E-03

                         ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                    29.0000
    DRIVER EFFICIENCY                                        1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                             30
    TOLERANCE                                                0.000100000

                          ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                         4,430.35
    PRESSURE CHANGE  PSI                                    29.0000
```

```
    NPSH AVAILABLE   FT-LBF/LB                             112.492
    FLUID POWER  HP                                          9.34401
    BRAKE POWER  HP                                         12.9400
    ELECTRICITY  KW                                          9.64933
    PUMP EFFICIENCY USED                                     0.72210
    NET WORK REQUIRED  HP                                   12.9400
    HEAD DEVELOPED FT-LBF/LB                                78.7781

 BLOCK:  P-107    MODEL: PUMP
 ----------------------------
   INLET STREAM:          17
   OUTLET STREAM:         18
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN                OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           810.330         810.330
0.140297E-15
       MASS(LB/HR   )           40005.9         40005.9
0.545616E-15
       ENTHALPY(BTU/HR  )      -0.136408E+09   -0.136408E+09
-0.429970E-07

                         ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                     0.0
    DRIVER EFFICIENCY                                        1.00000

    FLASH SPECIFICATIONS:
    2 PHASE FLASH
    MAXIMUM NUMBER OF ITERATIONS                             50
    TOLERANCE                                                0.000100000

                          ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                          552.977
    PRESSURE CHANGE  PSI                                     0.0
    NPSH AVAILABLE   FT-LBF/LB                              89.5810
    FLUID POWER  HP                                          0.0
    BRAKE POWER  HP                                          0.0
    ELECTRICITY  KW                                          0.0
    PUMP EFFICIENCY USED                                     0.48557
    NET WORK REQUIRED  HP                                    0.0
    HEAD DEVELOPED FT-LBF/LB                                 0.0
```

```
 BLOCK:  P-109    MODEL: PUMP
 ----------------------------
   INLET STREAM:          20
   OUTLET STREAM:         21
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                    IN                OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           300.820         300.820
0.207857E-14
       MASS(LB/HR   )           23912.2         23912.2
-0.152139E-15
       ENTHALPY(BTU/HR  )      -0.748702E+08   -0.748642E+08
-0.803456E-04

                         ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                     39.0000
    DRIVER EFFICIENCY                                         1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                              30
    TOLERANCE                                                 0.000100000

                          ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                           349.541
    PRESSURE CHANGE  PSI                                     39.0000
    NPSH AVAILABLE   FT-LBF/LB                                0.68104
    FLUID POWER  HP                                           0.99142
    BRAKE POWER  HP                                           2.36358
    ELECTRICITY  KW                                           1.76252
    PUMP EFFICIENCY USED                                      0.41946
    NET WORK REQUIRED  HP                                     2.36358
    HEAD DEVELOPED FT-LBF/LB                                 82.0928

 BLOCK:  P-110    MODEL: PUMP
 ----------------------------
   INLET STREAM:          23
   OUTLET STREAM:         24
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
```

```
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                   ***  MASS AND ENERGY BALANCE  ***
                            IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          509.510         509.510         0.00000
       MASS(LB/HR   )          16093.7         16093.7         0.00000
       ENTHALPY(BTU/HR  )     -0.521428E+08   -0.521319E+08
-0.208527E-03

                       ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                  71.7000
    DRIVER EFFICIENCY                                      1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                           30
    TOLERANCE                                              0.000100000

                        ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                        340.345
    PRESSURE CHANGE  PSI                                  71.7000
    NPSH AVAILABLE   FT-LBF/LB                             0.0
    FLUID POWER  HP                                        1.77474
    BRAKE POWER  HP                                        4.27174
    ELECTRICITY  KW                                        3.18544
    PUMP EFFICIENCY USED                                   0.41546
    NET WORK REQUIRED  HP                                  4.27174
    HEAD DEVELOPED FT-LBF/LB                             218.345

 BLOCK:  P-111    MODEL: PUMP
 ----------------------------
   INLET STREAM:          32
   OUTLET STREAM:         33
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                   ***  MASS AND ENERGY BALANCE  ***
                            IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          262.939         262.939
```

```
0.735027E-14
       MASS(LB/HR   )            8420.00         8420.00
0.108016E-14
       ENTHALPY(BTU/HR  )      -0.267428E+08   -0.267353E+08
-0.282881E-03

                          ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                        71.7000
    DRIVER EFFICIENCY                                            1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                                 30
    TOLERANCE                                                    0.000100000

                           ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                              178.680
    PRESSURE CHANGE  PSI                                        71.7000
    NPSH AVAILABLE   FT-LBF/LB                                   0.0
    FLUID POWER  HP                                              0.93173
    BRAKE POWER  HP                                              2.97049
    ELECTRICITY  KW                                              2.21509
    PUMP EFFICIENCY USED                                         0.31366
    NET WORK REQUIRED  HP                                        2.97049
    HEAD DEVELOPED FT-LBF/LB                                   219.101

 BLOCK:  P-112    MODEL: PUMP
 ----------------------------
   INLET STREAM:          36
   OUTLET STREAM:         37
   PROPERTY OPTION SET:   ELECNRTL  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                       ***  MASS AND ENERGY BALANCE  ***
                                  IN                OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)            813.903         813.903
-0.139681E-15
       MASS(LB/HR   )            226431.         226431.
0.128533E-15
       ENTHALPY(BTU/HR  )      -0.198761E+09   -0.198691E+09
-0.350873E-03
```

```
                         ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                        56.8500
    DRIVER EFFICIENCY                                            1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                                 30
    TOLERANCE                                                    0.000100000

                          ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                            4,837.54
    PRESSURE CHANGE  PSI                                        56.8500
    NPSH AVAILABLE   FT-LBF/LB                                   0.0
    FLUID POWER  HP                                             20.0010
    BRAKE POWER  HP                                             27.4057
    ELECTRICITY  KW                                             20.4364
    PUMP EFFICIENCY USED                                         0.72981
    NET WORK REQUIRED  HP                                       27.4057
    HEAD DEVELOPED FT-LBF/LB                                   174.896

 BLOCK:  P-114    MODEL: HEATER
 ------------------------------
   INLET STREAM:          HCL
   OUTLET STREAM:         41
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)          78.1237         78.1237
0.181902E-15
       MASS(LB/HR   )          1754.70         1754.70         0.00000
       ENTHALPY(BTU/HR  )    -0.854322E+07   -0.854354E+07
0.374776E-04

                         ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE              F                         76.7300
   SPECIFIED PRESSURE                 PSIA                      62.7000
   MAXIMUM NO. ITERATIONS                                       50
   CONVERGENCE TOLERANCE
0.000100000
```

```
                        ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                       76.730
   OUTLET PRESSURE       PSIA                                    62.700
   HEAT DUTY             BTU/HR                                 -320.19


   PRESSURE-DROP CORRELATION PARAMETER                     -0.76990E+10


 BLOCK:  P-115    MODEL: PUMP
 ----------------------------
   INLET STREAM:          PROWATER
   OUTLET STREAM:         45
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                   ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           11404.6         11404.6
-0.159496E-15
       MASS(LB/HR   )           205458.         205458.         0.00000
       ENTHALPY(BTU/HR  )      -0.140238E+10   -0.140234E+10
-0.281995E-04

                       ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                     45.0000
    DRIVER EFFICIENCY                                         1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                              30
    TOLERANCE                                                 0.000100000

                        ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                          3,300.33
    PRESSURE CHANGE  PSI                                     45.0000
    NPSH AVAILABLE   FT-LBF/LB                               32.9418
```

```
    FLUID POWER  HP                                            10.8011
    BRAKE POWER  HP                                            15.5425
    ELECTRICITY  KW                                            11.5900
    PUMP EFFICIENCY USED                                        0.69494
    NET WORK REQUIRED  HP                                      15.5425
    HEAD DEVELOPED FT-LBF/LB                                  104.090

 BLOCK:  P-117    MODEL: PUMP
 ----------------------------
   INLET STREAM:          49
   OUTLET STREAM:         50
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN               OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           6508.89        6508.89
0.139731E-15
       MASS(LB/HR   )           328115.        328115.
0.177400E-15
       ENTHALPY(BTU/HR  )      -0.918205E+09   -0.918122E+09
-0.898763E-04

                         ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                       56.0000
    DRIVER EFFICIENCY                                           1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                                30
    TOLERANCE                                                   0.000100000

                          ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                            5,949.81
    PRESSURE CHANGE  PSI                                       56.0000
    NPSH AVAILABLE   FT-LBF/LB                                 24.2612
    FLUID POWER  HP                                            24.2319
    BRAKE POWER  HP                                            32.4293
    ELECTRICITY  KW                                            24.1825
    PUMP EFFICIENCY USED                                        0.74722
    NET WORK REQUIRED  HP                                      32.4293
    HEAD DEVELOPED FT-LBF/LB                                  146.227
```

```
 BLOCK:  P-118    MODEL: PUMP
 ----------------------------
   INLET STREAM:          63
   OUTLET STREAM:         64
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN              OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           11521.9         11521.9
-0.157872E-15
       MASS(LB/HR   )           208830.         208830.
0.139366E-15
       ENTHALPY(BTU/HR  )      -0.139827E+10   -0.139823E+10
-0.260889E-04

                         ***  INPUT DATA  ***
    PRESSURE CHANGE  PSI                                       40.3000
    DRIVER EFFICIENCY                                           1.00000

    FLASH SPECIFICATIONS:
    LIQUID PHASE CALCULATION
    NO FLASH PERFORMED
    MAXIMUM NUMBER OF ITERATIONS                                30
    TOLERANCE                                                   0.000100000

                          ***  RESULTS  ***
    VOLUMETRIC FLOW RATE  CUFT/HR                           3,415.36
    PRESSURE CHANGE  PSI                                       40.3000
    NPSH AVAILABLE   FT-LBF/LB                                  8.17227
    FLUID POWER  HP                                            10.0101
    BRAKE POWER  HP                                            14.3369
    ELECTRICITY  KW                                            10.6910
    PUMP EFFICIENCY USED                                        0.69821
    NET WORK REQUIRED  HP                                      14.3369
    HEAD DEVELOPED FT-LBF/LB                                   94.9096

 BLOCK:  P-119    MODEL: HEATER
 ------------------------------
   INLET STREAM:          59
   OUTLET STREAM:         60
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
```

```
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                                  IN               OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           5735.28         5735.28
0.158579E-15
       MASS(LB/HR   )           103323.         103323.
0.328156E-13
       ENTHALPY(BTU/HR  )      -0.715842E+09   -0.707660E+09
-0.114297E-01

                         ***  INPUT DATA  ***
   TWO    PHASE  TP  FLASH
   SPECIFIED TEMPERATURE              F                        53.3300
   SPECIFIED PRESSURE                 PSIA                     71.7000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.000100000



                          ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                  53.330
   OUTLET PRESSURE       PSIA                               71.700
   HEAT DUTY             BTU/HR                            0.81819E+07
   OUTLET VAPOR FRACTION                                    0.0000
   PRESSURE-DROP CORRELATION PARAMETER                     -0.23246E+07



   V-L PHASE EQUILIBRIUM :

      COMP               F(I)           X(I)           Y(I)           K(I)
      TRIGLY-1         0.86905E-14    0.10000E-14    0.98949
0.86878E+14
      GLYCEROL         0.28767E-13    0.28767E-13    0.15776E-20
0.48150E-08
      METHANOL         0.20546E-12    0.20546E-12    0.28090E-13
0.12004E-01
      OH-               0.0000        0.10540E-08     0.0000         0.0000
      WATER             1.0000         1.0000        0.10515E-01
0.92323E-03
      K+               0.39583E-14    0.39583E-14     0.0000         0.0000
```

```
      CL-              0.10032E-10    0.10032E-10     0.0000         0.0000
      H3O+             0.10028E-10    0.10640E-08     0.0000         0.0000
      HCL               0.0000        0.10717E-23    0.29271E-15
0.23981E+08
      DIGLY-1          0.39974E-12    0.39974E-12    0.34354E-18
0.75457E-07
      DIGLY-2          0.90854E-13    0.90854E-13    0.86011E-20
0.83121E-08

 BLOCK:  P-120    MODEL: HEATER
 ------------------------------
   INLET STREAM:          56
   OUTLET STREAM:         57
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                      ***  MASS AND ENERGY BALANCE  ***
                                IN               OUT        RELATIVE
DIFF.
    TOTAL BALANCE
       MOLE(LBMOL/HR)           773.605        773.605
-0.293914E-15
       MASS(LB/HR   )           224793.        224793.
0.129470E-15
       ENTHALPY(BTU/HR  )      -0.244136E+09   -0.227374E+09
-0.686568E-01

                          ***  INPUT DATA  ***
   ONE    PHASE  TP  FLASH   SPECIFIED PHASE IS  LIQUID
   SPECIFIED TEMPERATURE              F                        53.3300
   SPECIFIED PRESSURE                 PSIA                     46.7000
   MAXIMUM NO. ITERATIONS                                      30
   CONVERGENCE TOLERANCE
0.000100000



                           ***  RESULTS  ***
   OUTLET TEMPERATURE    F                                  53.330
   OUTLET PRESSURE       PSIA                               46.700
   HEAT DUTY             BTU/HR                            0.16762E+08


   PRESSURE-DROP CORRELATION PARAMETER                     -0.24312E+06
```

```
 BLOCK:  R-101    MODEL: RSTOIC
 ------------------------------
   INLET STREAMS:         12          14          9           8
   OUTLET STREAM:         15
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                     ***  MASS AND ENERGY BALANCE  ***
                            IN          OUT       GENERATION   RELATIVE
DIFF.
   TOTAL BALANCE
   MOLE(LBMOL/HR)         1887.17       1887.17      0.261590E-12   0.00000
   MASS(LB/HR   )         274857.       274857.
0.211774E-15
   ENTHALPY(BTU/HR  )   -0.370359E+09 -0.377740E+09
0.195387E-01

                         ***  INPUT DATA  ***
   STOICHIOMETRY MATRIX:

    REACTION #   1:
     SUBSTREAM MIXED   :
     TRIGLY-1  -1.00    FAME-1     1.00    METHANOL  -1.00    DIGLY-1
1.00

    REACTION #   2:
     SUBSTREAM MIXED   :
     METHANOL  -1.00    FAME-2     1.00    TRIGLY-2  -1.00    DIGLY-2
1.00

    REACTION #   3:
     SUBSTREAM MIXED   :
     METHANOL  -1.00    FAME-3     1.00    TRIGLY-3  -1.00    DIGLY-3
1.00

    REACTION #   4:
     SUBSTREAM MIXED   :
     FAME-1     1.00    METHANOL  -1.00    DIGLY-1   -1.00    MONO-1
1.00

    REACTION #   5:
     SUBSTREAM MIXED   :
```

```
     METHANOL  -1.00    FAME-2     1.00    DIGLY-2   -1.00    MONO-2
1.00

    REACTION #   6:
     SUBSTREAM MIXED   :
     METHANOL  -1.00    FAME-3     1.00    DIGLY-3   -1.00    MONO-3
1.00

    REACTION #   7:
     SUBSTREAM MIXED   :
     FAME-1     1.00    GLYCEROL   1.00    METHANOL  -1.00    MONO-1
-1.00

    REACTION #   8:
     SUBSTREAM MIXED   :
     GLYCEROL   1.00    METHANOL  -1.00    FAME-2     1.00    MONO-2
-1.00

    REACTION #   9:
     SUBSTREAM MIXED   :
     GLYCEROL   1.00    METHANOL  -1.00    FAME-3     1.00    MONO-3
-1.00


   REACTION CONVERSION SPECS: NUMBER=    9
     REACTION #   1:
     SUBSTREAM:MIXED    KEY COMP:TRIGLY-1 CONV FRAC:  1.000
     REACTION #   2:
     SUBSTREAM:MIXED    KEY COMP:TRIGLY-2 CONV FRAC:  1.000
     REACTION #   3:
     SUBSTREAM:MIXED    KEY COMP:TRIGLY-3 CONV FRAC:  1.000
     REACTION #   4:
     SUBSTREAM:MIXED    KEY COMP:DIGLY-1  CONV FRAC: 0.9901
     REACTION #   5:
     SUBSTREAM:MIXED    KEY COMP:DIGLY-2  CONV FRAC: 0.9901
     REACTION #   6:
     SUBSTREAM:MIXED    KEY COMP:DIGLY-3  CONV FRAC: 0.9901
     REACTION #   7:
     SUBSTREAM:MIXED    KEY COMP:MONO-1   CONV FRAC: 0.9852
     REACTION #   8:
     SUBSTREAM:MIXED    KEY COMP:MONO-2   CONV FRAC: 0.9852
     REACTION #   9:
     SUBSTREAM:MIXED    KEY COMP:MONO-3   CONV FRAC: 0.9852


   HEAT OF REACTION SPECIFICATIONS:
```

REACTION NUMBER	REFERENCE COMPONENT	HEAT OF REACTION BTU/LBMOL
1	TRIGLY-1	-6402.7
2	TRIGLY-2	-24143.
3	TRIGLY-3	-6385.6
4	DIGLY-1	-6272.3
5	DIGLY-2	-6270.0
6	DIGLY-3	-6226.8
7	MONO-1	-13014.
8	MONO-2	-13010.
9	MONO-3	-12753.

ONE PHASE TP FLASH SPECIFIED PHASE IS LIQUID		
SPECIFIED TEMPERATURE	F	180.000
PRESSURE DROP	PSI	5.00000
MAXIMUM NO. ITERATIONS		30
CONVERGENCE TOLERANCE		0.000100000
SERIES REACTIONS		
GENERATE COMBUSTION REACTIONS FOR FEED SPECIES		NO

*** RESULTS ***

OUTLET TEMPERATURE	F	180.00
OUTLET PRESSURE	PSIA	52.700
HEAT DUTY	BTU/HR	-0.73805E+07

HEAT OF REACTIONS:

REACTION NUMBER	REFERENCE COMPONENT	HEAT OF REACTION BTU/LBMOL
1	TRIGLY-1	-6402.7
2	TRIGLY-2	-24143.
3	TRIGLY-3	-6385.6
4	DIGLY-1	-6272.3
5	DIGLY-2	-6270.0
6	DIGLY-3	-6226.8
7	MONO-1	-13014.
8	MONO-2	-13010.
9	MONO-3	-12753.

REACTION EXTENTS:

```
      REACTION          REACTION
      NUMBER            EXTENT
                        LBMOL/HR
      1                  176.36
      2                  49.503
      3                  29.700
      4                  174.61
      5                  49.012
      6                  29.405
      7                  172.02
      8                  48.285
      9                  28.969

 BLOCK:  R-102    MODEL: RSTOIC
 ------------------------------
   INLET STREAM:          29
   OUTLET STREAM:         30
   PROPERTY OPTION SET:   ENRTL-RK  ELECTROLYTE NRTL / REDLICH-KWONG
   HENRY-COMPS ID:        GLOBAL
   CHEMISTRY ID:          GLOBAL   - TRUE SPECIES

                      ***  MASS AND ENERGY BALANCE  ***
                             IN           OUT       GENERATION   RELATIVE
DIFF.
   TOTAL BALANCE
   MOLE(LBMOL/HR)         1076.84       1076.84      0.136280E-12
0.211148E-15
   MASS(LB/HR   )         234851.       234851.                     0.00000
   ENTHALPY(BTU/HR  )   -0.252573E+09 -0.252573E+09
-0.428316E-07

                          ***  INPUT DATA  ***
   STOICHIOMETRY MATRIX:

    REACTION #   1:
     SUBSTREAM MIXED   :
     TRIGLY-1  -1.00    FAME-1     1.00    METHANOL  -1.00    DIGLY-1
1.00

    REACTION #   2:
     SUBSTREAM MIXED   :
     METHANOL  -1.00    FAME-2     1.00    TRIGLY-2  -1.00    DIGLY-2
1.00

    REACTION #   3:
```

```
     SUBSTREAM MIXED   :
     METHANOL  -1.00    FAME-3     1.00    TRIGLY-3  -1.00    DIGLY-3
1.00

    REACTION #   4:
     SUBSTREAM MIXED   :
     FAME-1     1.00    METHANOL  -1.00    DIGLY-1   -1.00    MONO-1
1.00

    REACTION #   5:
     SUBSTREAM MIXED   :
     METHANOL  -1.00    FAME-2     1.00    DIGLY-2   -1.00    MONO-2
1.00

    REACTION #   6:
     SUBSTREAM MIXED   :
     METHANOL  -1.00    FAME-3     1.00    DIGLY-3   -1.00    MONO-3
1.00

    REACTION #   7:
     SUBSTREAM MIXED   :
     FAME-1     1.00    GLYCEROL   1.00    METHANOL  -1.00    MONO-1
-1.00

    REACTION #   8:
     SUBSTREAM MIXED   :
     GLYCEROL   1.00    METHANOL  -1.00    FAME-2     1.00    MONO-2
-1.00

    REACTION #   9:
     SUBSTREAM MIXED   :
     GLYCEROL   1.00    METHANOL  -1.00    FAME-3     1.00    MONO-3
-1.00


   REACTION CONVERSION SPECS: NUMBER=    9
     REACTION #   1:
     SUBSTREAM:MIXED    KEY COMP:TRIGLY-1 CONV FRAC:  1.000
     REACTION #   2:
     SUBSTREAM:MIXED    KEY COMP:TRIGLY-2 CONV FRAC:  1.000
     REACTION #   3:
     SUBSTREAM:MIXED    KEY COMP:TRIGLY-3 CONV FRAC:  1.000
     REACTION #   4:
     SUBSTREAM:MIXED    KEY COMP:DIGLY-1  CONV FRAC: 0.9901
     REACTION #   5:
     SUBSTREAM:MIXED    KEY COMP:DIGLY-2  CONV FRAC: 0.9901
```

```
  REACTION #   6:
  SUBSTREAM:MIXED    KEY COMP:DIGLY-3  CONV FRAC: 0.9901
  REACTION #   7:
  SUBSTREAM:MIXED    KEY COMP:MONO-1   CONV FRAC: 0.9852
  REACTION #   8:
  SUBSTREAM:MIXED    KEY COMP:MONO-2   CONV FRAC: 0.9852
  REACTION #   9:
  SUBSTREAM:MIXED    KEY COMP:MONO-3   CONV FRAC: 0.9852


HEAT OF REACTION SPECIFICATIONS:

  REACTION           REFERENCE          HEAT OF
  NUMBER             COMPONENT          REACTION
                                         BTU/LBMOL
      1                 TRIGLY-1          -6402.7
      2                 TRIGLY-2          -24143.
      3                 TRIGLY-3          -6385.6
      4                 DIGLY-1           -6272.3
      5                 DIGLY-2           -6270.0
      6                 DIGLY-3           -6226.8
      7                 MONO-1            -13014.
      8                 MONO-2            -13010.
      9                 MONO-3            -12753.


ONE    PHASE  PQ  FLASH   SPECIFIED PHASE IS  LIQUID
PRESSURE DROP         PSI                                 5.00000
SPECIFIED HEAT DUTY   BTU/HR                              0.0
MAXIMUM NO. ITERATIONS                                   30
CONVERGENCE TOLERANCE                                     0.000100000
SERIES REACTIONS
GENERATE COMBUSTION REACTIONS FOR FEED SPECIES          NO

                        ***  RESULTS  ***
OUTLET TEMPERATURE    F                                    143.73
OUTLET PRESSURE       PSIA                                 34.700



HEAT OF REACTIONS:

  REACTION           REFERENCE          HEAT OF
  NUMBER             COMPONENT          REACTION
                                         BTU/LBMOL
   1                    TRIGLY-1          -6402.7
```

2	TRIGLY-2	-24143.
3	TRIGLY-3	-6385.6
4	DIGLY-1	-6272.3
5	DIGLY-2	-6270.0
6	DIGLY-3	-6226.8
7	MONO-1	-13014.
8	MONO-2	-13010.
9	MONO-3	-12753.

REACTION EXTENTS:

REACTION NUMBER	REACTION EXTENT LBMOL/HR
1	0.59961E-02
2	0.16831E-02
3	0.10098E-02
4	1.7376
5	0.48762
6	0.29270
7	4.2626
8	1.1963
9	0.71790

E. Equipment Design Calculations

Distillation Tower
Applicable Blocks: D-101, D-102

Variable	Description	Unit
L	Liquid Mass Flow Rate	kg/hr
G	Vapor Mass Flow Rate	kg/hr
ρL	Liquid Density	kg/m^3
ρG	Vapor Density	kg/m^3
σ	Surface Tension	dyne/cm
P_o	Minimum Operating Pressure	PSI
ρS	Density of Carbon Steel	lb/in^3

Relevant Equations: **Source**

1) Compute the Vapor Flooding Velocity (Uf)

F_{LG}	=	$(L/G)*(\rho L)*(\rho G)$	SSLW (S19.6)
C_{SB}	=	Find from Fair Correlation Plot with F_{LG} and plate spacing	SSLW (S19.6)
F_{ST}	=	$(\sigma/20)^{0.20}$	SSLW (S19.6)
F_F	=	0.9 (Foaming Factor)	SSLW (S19.6)
F_{HA}	=	1 ($Ah/Aa \geq 0.10$)	SSLW (S19.6)
C	=	$C_{SB}*F_{ST}*F_F*F_{HA}$	SSLW (19.13)
Uf	=	$C*((\rho L-\rho G)/\rho G)^{0.5}$	SSLW (19.12)

2) Calculate the Tower Diameter (D_T)

A_d/A_T	=	0.1	$F_{LG} \leq 0.1$	SSLW (S19.6)
		$0.1 + (F_{LG}-0.1)/9$	$0.1 \leq F_{LG} \leq 1.0$	
		0.2	$F_{LG} \geq 1.0$	
f	=	0.85 (Fraction of Vapor Flooding Velocity)		SSLW (S19.6)
D_T	=	$((4*G)/(f*Uf*\pi*\rho G*(1-(A_d/A_T))$		SSLW (19.11)

3) Find the Tower Height (H_T)

N_T	=	Number of Trays
S_T	=	Tray Spacing
L_B	=	Bottoms Volumetric Flow Rate
τ_B	=	Sump Residence Time
H_D	=	Disengaging Height
F_H	=	Extra Feed Tray Height
M_H	=	Extra Manhole Height (1/10 trays)
H_T	=	$(N_T*S_T)+(L_B*\tau_B)/(\pi*0.25*(D_T)^2)+M_H+F_H$

4) Determine the Shell Thickess (Vacuum Tower)

P_d	=	$14.7\text{-}P_o$ (Maximum Pressure Difference)	SSLW (E22.14)
E_m	=	Modulus of Elasticity	SSLW(S22.5)
t_E	=	$1.3*D_T*(P_d*H_T/E_m*D_T)^{0.40}$	SSLW(22.63)
t_{EC}	=	$H_T*(0.18*D_T-2.2)*(10^{-5})-0.19$	SSLW(22.64)
t_C	=	0.125 (in) Corrosion Allowance	SSLW(S22.5)

t_v = $t_E+t_{EC}+t_C$

5) Find the Vessel Weight and Cost

W	=	$\pi*(D_T+t_V)*(H_T+0.8*D_T)*t_V*\rho S$	SSLW(22.59)
C_V	=	$\exp(7.2756+0.18255*(\ln(W))+0.02297*(\ln(W))^2)$	SSLW(22.57)
F_M	=	1 (Carbon Steel)	
C_{VESSEL}	=	C_V*F_M	

6) Find the Cost of Platforms and Ladders and Trays

C_{PL} = $300.9*(D_T)^{0.63316}(H_T)^{0.80161}$ SSLW(22.58)

F_{NT}	=	$2.25/(1.0414^{NT})$	SSLW(22.68)
F_{TT}	=	1 (Sieve Tray)	SSLW(S22.5)
F_{TM}	=	1 (Carbon Steel)	SSLW(S22.5)
C_{BT}	=	$468*\exp(0.1739*D_T)$	SSLW(22.67)

C_T = $N_T*F_{NT}*F_{TT}*F_{TM}*C_{BT}$ SSLW(22.66)

7) Find the Total Tower Cost

C_{FOB}	=	$C_{VESSEL}+C_{PL}+C_T$	
F_{BM}	=	4.16	SSLW(T22.11)

C_P = $C_{FOB}*F_{BM}$

Centrifugal Pump
Applicable Blocks: (MODULE 1): P-101, P-102, P-103, P-104, P-105, P-106, P-107, P-108 (MODULE 2): P-101, P-102, P-103, P-104 ((MODULE 3): P-101, P-102, P-103, P-104, P-105, P-106, P-107, P-108, P-109, P-110, P-111, P-112, P-113, P-114, P-115, P 116, P-117, P-118, P-119, P-120, P-121, P-122

Variable	Description	Unit
Q	Volumetric Flow Rate	Gal/min
ΔP	Pressure Rise	PSI
ρ_L	Liquid Density	lb/gal
A_G	Gravitational Accelera	ft/s^2

Relevant Equations: **Source**

1) Find Pump Head

P_H = $(\Delta P*144)/(\rho_L*A_G)$

2) Find Pump Cost

				Source
S	=	$Q*P_H^{0.5}$		SSLW (22.13)
C_B	=	$\exp(9.7171-0.6019*\ln(S)+0.0519*\ln(S)^2)$		SSLW (22.14)
F_M	=	1 (Cast Iron)		SSLW (S22.5)
F_T	=	1.0	$50<P_H<400$, Hp_{max} = 75	SSLW (S22.5)
		1.5	$50<P_H<200$, Hp_{max} = 200	
		1.7	$100<P_H<450$, Hp_{max} = 150	
		2.0	$50<P_H<500$, Hp_{max} = 250	
		2.7	$300<P_H<1{,}100$, Hp_{max} = 250	
		8.9	$650<P_H<3{,}200$, Hp_{max} = 1,450	

C_{PUMP} = $C_B*F_M*F_T$

3) Find Motor Cost

			Source
η_P	=	0.7	
P_B	=	$(Q*P_H*\rho_L)/(33{,}000*\eta_P)$	SSLW(22.16)
η_M	=	$0.80+0.0319*\ln(P_B)-0.00182*\ln(P_B)^2$	SSLW(22.18)
P_C	=	$(Q*P_H*\rho_L)/(33{,}000*\eta_P*\eta_M)$	SSLW(22.16)
C_B	=	$\exp(5.8259+0.13141*\ln(P_C)+0.053255*\ln(P_C)^2+0.028628*\ln(P_C)^3-0.0035549*\ln(P_C)^4$	SSLW(22.19)
F_T	=	1.8 (Explosion-proof enclosure)	SSLW(S22.5)

C_{MOTOR} = C_B*F_T

4) Find Total Cost

C_{FOB} = $C_{PUMP}+C_{MOTOR}$

F_{BM} = 3.3 SSLW(T22.11)

C_P = $C_{FOB}*F_{BM}$

Centrifugal Blower

Applicable Blocks: B-101, B-102

Variable	Description	Unit
Q	Volumetric Flow Rate	Gal/min
P_I	Inlet Pressure	PSI
P_O	Outlet Pressure	PSI

Relevant Equations: **Source**

1) Find Consumed Power

η_B	=	0.75	
P_B	=	$0.00436*(k/(k-1)*(Q*P_I/\eta_B)*((P_O/P_I)^{(k-1)/k}-1)$	SSLW (22.30)
η_M	=	$0.80+0.0319*\ln(P_B)-0.00182*\ln(P_B)^2$	SSLW (22.18)
P_C	=	P_B/η_M	

2) Find Total Blower Cost

C_B	=	$\exp(6.8929+0.7900*\ln(P_C))$	SSLW (22.32)
F_M	=	1 (Cast Iron)	
C_{FOB}	=	C_B*F_M	
F_{BM}	=	2.15	SSLW(T22.11)
C_P	=	$C_{FOB}*F_{BM}$	

Shell and Tube Heat Exchanger (Fixed Head)

Applicable Blocks: H-101, H-103, H-104, H-105, H-106, H-108, C-101, C-102, C-103, C-104, C-105, C-106, C-107

Variable	Description	Unit
$T_{H, in}$	Initial Temperature of the Hot Stream	F
$T_{H, out}$	Final Temperature of the Hot Stream	F
$T_{C, in}$	Initial Temperature of the Cold Stream	F
$T_{C, out}$	Final Temperature of the Cold Stream	F
Q	Heat Duty	BTU
U	Overall Heat Exchange Coefficient	BTU/ft^2 hr F
L	Tube Length	ft
P_S	Shell Pressure	PSIG
a	Material of Construction Factor	Dimensionless
b	Material of Construction Factor	Dimensionless

Relevant Equations:

1) Find the Log-Mean Temperature Difference

			Source
ΔT_1	=	$T_{H, in} - T_{C, out}$	SSLW(S18.3)
ΔT_2	=	$T_{H, out} - T_{C, in}$	SSLW(S18.3)
ΔT_{LM}	=	$(\Delta T_1 - \Delta T_2)/\ln(\Delta T_1/\Delta T_2)$	SSLW(18.3)

2) Find the Area Required for Heat Exchange

A	=	$Q/(U*\Delta T_{LM})$	SSLW(S18.3)

3) Find Vessel Cost:

F_M	=	$a+(A/100)^b$	SSLW(22.44)
F_L	=	1 (L = 20 ft)	SSLW(S22.5)
F_P	=	$0.9803 + 0.018(P_S/100) + 0.0017(P_S/100)^2$	SSLW(22.45)
C_B	=	$\exp(11.0545-0.9228*\ln(A)+0.09861*\ln(A)^2)$	SSLW(22.40)
C_{FOB}	=	$F_M*F_L*F_P*C_B$	
F_{BM}	=	3.17	SSLW(T22.11)
C_P	=	$C_{FOB}*F_{BM}$	

Shell and Tube Heat Exchanger (Kettle Vaporizer)

Applicable Blocks: H-107, H-109

Variable	Description	Unit
$T_{H, in}$	Initial Temperature of the Hot Stream	F
$T_{H, out}$	Final Temperature of the Hot Stream	F
$T_{C, in}$	Initial Temperature of the Cold Stream	F
$T_{C, out}$	Final Temperature of the Cold Stream	F
Q	Heat Duty	BTU
U	Overall Heat Exchange Coefficient	BTU/ft^2 hr F
L	Tube Length	ft
P_S	Shell Pressure	PSIG
a	Material of Construction Factor	Dimensionless
b	Material of Construction Factor	Dimensionless

Relevant Equations:

1) Find the Log-Mean Temperature Difference

			Source
ΔT_1	=	$T_{H, in} - T_{C, out}$	SSLW(S18.3)
ΔT_2	=	$T_{H, out} - T_{C, in}$	SSLW(S18.3)
ΔT_{LM}	=	$(\Delta T_1 - \Delta T_2)/\ln(\Delta T_1/\Delta T_2)$	SSLW(18.3)

2) Find the Area Required for Heat Exchange

A	=	$Q/(U*\Delta T_{LM})$	SSLW(S18.3)

3) Find Vessel Cost:

F_M	=	$a+(A/100)^b$	SSLW(22.44)
F_L	=	1 (L = 20 ft)	SSLW(S22.5)
F_P	=	$0.9803 + 0.018(P_S/100) + 0.0017(P_S/100)^2$	SSLW(22.45)
C_B	=	$\exp(11.967-0.8709*\ln(A)+0.09005*\ln(A)^2)$	SSLW(22.42)
C_{FOB}	=	$F_M*F_L*F_P*C_B$	
F_{BM}	=	3.17	SSLW(T22.11)
C_P	=	$C_{FOB}*F_{BM}$	

Double-Pipe Heat Exchanger

Applicable Blocks: H-102

Variable	Description	Unit
$T_{H, in}$	Initial Temperature of the Hot Stream	F
$T_{H, out}$	Final Temperature of the Hot Stream	F
$T_{C, in}$	Initial Temperature of the Cold Stream	F
$T_{C, out}$	Final Temperature of the Cold Stream	F
Q	Heat Duty	BTU
U	Overall Heat Exchange Coefficient	BTU/ft^2 hr F
L	Tube Length	ft
P_S	Shell Pressure	PSIG
a	Material of Construction Factor	Dimensionless
b	Material of Construction Factor	Dimensionless

Relevant Equations:

1) Find the Log-Mean Temperature Difference

			Source
ΔT_1	=	$T_{H, in} - T_{C, out}$	SSLW(S18.3)
ΔT_2	=	$T_{H, out} - T_{C, in}$	SSLW(S18.3)
ΔT_{LM}	=	$(\Delta T_1 - \Delta T_2)/\ln(\Delta T_1/\Delta T_2)$	SSLW(18.3)

2) Find the Area Required for Heat Exchange

A	=	$Q/(U^*\Delta T_{LM})$	SSLW(S18.3)

3) Find Vessel Cost:

F_M	=	2 (Carbon Steel Outer Pipe, Stainless Steel Inner Pipe)	SSLW(S22.5)
F_P	=	$0.8510+0.1292(P/600)+0.0198^*(P/600)^2$	SSLW(22.48)
C_B	=	$\exp(7.1460+0.16^*\ln(A))$	SSLW(22.46)
C_{FOB}	=	$F_M^*F_P^*C_B$	SSLW(22.47)
F_{BM}	=	3.17	SSLW(T22.11)
C_P	=	$C_{FOB}^*F_{BM}$	

Floating-Roof Storage Tank

Applicable Blocks: T-101, T-102, T-103, T-104, T-105, T-106, T-109, T-110

Variable	Description	Unit
Q	Inlet/Outlet Volumetric Flow Rate	gal/hr
τ	Residence Time	hr

Relevant Equations:

1) Find the Tank Volume

			Source
V	=	$Q*\tau$	

2) Calculate Vessel Cost

C_{FOB}	=	$475*V^{0.51}$	SSLW(T22.32)
F_{BM}	=	3.05	SSLW(T22.11)
C_P	=	$C_{FOB}*F_{BM}$	

Cone-Roof Storage Tank

Applicable Blocks: T-107, T-108

Variable	Description	Unit
Q	Inlet/Outlet Volumetric Flow Rate	gal/hr
τ	Residence Time	hr

Relevant Equations:

1) Find the Tank Volume

		Source
V	$= Q*\tau$	

2) Calculate Vessel Cost

C_{FOB}	$= 265*V^{0.51}$	SSLW(T22.32)
F_{BM}	$= 3.05$	SSLW(T22.11)
C_P	$= C_{FOB}*F_{BM}$	

Horizontal Pressure Vessel

Applicable Blocks: S-101, S-102, S-103, S-104

Variable	Description	Unit
Q	Volumetric Flow Rate	ft^3/min
τ	Residence Time	min
AR	Aspect Ratio	Dimensionless
P_O	Operating Pressure	PSIG
E	Weld Efficiency	Dimensionless
S	Maximum Allowable Shell Stress	PSI
ρS	Density of Carbon Steel	lb/in^3

Relevant Equations: **Source**

1) Find Diameter and Length:

$V = Q*\tau$

$D = ((4*V)/(AR*\pi))^{1/3}$

$L = D*AR$

2) Shell Thickness

$P_d = \exp(0.60608+0.91615*\ln(P_O)+0.0015655*\ln(P_O)^2$ SSLW (22.61)

$t_P = (P_d*D)/(2*S*E-1.2*P_d)$ SSLW(22.60)

t_C = 0.125 (in) Corrosion Allowance

$t_s = t_P+t_C$

3) Find Vessel Weight and Cost

$W = \pi*(D+t_S)*(L+0.8*D)*t_V*\rho S$ SSLW(22.59)

$C_V = \exp(8.9552-0.2330*\ln(W)+0.04333*\ln(W)^2)$ SSLW(22.53)

F_M = 1 (Carbon Steel)

$C_{VESSEL} = C_V*F_M$

4) Find Cost of Platforms and Ladders

$C_{PL} = 2005*D^{0.20294}$ SSLW(22.55)

5) Find Total Horizontal Pressure Vessel Cost

$C_{FOB} = C_{VESSEL} + C_{PL}$

$F_{BM} = 3.05$ SSLW(T22.11)

$C_P = C_{FOB}*F_{BM}$

Vertical Pressure Vessel
Applicable Blocks: R-102

Variable	Description	Unit
Q	Volumetric Flow Rate	ft^3/min
τ	Reaction Time	min
AR	Aspect Ratio	Dimensionless
P_O	Operating Pressure	PSIG
E	Weld Efficiency	Dimensionless
S	Maximum Allowable Shell Stress	PSI
ρS	Density of Carbon Steel	lb/in^3

Relevant Equations:			**Source**
1) Find Diameter and Length:			
V	=	$Q*\tau$	
D	=	$((4*V)/(AR*\pi))^{1/3}$	
L	=	$D*AR$	
2) Shell Thickness			
P_d	=	$\exp(0.60608+0.91615*\ln(P_O)+0.0015655*\ln(P_O)^2$	SSLW (22.61)
t_P	=	$(P_d*D)/(2*S*E-1.2*P_d)$	SSLW(22.60)
t_C	=	0.125 (in) Corrosion Allowance	
t_s	=	t_P+t_C	
3) Find Vessel Weight and Cost			
W	=	$\pi*(D+t_S)*(L+0.8*D)*t_V*\rho S$	SSLW(22.59)
C_V	=	$\exp(7.2756+0.18255*\ln(W)+0.02297*\ln(W)^2)$	SSLW(22.53)
F_M	=	1 (Carbon Steel)	
C_{VESSEL}	=	C_V*F_M	
4) Find Cost of Platforms and Ladders			
C_{PL}	=	$300.9*(D)^{0.63316}(L)^{0.80161}$	SSLW(22.58)
5) Find Total Veritical Pressure Vessel Cost			
C_{FOB}	=	$C_{VESSEL} + C_{PL}$	
F_{BM}	=	4.16	SSLW(T22.11)
C_P	=	$C_{FOB}*F_{BM}$	

Liquid-Liquid Extractor

Applicable Blocks: L-101

Variable	Description	Unit
Q_S	Solvent Volumetric Flow Rate	ft^3/hr
Q_F	Feed Volumetric Flow Rate	ft^3/hr
T_M	Maximum Throughput	$ft^3/hr\text{-}ft^2$
f	Fraction of Maximum Throughput	Dimensionless
N_S	Number of Equilibrium Stages	Dimensionless
HETP	Height of Theoretical Plate	ft
H_T	Additional Top Height	ft
H_B	Additional Bottom Height	ft

Relevant Equations: **Source**

1) Find Column Diameter:

$Q_T = Q_S+Q_F$		SSLW(S22.6)
$A_{CS,\ MIN} = Q_T/T_M$		SSLW(S22.6)
$A_{CS,\ ACTUAL} = A_{CS,\ MAX}/f$		SSLW(S22.6)
$D = ((4*A_{CS,\ ACTUAL})/\pi)^{0.5}$		SSLW(S22.6)

2) Find the Column Height:

$H = N_S*HETP+H_T+H_B$	SSLW(S22.6)

3) Find Vessel Cost:

$S = H*D^{1.5}$	SSLW(S22.6)
$C_V = 320*S^{0.84}$	SSLW(S22.6)
F_M = 2 (Stainless Steel)	SSLW(S22.6)
$C_{PL} = 300.9*(D)^{0.63316}(L)^{0.80161}$	
$C_{FOB} = C_V*F_M+C_{PL}$	
F_{BM} = 4.16	SSLW(T22.11)
$C_P = C_{FOB}*F_{BM}$	

F. Profitability Analysis (For both 10 and 15 years)

General Information

Process Title:	**Algae to Biodiesel**
Product:	**Biodiesel Fuel**
Plant Site Location:	**Thompson, TX**
Site Factor:	**1.00**
Operating Hours per Year:	**7920**
Operating Days Per Year:	**330**
Operating Factor:	**0.9041**

Product Information

This Process will Yield

33,963	gal of Biodiesel Fuel per hour
815,113	gal of Biodiesel Fuel per day
268,987,232	gal of Biodiesel Fuel per year

Price **$3.30 /gal**

Chronology

Year	Action	Distribution of Permanent Investment	Production Capacity	Depreciation 5 year MACRS	Product Price
2012	Design		0.0%		
2013	Construction	100%	0.0%		
2014	Production	0%	45.0%	20.00%	$3.30
2015	Production	0%	67.5%	32.00%	$3.38
2016	Production	0%	90.0%	19.20%	$3.47
2017	Production		90.0%	11.52%	$3.55
2018	Production		90.0%	11.52%	$3.64
2019	Production		90.0%	5.76%	$3.73
2020	Production		90.0%		$3.83
2021	Production		90.0%		$3.92
2022	Production		90.0%		$4.02
2023	Production		90.0%		$4.12
2024	Production		90.0%		$4.22
2025	Production		90.0%		$4.33
2026	Production		90.0%		$4.44
2027	Production		90.0%		$4.55
2028	Production		90.0%		$4.66

Equipment Costs

Equipment Description		Bare Module Cost
S-101	Storage	$121,518,100
S-102	Storage	$121,518,100
C-101	Process Machinery	$62,113,747
C-102	Process Machinery	$62,171,356
F-101	Process Machinery	$150,637,060
P-101	Process Machinery	$1,449,030
P-102	Process Machinery	$1,449,030
P-103	Process Machinery	$23,102,640
P-104	Process Machinery	$2,908,950
P-105	Process Machinery	$2,710,290
P-106	Process Machinery	$2,908,620
P-107	Process Machinery	$1,268,190
P-108	Process Machinery	$2,669,040
Spare Pumps MOD 1	Spares	$38,471,000
Land	Other Equipment	$128,400,000
G-101	Process Machinery	$3,578,220
G-102	Process Machinery	$3,578,220
C-101	Process Machinery	$771,400
C-102	Process Machinery	$771,400
C-103	Process Machinery	$771,400
C-104	Process Machinery	$771,400
C-105	Process Machinery	$771,400
D-101	Process Machinery	$688,658
D-102	Process Machinery	$688,658
D-103	Process Machinery	$688,658
D-104	Process Machinery	$688,658
D-105	Process Machinery	$688,658
D-106	Process Machinery	$688,658
D-107	Process Machinery	$688,500
D-108	Process Machinery	$688,500
D-109	Process Machinery	$688,500
D-110	Process Machinery	$688,500
D-111	Process Machinery	$688,500
D-112	Process Machinery	$688,500
D-113	Process Machinery	$688,500
D-114	Process Machinery	$688,500
D-115	Process Machinery	$688,500
D-116	Process Machinery	$688,500
Additional Equipment		$33,734,000
Total		***$779,059,541***

Raw Materials

	Raw Material:	Unit:	Required Ratio:		Cost of Raw Material:	
1	Methanol	MT	0.0003609	MT per gal of Biodiesel	$460.000	per MT
2	Potassium Hydroxide	Ton	3.635E-05	Ton per gal of Biodiesel	$600.00	per Ton
3	Hydrochloric Acid	MT	2.597E-05	MT per gal of Biodiesel	$130.00	per MT
4	f/2 Guillard's Algae Nutr	L	1.2803126	L per gal of Biodiesel F	$0.01	per L

Total Weighted Average: $0.204 per gal of Biodiesel Fuel

Byproducts

	Byproduct:	Unit:	Ratio to Product		Byproduct Selling Price	
1	Glycerol	lb	0.6454242	lb per gal of Biodiesel F	$0.220	per lb
2	Algae Biomass	lb	13.628757	lb per gal of Biodiesel F	$0.097	per lb
3		MT	7.547E-06	MT per gal of Biodiesel	$2.750	per MT

Total Weighted Average: $1.464 per gal of Biodiesel Fuel

Utilities

	Utility:	Unit:	Required Ratio		Utility Cost	
1	High Pressure Steam	lb	1.3433203	lb per gal of Biodiesel F	$6.600E-03	per lb
2	Medium Pressure Stear	lb	0.4897519	lb per gal of Biodiesel F	$4.800E-03	per lb
3	Process Water	gal	34.248888	gal per gal of Biodiesel	$7.500E-04	per gal
4	Cooling Water	gal	5.4675309	gal per gal of Biodiesel	$7.500E-05	per gal
5	Electricity	kWh	10.768677	kWh per gal of Biodiese	$0.060	per kWh
6	Chilled Brine	ton-day	0.0007498	ton-day per gal of Biodi	$2.400	per ton-day
7	Chilled Water	ton-day	0.016502	ton-day per gal of Biodi	$1.20	per ton-day
8	Wastewater Treatment	lb	0.0680107	lb per gal of Biodiesel F	$0.15	per lb

Total Weighted Average: $0.715 per gal of Biodiesel Fuel

Variable Costs

General Expenses:

Selling / Transfer Expenses:	**3.00%**	**of Sales**
Direct Research:	**4.80%**	**of Sales**
Allocated Research:	**0.50%**	**of Sales**
Administrative Expense:	**2.00%**	**of Sales**
Management Incentive Compensation:	**1.25%**	**of Sales**

Working Capital

Accounts Receivable	⇨	**30**	**Days**
Cash Reserves (excluding Raw Materi	⇨	**30**	**Days**
Accounts Payable	⇨	**30**	**Days**
Biodiesel Fuel Inventory	⇨	**5**	**Days**
Raw Materials	⇨	**5**	**Days**

Total Permanent Investment

Cost of Site Preparations:	**5.00%**	**of Total Bare Module Costs**
Cost of Service Facilities:	**5.00%**	**of Total Bare Module Costs**
Allocated Costs for utility plants and related facilities:	**$0**	
Cost of Contingencies and Contractor Fees:	**18.00%**	**of Direct Permanent Investment**
Cost of Land:	**0.00%**	**of Total Depreciable Capital**
Cost of Royalties:	**$0**	
Cost of Plant Start-Up:	**10.00%**	**of Total Depreciable Capital**

Fixed Costs

Operations

Operators per Shift:	**23**	**(assuming 5 shifts)**
Direct Wages and Benefits:	**$35**	**/operator hour**
Direct Salaries and Benefits:	**15%**	**of Direct Wages and Benefits**
Operating Supplies and Services:	**6%**	**of Direct Wages and Benefits**
Technical Assistance to Manufacturing:	**$60,000.00**	**per year, for each Operator per Shift**
Control Laboratory:	**$65,000.00**	**per year, for each Operator per Shift**

Maintenance

Wages and Benefits:	**4.50%**	**of Total Depreciable Capital**
Salaries and Benefits:	**25%**	**of Maintenance Wages and Benefits**
Materials and Services:	**100%**	**of Maintenance Wages and Benefits**
Maintenance Overhead:	**5%**	**of Maintenance Wages and Benefits**

Operating Overhead

General Plant Overhead:	**7.10%**	**of Maintenance and Operations Wages and Benefits**
Mechanical Department Services:	**2.40%**	**of Maintenance and Operations Wages and Benefits**
Employee Relations Department:	**5.90%**	**of Maintenance and Operations Wages and Benefits**
Business Services:	**7.40%**	**of Maintenance and Operations Wages and Benefits**

Property Taxes and Insurance

Property Taxes and Insurance:	**2%**	**of Total Depreciable Capital**

Straight Line Depreciation

Direct Plant:	8.00%	**of Total Depreciable Capital, less 1.18 times the Allocated Costs for Utility Plants and Related Facilities**
Allocated Plant:	6.00%	**of 1.18 times the Allocated Costs for Utility Plants and Related Facilities**

Other Annual Expenses

Rental Fees (Office and Laboratory Space):	**$0**
Licensing Fees:	**$0**
Miscellaneous:	**$0**

Depletion Allowance

Annual Depletion Allowance:	**$0**

Variable Cost Summary

Variable Costs at 100% Capacity:

General Expenses			
Selling / Transfer Expenses:		$	26,629,736
Direct Research:		$	42,607,578
Allocated Research:		$	4,438,289
Administrative Expense:		$	17,753,157
Management Incentive Compensation:		$	11,095,723
Total General Expenses		$	102,524,483
Raw Materials	$0.203732 per gal of Biodiesel Fue		$54,801,383
Byproducts	$1.464004 per gal of Biodiesel Fue		($393,798,258)
Utilities	$0.715238 per gal of Biodiesel Fue		$192,389,823
Total Variable Costs		**$**	**(44,082,569)**

Fixed Cost Summary

Operations		
Direct Wages and Benefits	$	8,372,000
Direct Salaries and Benefits	$	1,255,800
Operating Supplies and Services	$	502,320
Technical Assistance to Manufacturing	$	6,900,000
Control Laboratory	$	7,475,000
Total Operations	**$**	**24,505,120**
Maintenance		
Wages and Benefits	$	45,504,868
Salaries and Benefits	$	11,376,217
Materials and Services	$	45,504,868
Maintenance Overhead	$	2,275,243
Total Maintenance	**$**	**104,661,196**
Operating Overhead		
General Plant Overhead:	$	4,722,131
Mechanical Department Services:	$	1,596,213
Employee Relations Department:	$	3,924,024
Business Services:	$	4,921,657
Total Operating Overhead	**$**	**15,164,026**
Property Taxes and Insurance		
Property Taxes and Insurance:	$	20,224,386
Other Annual Expenses		
Rental Fees (Office and Laboratory Space):	$	-
Licensing Fees:	$	-
Miscellaneous:	$	-
Total Other Annual Expenses	**$**	**-**
Total Fixed Costs	**$**	**164,554,727**

Investment Summary

Bare Module Costs				
Fabricated Equipment	$	-		
Process Machinery	$	345,516,141		
Spares	$	39,533,700		
Storage	$	265,609,700		
Other Equipment	$	128,400,000		
Catalysts	$	-		
Computers, Software, Etc.	$	-		
Total Bare Module Costs:			**$**	**779,059,541**
Direct Permanent Investment				
Cost of Site Preparations:	$	38,952,977		
Cost of Service Facilities:	$	38,952,977		
Allocated Costs for utility plants and related facilities:	$	-		
Direct Permanent Investment			**$**	**856,965,495**
Total Depreciable Capital				
Cost of Contingencies & Contractor Fees	$	154,253,789		
Total Depreciable Capital			**$**	**1,011,219,284**
Total Permanent Investment				
Cost of Land:	$	-		
Cost of Royalties:	$	-		
Cost of Plant Start-Up:	$	101,121,928		
Total Permanent Investment - Unadjusted			$	1,112,341,213
Site Factor				1.00
Total Permanent Investment			**$**	**1,112,341,213**

Working Capital

		2013		2014		2015
Accounts Receivable	$	32,831,181	$	16,415,591	$	16,415,591
Cash Reserves	$	13,202,059	$	6,601,029	$	6,601,029
Accounts Payable	$	(9,142,688)	$	(4,571,344)	$	(4,571,344)
Biodiesel Fuel Inventory	$	5,471,864	$	2,735,932	$	2,735,932
Raw Materials	$	337,817	$	168,908	$	168,908
Total	**$**	**42,700,232**	**$**	**21,350,116**	**$**	**21,350,116**
Present Value at 15%	*$*	*37,130,636*	*$*	*16,143,755*	*$*	*14,038,048*
Total Capital Investment			**$**	**1,179,653,652**		

10 Year Cash Flow Summary

Cash Flow Summary

Year	Percentage of Design Capacity	Product Unit Price	Sales	Capital Costs	Working Capital	Var Costs	Fixed Costs	Depreciation	Depletion Allowance	Taxible Income	Taxes	Taxes	Net Earnings	Cash Flow	Cumulative Net Present Value at 15%
2012	0%		-	-	-	-	-	-	-	-	-	-	-	-	-
2013	0%		-	(1,112,341,200)	(42,700,200)	-	-	-	-	-	-	-	-	(1,155,041,400)	(1,004,383,900)
2014	45%	$3.30	399,446,000	-	(21,350,100)	19,837,200	(164,554,700)	(202,243,900)	-	52,484,600	(19,419,306)	(19,419,300)	33,065,300	213,959,000	(842,600,100)
2015	68%	$3.38	614,148,300	-	(21,350,100)	30,499,600	(169,491,400)	(323,590,200)	-	151,566,400	(56,079,558)	(56,079,600)	95,486,800	397,726,900	(581,088,200)
2016	90%	$3.47	839,336,000	-	-	41,682,800	(174,576,100)	(194,154,100)	-	512,288,600	(189,546,783)	(189,546,800)	322,741,800	516,895,900	(285,551,300)
2017	90%	$3.55	860,319,400	-	-	42,724,900	(179,813,400)	(116,492,500)	-	606,738,400	(224,493,219)	(224,493,200)	382,245,200	498,737,700	(37,590,500)
2018	90%	$3.64	881,827,400	-	-	43,793,000	(185,207,800)	(116,492,500)	-	623,920,100	(230,850,450)	(230,850,400)	393,069,700	509,562,100	182,707,300
2019	90%	$3.73	903,873,100	-	-	44,887,800	(190,764,000)	(58,246,200)	-	699,750,600	(258,907,737)	(258,907,700)	440,842,900	499,089,100	370,333,300
2020	90%	$3.83	926,469,900	-	-	46,010,000	(196,487,000)	-	-	775,993,000	(287,117,400)	(287,117,400)	488,875,600	488,875,600	530,147,600
2021	90%	$3.92	949,631,600	-	-	47,160,300	(202,381,600)	-	-	794,410,400	(293,931,835)	(293,931,800)	500,478,500	500,478,500	672,414,900
2022	90%	$4.02	973,372,400	-	-	48,339,300	(208,453,000)	-	-	813,258,700	(300,905,725)	(300,905,700)	512,353,000	512,353,000	799,060,700
2023	90%	$4.12	997,706,700	-	85,400,500	49,547,800	(214,706,600)	-	-	832,547,900	(308,042,730)	(308,042,700)	524,505,200	609,905,700	930,155,800

15 Year Cash Flow Summary

Cash Flow Summary

Year	Percentage of Design Capacity	Product Unit Price	Sales	Capital Costs	Working Capital	Var Costs	Fixed Costs	Depreciation	Depletion Allowance	Taxible Income	Taxes	Taxes	Net Earnings	Cash Flow	Cumulative Net Present Value at 15%
2012	0%		-	-	-	-	-	-	-	-	-	-	-	-	-
2013	0%		-	(1,112,341,200)	(42,700,200)	-	-	-	-	-	-	-	-	(1,155,041,400)	(1,004,383,900)
2014	45%	$3.30	399,446,000	-	(21,350,100)	19,837,200	(164,554,700)	(202,243,900)	-	52,484,600	(19,419,306)	(19,419,300)	33,065,300	213,959,000	(842,600,100)
2015	68%	$3.38	614,148,300	-	(21,350,100)	30,499,600	(169,491,400)	(323,590,200)	-	151,566,400	(56,079,558)	(56,079,600)	95,486,800	397,726,900	(581,088,200)
2016	90%	$3.47	839,336,000	-	-	41,682,800	(174,576,100)	(194,154,100)	-	512,288,600	(189,546,783)	(189,546,800)	322,741,800	516,895,900	(285,551,300)
2017	90%	$3.55	860,319,400	-	-	42,724,900	(179,813,400)	(116,492,500)	-	606,738,400	(224,493,219)	(224,493,200)	382,245,200	498,737,700	(37,590,500)
2018	90%	$3.64	881,827,400	-	-	43,793,000	(185,207,800)	(116,492,500)	-	623,920,100	(230,850,450)	(230,850,400)	393,069,700	509,562,100	182,707,300
2019	90%	$3.73	903,873,100	-	-	44,887,800	(190,764,000)	(58,246,200)	-	699,750,600	(258,907,737)	(258,907,700)	440,842,900	499,089,100	370,333,300
2020	90%	$3.83	926,469,900	-	-	46,010,000	(196,487,000)	-	-	775,993,000	(287,117,400)	(287,117,400)	488,875,600	488,875,600	530,147,600
2021	90%	$3.92	949,631,600	-	-	47,160,300	(202,381,600)	-	-	794,410,400	(293,931,835)	(293,931,800)	500,478,500	500,478,500	672,414,900
2022	90%	$4.02	973,372,400	-	-	48,339,300	(208,453,000)	-	-	813,258,700	(300,905,725)	(300,905,700)	512,353,000	512,353,000	799,060,700
2023	90%	$4.12	997,706,700	-	-	49,547,800	(214,706,600)	-	-	832,547,900	(308,042,730)	(308,042,700)	524,505,200	524,505,200	911,799,500
2024	90%	$4.22	1,022,649,400	-	-	50,786,500	(221,147,800)	-	-	852,288,100	(315,346,591)	(315,346,600)	536,941,500	536,941,500	1,012,157,700
2025	90%	$4.33	1,048,215,600	-	-	52,056,100	(227,782,200)	-	-	872,489,500	(322,821,132)	(322,821,100)	549,668,400	549,668,400	1,101,494,200
2026	90%	$4.44	1,074,421,000	-	-	53,357,500	(234,615,700)	-	-	893,162,900	(330,470,263)	(330,470,300)	562,692,600	562,692,600	1,181,018,800
2027	90%	$4.55	1,101,281,600	-	-	54,691,500	(241,654,200)	-	-	914,318,900	(338,297,981)	(338,298,000)	576,020,900	576,020,900	1,251,808,600
2028	90%	$4.66	1,128,813,600	-	85,400,500	56,058,800	(248,903,800)	-	-	935,968,600	(346,308,370)	(346,308,400)	589,660,200	675,060,700	1,323,948,800

10 Year Profitability Summary

Profitability Measures

The Internal Rate of Return (IRR) for this project is 33.42%

The Net Present Value (NPV) of this project in 2012 is $ 930,155,800

ROI Analysis (Third Production Year)

Annual Sales	**839,335,990**
Annual Costs	**(132,893,286)**
Depreciation	**(88,987,297)**
Income Tax	**(228,458,501)**
Net Earnings	**388,996,907**
Total Capital Investmer	**1,197,741,676**
ROI	**32.48%**

Sensitivity Analyses

	Vary Initial Value by +/-
x-axis	50%
y-axis	50%

Product Price	Variable Costs										
	-22,041,285	-26,449,542	-30,857,799	-35,266,056	-39,674,312	***-44,082,569***	-48,490,826	-52,899,083	-57,307,340	-61,715,597	-66,123,854
$1.65	10.81%	11.09%	11.38%	11.66%	11.94%	12.21%	12.49%	12.76%	13.03%	13.30%	13.57%
$1.98	16.09%	16.33%	16.58%	16.83%	17.07%	17.31%	17.56%	17.80%	18.04%	18.27%	18.51%
$2.31	20.73%	20.96%	21.18%	21.40%	21.63%	21.85%	22.06%	22.28%	22.50%	22.72%	22.93%
$2.64	24.95%	25.16%	25.36%	25.57%	25.77%	25.98%	26.18%	26.38%	26.58%	26.79%	26.99%
$2.97	28.85%	29.04%	29.24%	29.43%	29.62%	29.81%	30.00%	30.19%	30.38%	30.57%	30.76%
$3.30	32.51%	32.69%	32.87%	33.06%	33.24%	**33.42%**	33.60%	33.78%	33.96%	34.14%	34.32%
$3.63	35.97%	36.14%	36.32%	36.49%	36.67%	36.84%	37.01%	37.19%	37.36%	37.53%	37.71%
$3.96	39.27%	39.44%	39.61%	39.78%	39.94%	40.11%	40.28%	40.44%	40.61%	40.78%	40.94%
$4.29	42.44%	42.60%	42.77%	42.93%	43.09%	43.25%	43.41%	43.58%	43.74%	43.90%	44.06%
$4.62	45.50%	45.66%	45.81%	45.97%	46.13%	46.29%	46.44%	46.60%	46.76%	46.91%	47.07%
$4.95	48.46%	48.61%	48.76%	48.92%	49.07%	49.22%	49.38%	49.53%	49.68%	49.83%	49.99%

Profitability Measures

The Internal Rate of Return (IRR) for this project is 35.16%

The Net Present Value (NPV) of this project in 2012 is $ 1,323,948,800

ROI Analysis (Third Production Year)

Annual Sales	**839,335,990**
Annual Costs	**(132,893,286)**
Depreciation	**(88,987,297)**
Income Tax	**(228,458,501)**
Net Earnings	**388,996,907**
Total Capital Investmer	**1,197,741,676**
ROI	**32.48%**

Sensitivity Analyses

	Vary Initial Value by +/-
x-axis	50%
y-axis	50%

Product Price	Variable Costs: -22,041,285	-26,449,542	-30,857,799	-35,266,056	-39,674,312	***-44,082,569***	-48,490,826	-52,899,083	-57,307,340	-61,715,597	-66,123,854
$1.65	14.61%	14.87%	15.12%	15.37%	15.62%	15.86%	16.11%	16.35%	16.60%	16.84%	17.08%
$1.98	19.32%	19.54%	19.76%	19.98%	20.20%	20.42%	20.64%	20.85%	21.07%	21.28%	21.50%
$2.31	23.50%	23.70%	23.91%	24.11%	24.31%	24.51%	24.71%	24.91%	25.10%	25.30%	25.50%
$2.64	27.33%	27.52%	27.71%	27.90%	28.09%	28.28%	28.46%	28.65%	28.83%	29.02%	29.20%
$2.97	30.92%	31.10%	31.28%	31.45%	31.63%	31.81%	31.99%	32.16%	32.34%	32.52%	32.69%
$3.30	34.31%	34.48%	34.65%	34.82%	34.99%	**35.16%**	35.33%	35.50%	35.67%	35.84%	36.01%
$3.63	37.55%	37.71%	37.87%	38.04%	38.20%	38.37%	38.53%	38.69%	38.86%	39.02%	39.18%
$3.96	40.66%	40.82%	40.98%	41.14%	41.29%	41.45%	41.61%	41.77%	41.93%	42.09%	42.24%
$4.29	43.66%	43.82%	43.97%	44.13%	44.28%	44.44%	44.59%	44.75%	44.90%	45.05%	45.21%
$4.62	46.58%	46.73%	46.88%	47.03%	47.19%	47.34%	47.49%	47.64%	47.79%	47.94%	48.09%
$4.95	49.42%	49.57%	49.71%	49.86%	50.01%	50.16%	50.30%	50.45%	50.60%	50.75%	50.89%

G. Materials, Safety and Data Sheets

Material Safety Data Sheet
Instant FAME/Instant Anaerobe Methods
Methanol

SECTION 1 – CHEMICAL PRODUCT AND COMPANY IDENTIFICATION

MSDS Name: Methanol
MSDS Preparation Date: 06/19/2009
Synonyms or Generic ID for Methanol: Carbinol; Methyl alcohol; Methyl hydroxide; Monohydroxymethane; Wood alcohol; Wood naptha; Wood spirits; Columbian spirits; Methanol.
Chemical Family: Methanol Family
Formula: CH_3OH
Molecular Weight: N/A
PIN (UN#/ NA#): UN1230
Company Identification:
Microbial ID.
125 Sandy Drive
Newark, DE 19713
For Information, call: (800)276-8068, (302)737-4297
For Domestic CHEMTREC assistance, call: 800-424-9300
For International CHEMTREC assistance, call: 703-527-3887

SECTION 2 – COMPOSITION, INFORMATION ON INGREDIENTS

67-56-1	Methanol	<99%	200-659-6	Irritant, Flammable

NFPA Rating: (estimated) Health: 1; Flammability: 3; Instability: 0

State: Liquid	Appearance: colorless	Odor: Alcohol-like, weak odor
Boiling Point: 64.7°C@760mmHg	pH: Not available	Specific Gravity: 7910g/cm3@20°C
Vapor Pressure (mm Hg): 128mmHg @20°C		Vapor Density (AIR=1): 1.11
Flash Point: 12°C		Solubility in Water: miscible

SECTION 3 – HAZARDS IDENTIFICATION

Appearance: Colorless liquid, Flash Point: 12°C, 53.6°F.

Danger! Poison! May be fatal or cause blindness if swallowed. Vapor harmful. **Flammable liquid and vapor.** Harmful if swallowed, inhaled, or absorbed through the skin. Causes eye, skin, and respiratory tract irritation. May cause central nervous system depression. Cannot be made non-poisonous.
Target Organs: Eyes, nervous system, optic nerve.

Potential Health Effects
Eye: May cause painful sensitization to light. Methanol is a mild to moderate eye irritant. Inhalation, ingestion or skin absorption of methanol can cause significant disturbance in vision, including blindness.
Skin: Causes moderate skin irritation. May be absorbed through the skin in harmful amounts. Prolonged and or repeated contact may cause defatting of skin and dermatitis. Methanol can be absorbed through the skin, producing systemic effects that include visual disturbances.
Ingestion: May be fatal or cause blindness if swallowed. Aspiration hazard. Cannot be made non-poisonous. May cause gastrointestinal irritation with nausea, vomiting and diarrhea. May cause systematic toxicity with acidosis. May cause central nervous system depression, characterized by excitement, followed by headache, dizziness, drowsiness, and nausea. Advanced stages may cause collapse, unconsciousness, coma, and possible death due to failed respiratory failure. May cause cardiopulmonary system effects.

Inhalation: Methanol is toxic and can very readily form extremely high vapor concentrations at room temperature. Inhalation is the most common route of occupational exposure. At first, methanol causes CNS depression with nausea, headache, vomiting, dizziness and incoordination. A time period with no obvious symptoms follows (typically 8-24 hrs). This latent period is followed by metabolic acidosis and severe visual effects which may include reduced reactivity and/or increased sensitivity to light, blurred, doubl and/or snowy vision, and blindness. Depending on the severity of exposure and the promptness of treatment, survivors may recover completely or may have permanent blindness, vision disturbances and/or nervous system effects.
Chronic: Prolonged or repeated skin contact may cause dermatitis. Chronic exposure may cause effects similar to those of acute exposure. Methanol is only very slowly eliminated from the body. Because of this slow elimination, methanol should be regarded as a cumulative poison. Though a single exposure may cause no effect, daily exposures may result in the accumulation of a harmful amount. Methanol has produced fetotoxicity in rats and teratogenicity in mice exposed by inhalation to high concentrations that did not produce significant maternal toxicity.

SECTION 4 – FIRST AID MEASURES

Eyes: In case of contact, immediately flush eyes with plenty of water for a t least 15 minutes. Get medical aid.
Skin: In case of contact, immediately flush skin with plenty of water for at least 15 minutes while removing contaminated clothing and shoes. Get medical aid immediately. Wash clothing before reuse.
Ingestion: Potential for aspiration if swallowed. Get medical aid immediately. Do not induce vomiting unless directed to do so by medical personnel. Never give anything by mouth to an unconscious person. If vomiting occurs naturally, have victim lean forward.
Inhalation: If inhaled, remove to fresh air. If not breathing, give artificial respiration. If breathing is difficult, give oxygen. Get medical aid.
Notes to Physician: Effects may be delayed.
Antidote: Ethanol may inhibit methanol metabolism.

SECTION 5 – FIRE FIGHTING MEASURES

General Information: Ethanol may inhibit methanol metabolism. As in any fire, wear a self-contained breathing apparatus in pressure-demand, MSHA/NIOSH (approved or equivalent), and full protective gear. During a fire, irritating and highly toxic gases may be generated by thermal decomposition or combustion. Use water spray to keep fire-exposed containers cool. Water may be ineffective. Material is lighter than water and a fire may be spread by the use of water. Vapors are heavier than air and may travel to a source of ignition and flash back. Vapors can spread along the ground and collect in low or confined areas.
Extinguishing Media: For small fires, use dry chemical, carbon dioxide, water spray or alcohol-resistant foam. Water may be ineffective. For large fires, use water spray, fog or alcohol-resistant foam. Do NOT use straight streams of water.
Flash Point: 12 deg C (53.60 deg F)
Autoignition Temperature: 455 deg C (851.00 deg F)
Explosion Limits, Lower:6.0 vol %
Upper: 31.00 vol %
NFPA Rating: (estimated) Health: 1; Flammability: 3; Instability: 0

SECTION 6 – ACCIDENTAL RELEASE MEASURES

General Information: Use proper personal protective equipment as indicated in Section 8.
Spills/Leaks: Use water spray to disperse the gas/vapor. Remove all sources of ignition. Absorb spill using an absorbent, non-combustible material such as earth, sand, or vermiculite. Do not use combustible materials such as sawdust. Use a spark-proof tool. Provide ventilation. A vapor suppressing foam may be used to reduce vapors. Water spray may reduce vapor but may not prevent ignition in closed spaces.

SECTON 10 – STABILITY AND REACTIVITY

Chemical Stability: Stable under normal temperatures and pressures.
Conditions to Avoid: High temperatures, ignition sources, confined spaces.
Incompatibilities with Other Materials: Oxidizing agents, reducing agents, acids, alkali metals, potassium, sodium, metals as powders (e.g. hafnium, raney nickel), acid anhydrides, acid chlorides, powdered aluminum, powdered magnesium.
Hazardous Decomposition Products: Carbon monoxide, irritating and toxic fumes and gases, carbon dioxide, formaldehyde.
Hazardous Polymerization: Will not occur.

SECTION 11 – TOXICOLOGICAL INFORMATION

RTECS#:
CAS# 67-56-1: PC1400000
LD50/LC50:
CAS# 67-56-1:
Draize test, rabbit, eye: 40 mg Moderate;
Draize test, rabbit, eye: 100 mg/24H Moderate;
Draize test, rabbit, skin: 20 mg/24H Moderate;
Inhalation, rabbit: LC50 = 81000 mg/m3/14H;
Inhalation, rat: LC50 = 64000 ppm/4H;
Oral, mouse: LD50 = 7300 mg/kg;
Oral, rabbit: LD50 = 14200 mg/kg;
Oral, rat: LD50 = 5600 mg/kg;
Skin, rabbit: LD50 = 15800 mg/kg;

Human LDLo Oral: 143 mg/kg; Human LDLo Oral: 428 mg/kg; Human TCLo Inhalation; 300 ppm caused visual field changes & headache; Monkey LDLo Skin: 393 mg/kg. Methanol is significantly less toxic to most experimental animals than humans, because most animal species metabolize methanol differently. Non-primate species do not ordinarily show symptoms of metabolic acidosis or the visual effects which have been observed in primates and humans.
Carcinogenicity:
CAS# 67-56-1: Not listed by ACGIH, IARC, NTP, or CA Prop 65.

Epidemiology: No information found
Teratogenicity: There is no human information available. Methanol is considered to be a potential developmental hazard based on animal data. In animal experiments, methanol has caused fetotoxic or teratogenic effects without maternal toxicity.
Reproductive Effects: See actual entry in RTECS for complete information.
Mutagenicity: See actual entry in RTECS for complete information.
Neurotoxicity: ACGIH cites neuropathy, vision and CNS under TLV basis.

SECTION 12 – ECOLOGICAL INFORMATION

Ecotoxicity: Fish: Fathead Minnow: 29.4 g/L; 96 Hr; LC50 (unspecified)Fish: Goldfish: 250 ppm; 11 Hr; resulted in deathFish: Rainbow trout: 8000 mg/L; 48 Hr; LC50 (unspecified)Fish: Rainbow trout: LC50 = 13-68 mg/L; 96 Hr.; 12 degrees CFish: Fathead Minnow: LC50 = 29400 mg/L; 96 Hr.; 25 degrees C, pH 7.63Fish: Rainbow trout: LC50 = 8000 mg/L; 48 Hr.; UnspecifiedBacteria: Phytobacterium phosphoreum: EC50 = 51,000-320,000 mg/L; 30 minutes; Microtox test No data available.
Environmental: Dangerous to aquatic life in high concentrations. Aquatic toxicity rating: TLm 96>1000 ppm. May be dangerous if it enters water intakes. Methyl alcohol is expected to biodegrade in soil and water very rapidly. This product will show high soil mobility and will be degraded from the ambient atmosphere by the reaction with photochemically produced hyroxyl radicals with an estimated half-life of 17.8 days. Bioconcentration factor for fish (golden ide) < 10. Based on a log Kow of -0.77, the BCF value for methanol can be estimated to be 0.2.
Physical: No information available.
Other: No information available.

SECTION 13 – DISPOSAL CONSIDERATIONS

Chemical waste generators must determine whether a discarded chemical is classified as a hazardous waste. US EPA guidelines for the classification determination are listed in 40 CFR Parts 261.3. Additionally, waste generators must consult state and local hazardous waste regulations to ensure complete and accurate classification.

RCRA P-Series: None listed.

RCRA U-Series:

CAS# 67-56-1: waste number U154 (Ignitable waste).

SECTION 14 – TRANSPORT INFORMATION

	US DOT	CANADA TDG
Shipping Name:	Methanol	Methanol
Hazard Class:	3	3
UN Number:	UN1230	UN1230
Packing Group:	II	II
Additional Information		Flash Point 12°C

SECTION 15 – REGULATORY INFORMATION

US FEDERAL

TSCA

CAS# 67-56-1 is listed on the TSCA inventory.

Health & Safety Reporting List

None of the chemicals are on the Health & Safety Reporting List.

Chemical Test Rules

None of the chemicals in this product are under a Chemical Test Rule.

Section 12b

None of the chemicals are listed under TSCA Section 12b.

TSCA Significant New Use Rule

None of the chemicals in this material have a SNUR under TSCA.

CERCLA Hazardous Substances and corresponding RQs

CAS# 67-56-1: 5000 lb final RQ; 2270 kg final RQ

SARA Section 302 Extremely Hazardous Substances

None of the chemicals in this product have a TPQ.

SARA Codes

CAS # 67-56-1: immediate, fire.

Section 313

This material contains Methanol (CAS# 67-56-1, > 99%),which is subject to the reporting requirements of Section 313 of SARA Title III and 40 CFR Part 373.

Clean Air Act:

CAS# 67-56-1 is listed as a hazardous air pollutant (HAP).
This material does not contain any Class 1 Ozone depletors.
This material does not contain any Class 2 Ozone depletors.

Clean Water Act:

None of the chemicals in this product are listed as Hazardous Substances under the CWA.
None of the chemicals in this product are listed as Priority Pollutants under the CWA.
None of the chemicals in this product are listed as Toxic Pollutants under the CWA.

OSHA:

None of the chemicals in this product are considered highly hazardous by OSHA.

STATE

CAS# 67-56-1 can be found on the following state right to know lists: California, New Jersey, Pennsylvania, Minnesota, Massachusetts.

California Prop 65
California No Significant Risk Level: None of the chemicals in this product are listed.

European/International Regulations

European Labeling in Accordance with EC Directives
Hazard Symbols:
T F
Risk Phrases:
R 11 Highly flammable.
R 23/24/25 Toxic by inhalation, in contact with skin and if swallowed.
R 39/23/24/25 Toxic : danger of very serious irreversible effects through inhalation, in contact with skin and if swallowed.

Safety Phrases:
S 16 Keep away from sources of ignition - No smoking.
S 36/37 Wear suitable protective clothing and gloves.
S 45 In case of accident or if you feel unwell, seek medical advice immediately (show the label where possible).
S 7 Keep container tightly closed.

WGK (Water Danger/Protection)
CAS# 67-56-1: 1
Canada - DSL/NDSL
CAS# 67-56-1 is listed on Canada's DSL List.
Canada - WHMIS
This product has a WHMIS classification of B2, D1B, D2B.
This product has been classified in accordance with the hazard criteria of the Controlled Products Regulations and the MSDS contains all of the information required by those regulations.
Canadian Ingredient Disclosure List
CAS# 67-56-1 is listed on the Canadian Ingredient Disclosure List.

SECTION 16 – Other Information

This Material Safety Data Sheet has been prepared in accordance with 29 CFR 1910.1200 and contains information believed to be accurate and complete at the date of preparation. The statements contained herein are offered for informational purposes only and are based upon technical data. MIDI Inc. believes them to be accurate but does not purport to be all-inclusive. The above-stated product is intended for use only by persons having the necessary technical skills and facilities for handling the product at their discretion and risk. Since conditions and manner of use are outside our control, we (MIDI Inc.) make no warranty of merchantability or any such warranty, express or implied with respect to information and we assume no liability resulting from the above product or its use. Users should make their own investigations to determine suitability of information and product for their particular purposes.

MSDS Number: **H3880** * * * * * *Effective Date:* ***11/21/08*** * * * * * *Supercedes:* ***01/19/06***

MSDS Material Safety Data Sheet

24 Hour Emergency Telephone: 908-859-2151
CHEMTREC: 1-800-424-9300

National Response in Canada
CANUTEC: 613-996-6666

Outside U.S. and Canada
Chemtrec: 703-527-3887

NOTE: CHEMTREC, CANUTEC and National Response Center emergency numbers to be used only in the event of chemical emergencies involving a spill, leak, fire, exposure or accident involving chemicals.

From: Mallinckrodt Baker, Inc.
222 Red School Lane
Phillipsburg, NJ 08865

Mallinckrodt CHEMICALS J.T.Baker

All non-emergency questions should be directed to Customer Service (1-800-582-2537) for assistance.

HYDROCHLORIC ACID, 33 - 40%

1. Product Identification

Synonyms: Muriatic acid; hydrogen chloride, aqueous
CAS No.: 7647-01-0
Molecular Weight: 36.46
Chemical Formula: HCl
Product Codes:
J.T. Baker: 5367, 5537, 5575, 5800, 5814, 5821, 5839, 5861, 5862, 5894, 5962, 5963, 5972, 5994, 6900, 7831, 9529, 9530, 9534, 9535, 9536, 9538, 9539, 9540, 9544, 9548, 9551
Mallinckrodt: 2062, 2515, 2612, 2624, 2626, 3861, 5583, 5587, H611, H613, H616, H987, H992, H999, V078, V628

2. Composition/Information on Ingredients

Ingredient	CAS No	Percent	Hazardous
Hydrogen Chloride	7647-01-0	33 - 40%	Yes
Water	7732-18-5	60 - 67%	No

3. Hazards Identification

Emergency Overview

POISON! DANGER! CORROSIVE. LIQUID AND MIST CAUSE SEVERE BURNS TO ALL BODY TISSUE. MAY BE FATAL IF SWALLOWED OR INHALED. INHALATION MAY CAUSE LUNG DAMAGE.

SAF-T-DATA$^{(tm)}$ Ratings (Provided here for your convenience)

Health Rating: 3 - Severe (Poison)

Flammability Rating: 0 - None
Reactivity Rating: 2 - Moderate
Contact Rating: 4 - Extreme (Corrosive)
Lab Protective Equip: GOGGLES & SHIELD; LAB COAT & APRON; VENT HOOD; PROPER GLOVES
Storage Color Code: White (Corrosive)

Potential Health Effects

Inhalation:
Corrosive! Inhalation of vapors can cause coughing, choking, inflammation of the nose, throat, and upper respiratory tract, and in severe cases, pulmonary edema, circulatory failure, and death.
Ingestion:
Corrosive! Swallowing hydrochloric acid can cause immediate pain and burns of the mouth, throat, esophagus and gastrointestinal tract. May cause nausea, vomiting, and diarrhea. Swallowing may be fatal.
Skin Contact:
Corrosive! Can cause redness, pain, and severe skin burns. Concentrated solutions cause deep ulcers and discolor skin.
Eye Contact:
Corrosive! Vapors are irritating and may cause damage to the eyes. Contact may cause severe burns and permanent eye damage.
Chronic Exposure:
Long-term exposure to concentrated vapors may cause erosion of teeth. Long term exposures seldom occur due to the corrosive properties of the acid.
Aggravation of Pre-existing Conditions:
Persons with pre-existing skin disorders or eye disease may be more susceptible to the effects of this substance.

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4. First Aid Measures

Inhalation:
Remove to fresh air. If not breathing, give artificial respiration. If breathing is difficult, give oxygen. Get medical attention immediately.
Ingestion:
DO NOT INDUCE VOMITING! Give large quantities of water or milk if available. Never give anything by mouth to an unconscious person. Get medical attention immediately.
Skin Contact:
In case of contact, immediately flush skin with plenty of water for at least 15 minutes while removing contaminated clothing and shoes. Wash clothing before reuse. Thoroughly clean shoes before reuse. Get medical attention immediately.
Eye Contact:
Immediately flush eyes with plenty of water for at least 15 minutes, lifting lower and upper eyelids occasionally. Get medical attention immediately.

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5. Fire Fighting Measures

Fire:
Extreme heat or contact with metals can release flammable hydrogen gas.
Explosion:
Not considered to be an explosion hazard.
Fire Extinguishing Media:
If involved in a fire, use water spray. Neutralize with soda ash or slaked lime.
Special Information:

In the event of a fire, wear full protective clothing and NIOSH-approved self-contained breathing apparatus with full facepiece operated in the pressure demand or other positive pressure mode. Structural firefighter's protective clothing is ineffective for fires involving hydrochloric acid. Stay away from ends of tanks. Cool tanks with water spray until well after fire is out.

6. Accidental Release Measures

Ventilate area of leak or spill. Wear appropriate personal protective equipment as specified in Section 8. Isolate hazard area. Keep unnecessary and unprotected personnel from entering. Contain and recover liquid when possible. Neutralize with alkaline material (soda ash, lime), then absorb with an inert material (e. g., vermiculite, dry sand, earth), and place in a chemical waste container. Do not use combustible materials, such as saw dust. Do not flush to sewer! US Regulations (CERCLA) require reporting spills and releases to soil, water and air in excess of reportable quantities. The toll free number for the US Coast Guard National Response Center is (800) 424-8802.

J. T. Baker NEUTRASORB® acid neutralizers are recommended for spills of this product.

7. Handling and Storage

Store in a cool, dry, ventilated storage area with acid resistant floors and good drainage. Protect from physical damage. Keep out of direct sunlight and away from heat, water, and incompatible materials. Do not wash out container and use it for other purposes. When diluting, the acid should always be added slowly to water and in small amounts. Never use hot water and never add water to the acid. Water added to acid can cause uncontrolled boiling and splashing. When opening metal containers, use non-sparking tools because of the possibility of hydrogen gas being present. Containers of this material may be hazardous when empty since they retain product residues (vapors, liquid); observe all warnings and precautions listed for the product.

8. Exposure Controls/Personal Protection

Airborne Exposure Limits:
For Hydrochloric acid:
- OSHA Permissible Exposure Limit (PEL):
5 ppm (Ceiling)
- ACGIH Threshold Limit Value (TLV):
2 ppm (Ceiling), A4 Not classifiable as a human carcinogen
Ventilation System:
A system of local and/or general exhaust is recommended to keep employee exposures below the Airborne Exposure Limits. Local exhaust ventilation is generally preferred because it can control the emissions of the contaminant at its source, preventing dispersion of it into the general work area. Please refer to the ACGIH document, *Industrial Ventilation, A Manual of Recommended Practices*, most recent edition, for details.
Personal Respirators (NIOSH Approved):
If the exposure limit is exceeded, a full facepiece respirator with an acid gas cartridge may be worn up to 50 times the exposure limit or the maximum use concentration specified by the appropriate regulatory agency or respirator supplier, whichever is lowest. For emergencies or instances where the exposure levels are not known, use a full-facepiece positive-pressure, air-supplied respirator. WARNING: Air purifying respirators do not protect workers in oxygen-deficient atmospheres.
Skin Protection:
Rubber or neoprene gloves and additional protection including impervious boots, apron, or coveralls, as needed in areas of unusual exposure to prevent skin contact.
Eye Protection:
Use chemical safety goggles and/or a full face shield where splashing is possible. Maintain eye wash fountain

and quick-drench facilities in work area.

9. Physical and Chemical Properties

Appearance:
Colorless, fuming liquid.
Odor:
Pungent odor of hydrogen chloride.
Solubility:
Infinite in water with slight evolution of heat.
Density:
1.18
pH:
For HCL solutions: 0.1 (1.0 N), 1.1 (0.1 N), 2.02 (0.01 N)
% Volatiles by volume @ 21C (70F):
100
Boiling Point:
53C (127F) Azeotrope (20.2%) boils at 109C (228F)
Melting Point:
-74C (-101F)
Vapor Density (Air=1):
No information found.
Vapor Pressure (mm Hg):
190 @ 25C (77F)
Evaporation Rate (BuAc=1):
No information found.

10. Stability and Reactivity

Stability:
Stable under ordinary conditions of use and storage. Containers may burst when heated.
Hazardous Decomposition Products:
When heated to decomposition, emits toxic hydrogen chloride fumes and will react with water or steam to produce heat and toxic and corrosive fumes. Thermal oxidative decomposition produces toxic chlorine fumes and explosive hydrogen gas.
Hazardous Polymerization:
Will not occur.
Incompatibilities:
A strong mineral acid, concentrated hydrochloric acid is incompatible with many substances and highly reactive with strong bases, metals, metal oxides, hydroxides, amines, carbonates and other alkaline materials. Incompatible with materials such as cyanides, sulfides, sulfites, and formaldehyde.
Conditions to Avoid:
Heat, direct sunlight.

11. Toxicological Information

Inhalation rat LC50: 3124 ppm/1H; oral rabbit LD50: 900 mg/kg (Hydrochloric acid concentrated); investigated as a tumorigen, mutagen, reproductive effector.

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--------\Cancer Lists\------------------------------------------------------
                                           ---NTP Carcinogen---
```

Ingredient	Known	Anticipated	IARC Category
Hydrogen Chloride (7647-01-0)	No	No	3
Water (7732-18-5)	No	No	None

12. Ecological Information

Environmental Fate:
When released into the soil, this material is not expected to biodegrade. When released into the soil, this material may leach into groundwater.
Environmental Toxicity:
This material is expected to be toxic to aquatic life.

13. Disposal Considerations

Whatever cannot be saved for recovery or recycling should be handled as hazardous waste and sent to a RCRA approved waste facility. Processing, use or contamination of this product may change the waste management options. State and local disposal regulations may differ from federal disposal regulations. Dispose of container and unused contents in accordance with federal, state and local requirements.

14. Transport Information

Domestic (Land, D.O.T.)

Proper Shipping Name: HYDROCHLORIC ACID
Hazard Class: 8
UN/NA: UN1789
Packing Group: II
Information reported for product/size: 475LB

International (Water, I.M.O.)

Proper Shipping Name: HYDROCHLORIC ACID
Hazard Class: 8
UN/NA: UN1789
Packing Group: II
Information reported for product/size: 475LB

15. Regulatory Information

Risk and Safety Phrases:
Symbol: C
Risk: 34-37
Safety: (1/2-)26-45

--------\Chemical Inventory Status - Part 1\----------------------------------

Ingredient	TSCA	EC	Japan	Australia
Hydrogen Chloride (7647-01-0)	Yes	Yes	Yes	Yes
Water (7732-18-5)	Yes	Yes	Yes	Yes

--------\Chemical Inventory Status - Part 2\---------------------------------

Ingredient	Korea	Canada DSL	Canada NDSL	Phil.
Hydrogen Chloride (7647-01-0)	Yes	Yes	No	Yes
Water (7732-18-5)	Yes	Yes	No	Yes

--------\Federal, State & International Regulations - Part 1\----------------

Ingredient	SARA 302- RQ	SARA 302- TPQ	SARA 313 List	SARA 313 Chemical Catg.
Hydrogen Chloride (7647-01-0)	5000	500*	Yes	No
Water (7732-18-5)	No	No	No	No

--------\Federal, State & International Regulations - Part 2\----------------

Ingredient	CERCLA	-RCRA- 261.33	-TSCA- 8(d)
Hydrogen Chloride (7647-01-0)	5000	No	No
Water (7732-18-5)	No	No	No

Chemical Weapons Convention: No TSCA 12(b): No CDTA: Yes
SARA 311/312: Acute: Yes Chronic: Yes Fire: No Pressure: No
Reactivity: No (Mixture / Liquid)

Australian Hazchem Code: 2R
Poison Schedule: None allocated.
WHMIS:
This MSDS has been prepared according to the hazard criteria of the Controlled Products Regulations (CPR) and the MSDS contains all of the information required by the CPR.

---

16. Other Information

NFPA Ratings: Health: **3** Flammability: **0** Reactivity: **1**
Label Hazard Warning:
POISON! DANGER! CORROSIVE. LIQUID AND MIST CAUSE SEVERE BURNS TO ALL BODY TISSUE. MAY BE FATAL IF SWALLOWED OR INHALED. INHALATION MAY CAUSE LUNG DAMAGE.
Label Precautions:
Do not get in eyes, on skin, or on clothing.
Do not breathe vapor or mist.
Use only with adequate ventilation.
Wash thoroughly after handling.
Store in a tightly closed container.
Remove and wash contaminated clothing promptly.
Label First Aid:
In case of contact, immediately flush eyes or skin with plenty of water for at least 15 minutes while removing contaminated clothing and shoes. Wash clothing before reuse. If swallowed, DO NOT INDUCE VOMITING. Give large quantities of water. Never give anything by mouth to an unconscious person. If inhaled, remove to fresh air. If not breathing, give artificial respiration. If breathing is difficult, give oxygen. In all cases get medical attention immediately.
Product Use:
Laboratory Reagent.
Revision Information:
No Changes.
Disclaimer:
****

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Prepared by: Environmental Health & Safety
Phone Number: (314) 654-1600 (U.S.A.)

MSDS Number: **P5884** * * * * * *Effective Date:* ***09/01/09*** * * * * * *Supercedes:* ***02/01/07***

MSDS Material Safety Data Sheet

24 Hour Emergency Telephone: 908-859-2151
CHEMTREC: 1-800-424-9300

National Response in Canada
CANUTEC: 613-996-6666

Outside U.S. and Canada
Chemtrec: 703-527-3887

NOTE: CHEMTREC, CANUTEC and National Response Center emergency numbers to be used only in the event of chemical emergencies involving a spill, leak, fire, exposure or accident involving chemicals.

From: Mallinckrodt Baker, Inc.
222 Red School Lane
Phillipsburg, NJ 08865

Mallinckrodt CHEMICALS J.T.Baker

All non-emergency questions should be directed to Customer Service (1-800-582-2537) for assistance.

POTASSIUM HYDROXIDE

1. Product Identification

Synonyms: Caustic potash; potassium hydrate
CAS No.: 1310-58-3
Molecular Weight: 56.11
Chemical Formula: KOH
Product Codes:
J.T. Baker: 3140, 3141, 3146, 3150, 3152, 5685
Mallinckrodt: 6964, 6976, 6984, 7704, 7815

2. Composition/Information on Ingredients

Ingredient	CAS No	Percent	Hazardous
Potassium Hydroxide	1310-58-3	85 - 90%	Yes
Water	7732-18-5	10 - 15%	No

3. Hazards Identification

Emergency Overview

POISON! DANGER! CORROSIVE. CAUSES SEVERE BURNS TO SKIN, EYES, RESPIRATORY TRACT, AND GASTROINTESTINAL TRACT. MATERIAL IS EXTREMELY DESTRUCTIVE TO ALL BODY TISSUES. MAY BE FATAL IF SWALLOWED. HARMFUL IF INHALED.

SAF-T-DATA(tm) Ratings (Provided here for your convenience)

Health Rating: 3 - Severe (Poison)
Flammability Rating: 0 - None
Reactivity Rating: 2 - Moderate

Contact Rating: 4 - Extreme (Corrosive)
Lab Protective Equip: GOGGLES & SHIELD; LAB COAT & APRON; VENT HOOD; PROPER GLOVES
Storage Color Code: White Stripe (Store Separately)

Potential Health Effects

Inhalation:
Severe irritant. Effects from inhalation of dust or mist vary from mild irritation to serious damage of the upper respiratory tract, depending on the severity of exposure. Symptoms may include coughing, sneezing, damage to the nasal or respiratory tract. High concentrations can cause lung damage.
Ingestion:
Toxic! Swallowing may cause severe burns of mouth, throat and stomach. Other symptoms may include vomiting, diarrhea. Severe scarring of tissue and death may result. Estimated lethal dose: 5 grams.
Skin Contact:
Corrosive! Contact with skin can cause irritation or severe burns and scarring with greater exposures.
Eye Contact:
Highly Corrosive! Causes irritation of eyes with tearing, redness, swelling. Greater exposures cause severe burns with possible blindness resulting.
Chronic Exposure:
Prolonged contact with dilute solutions or dust of potassium hydroxide has a destructive effect on tissue.
Aggravation of Pre-existing Conditions:
Persons with pre-existing skin disorders or eye problems or impaired respiratory function may be more susceptible to the effects of the substance.

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4. First Aid Measures

Inhalation:
Remove to fresh air. If not breathing, give artificial respiration. If breathing is difficult, give oxygen. Call a physician.
Ingestion:
If swallowed, DO NOT INDUCE VOMITING. Give large quantities of water. Never give anything by mouth to an unconscious person. Get medical attention immediately.
Skin Contact:
In case of contact, immediately flush skin with plenty of water for at least 15 minutes while removing contaminated clothing and shoes. Wash clothing before reuse. Thoroughly clean shoes before reuse. Get medical attention immediately.
Eye Contact:
Immediately flush eyes with plenty of water for at least 15 minutes, lifting lower and upper eyelids occasionally. Get medical attention immediately.

---

5. Fire Fighting Measures

Fire:
Not combustible, but contact with water or moisture may generate enough heat to ignite combustibles.
Explosion:
Can react with chemically reactive metals such as aluminum, zinc, magnesium, copper, etc. to release hydrogen gas which can form explosive mixtures with air.
Fire Extinguishing Media:
Use any means suitable for extinguishing surrounding fire.
Special Information:
Solution process causes formation of corrosive mists. Hot or molten material can react violently with water. In

the event of a fire, wear full protective clothing and NIOSH-approved self-contained breathing apparatus with full facepiece operated in the pressure demand or other positive pressure mode.

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6. Accidental Release Measures

Ventilate area of leak or spill. Keep unnecessary and unprotected people away from area of spill. Wear appropriate personal protective equipment as specified in Section 8. Spills: Pick up and place in a suitable container for reclamation or disposal, using a method that does not generate dust. Do not flush caustic residues to the sewer. Residues from spills can be diluted with water, neutralized with dilute acid such as acetic, hydrochloric or sulfuric. Absorb neutralized caustic residue on clay, vermiculite or other inert substance and package in a suitable container for disposal.
US Regulations (CERCLA) require reporting spills and releases to soil, water and air in excess of reportable quantities. The toll free number for the US Coast Guard National Response Center is (800) 424-8802.

J. T. Baker NEUTRACIT®-2 or BuCAIM® caustic neutralizers are recommended for spills of solutions of this product.

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7. Handling and Storage

Keep in a tightly closed container, stored in a cool, dry, ventilated area. Protect against physical damage. Isolate from incompatible substances. Protect from moisture. Addition to water releases heat which can result in violent boiling and spattering. Always add slowly and in small amounts. Never use hot water. Containers of this material may be hazardous when empty since they retain product residues (dust, solids); observe all warnings and precautions listed for the product.

---

8. Exposure Controls/Personal Protection

Airborne Exposure Limits:
- OSHA Permissible Exposure Limit (PEL):
2 mg/m3 Ceiling
- ACGIH Threshold Limit Value (TLV):
2 mg/m3 Ceiling

Ventilation System:
A system of local and/or general exhaust is recommended to keep employee exposures below the Airborne Exposure Limits. Local exhaust ventilation is generally preferred because it can control the emissions of the contaminant at its source, preventing dispersion of it into the general work area. Please refer to the ACGIH document, *Industrial Ventilation, A Manual of Recommended Practices*, most recent edition, for details.
Personal Respirators (NIOSH Approved):
If the exposure limit is exceeded and engineering controls are not feasible, a half facepiece particulate respirator (NIOSH type N95 or better filters) may be worn for up to ten times the exposure limit or the maximum use concentration specified by the appropriate regulatory agency or respirator supplier, whichever is lowest.. A full-face piece particulate respirator (NIOSH type N100 filters) may be worn up to 50 times the exposure limit, or the maximum use concentration specified by the appropriate regulatory agency, or respirator supplier, whichever is lowest. If oil particles (e.g. lubricants, cutting fluids, glycerine, etc.) are present, use a NIOSH type R or P filter. For emergencies or instances where the exposure levels are not known, use a full-facepiece positive-pressure, air-supplied respirator. WARNING: Air-purifying respirators do not protect workers in oxygen-deficient atmospheres.
Skin Protection:
Rubber or neoprene gloves and additional protection including impervious boots, apron, or coveralls, as needed in areas of unusual exposure.

Eye Protection:
Use chemical safety goggles and/or a full face shield where splashing is possible. Maintain eye wash fountain and quick-drench facilities in work area.

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9. Physical and Chemical Properties

Appearance:
White deliquescent solid
Odor:
Odorless.
Solubility:
52.8% in water @ 20C (68F)
Specific Gravity:
2.04
pH:
13.5 (0.1 molar solution)
% Volatiles by volume @ 21C (70F):
0
Boiling Point:
1320C (2408F)
Melting Point:
360C (680F)
Vapor Density (Air=1):
No information found.
Vapor Pressure (mm Hg):
1.0 @ 714C (1317F)
Evaporation Rate (BuAc=1):
No information found.

---

10. Stability and Reactivity

Stability:
Stable under ordinary conditions of use and storage.
Hazardous Decomposition Products:
Carbon monoxide when reacting with carbohydrates, and hydrogen gas when reacting with aluminum, zinc and tin. Thermal oxidation can produce toxic fumes of potassium oxide (K_2O).
Hazardous Polymerization:
Will not occur.
Incompatibilities:
Contact with water, acids, flammable liquids and organic halogen compounds, especially trichloroethylene, may cause fire or explosion. Contact with nitromethane and other similar nitro compounds cause formation of shock sensitive salts. Contact with metals such as aluminum, tin and zinc causes formation of flammable hydrogen gas.
Conditions to Avoid:
Heat, moisture, incompatibles.

---

11. Toxicological Information

For potassium hydroxide: Oral rat LD50: 273 mg/kg; Investigated as a mutagen. Skin Irritation Data (std Draize, 50 mg/24 H): Human, Severe; Rabbit, Severe. Eye Irritation Data(Rabbit, non-std test,1 mg/24 H, rinse): Moderate.

--------\Cancer Lists\--

Ingredient	---NTP Carcinogen--- Known	Anticipated	IARC Category
Potassium Hydroxide (1310-58-3)	No	No	None
Water (7732-18-5)	No	No	None

12. Ecological Information

Environmental Fate:
No information found.
Environmental Toxicity:
Potassium Hydroxide: TLm: 80 ppm/Mosquito fish/ 24 hr./ Fresh water

13. Disposal Considerations

Whatever cannot be saved for recovery or recycling should be handled as hazardous waste and sent to a RCRA approved waste facility. Processing, use or contamination of this product may change the waste management options. State and local disposal regulations may differ from federal disposal regulations. Dispose of container and unused contents in accordance with federal, state and local requirements.

14. Transport Information

Domestic (Land, D.O.T.)

Proper Shipping Name: POTASSIUM HYDROXIDE, SOLID
Hazard Class: 8
UN/NA: UN1813
Packing Group: II
Information reported for product/size: 110LB

International (Water, I.M.O.)

Proper Shipping Name: POTASSIUM HYDROXIDE, SOLID
Hazard Class: 8
UN/NA: UN1813
Packing Group: II
Information reported for product/size: 110LB

International (Air, I.C.A.O.)

Proper Shipping Name: POTASSIUM HYDROXIDE, SOLID
Hazard Class: 8
UN/NA: UN1813
Packing Group: II
Information reported for product/size: 110LB

15. Regulatory Information

--------\Chemical Inventory Status - Part 1\---------------------------------

Ingredient	TSCA	EC	Japan	Australia
Potassium Hydroxide (1310-58-3)	Yes	Yes	Yes	Yes

```
  Water (7732-18-5)                                Yes  Yes   Yes      Yes

  --------\Chemical Inventory Status - Part 2\---------------------------------
                                                          --Canada--
  Ingredient                                      Korea  DSL   NDSL  Phil.
  -----------------------------------------------  -----  ---   ----  -----
  Potassium Hydroxide (1310-58-3)                  Yes   Yes   No     Yes
  Water (7732-18-5)                                Yes   Yes   No     Yes

  --------\Federal, State & International Regulations - Part 1\----------------
                                            -SARA 302-    ------SARA 313------
  Ingredient                                RQ    TPQ     List  Chemical Catg.
  -----------------------------------------  ---   -----   ----  --------------
  Potassium Hydroxide (1310-58-3)           No    No      No          No
  Water (7732-18-5)                         No    No      No          No

  --------\Federal, State & International Regulations - Part 2\----------------
                                                       -RCRA-    -TSCA-
  Ingredient                                CERCLA     261.33     8(d)
  -----------------------------------------  ------     ------    ------
  Potassium Hydroxide (1310-58-3)           1000       No         No
  Water (7732-18-5)                         No         No         No


Chemical Weapons Convention:  No     TSCA 12(b):  No     CDTA:  No
SARA 311/312:  Acute: Yes      Chronic: Yes  Fire: No  Pressure: No
Reactivity: Yes         (Mixture / Solid)
```

Australian Hazchem Code: 2R
Poison Schedule: S6
WHMIS:
This MSDS has been prepared according to the hazard criteria of the Controlled Products Regulations (CPR) and the MSDS contains all of the information required by the CPR.

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16. Other Information

NFPA Ratings: Health: **3** Flammability: **0** Reactivity: **1**
Label Hazard Warning:
POISON! DANGER! CORROSIVE. CAUSES SEVERE BURNS TO SKIN, EYES, RESPIRATORY TRACT, AND GASTROINTESTINAL TRACT. MATERIAL IS EXTREMELY DESTRUCTIVE TO ALL BODY TISSUES. MAY BE FATAL IF SWALLOWED. HARMFUL IF INHALED.
Label Precautions:
Do not get in eyes, on skin, or on clothing.
Do not breathe dust.
Keep container closed.
Use only with adequate ventilation.
Wash thoroughly after handling.
Label First Aid:
If swallowed, DO NOT INDUCE VOMITING. Give large quantities of water. Never give anything by mouth to an unconscious person. In case of contact, immediately flush eyes or skin with plenty of water for at least 15 minutes while removing contaminated clothing and shoes. Wash clothing before reuse. If inhaled, remove to fresh air. If not breathing give artificial respiration. If breathing is difficult, give oxygen. In all cases get medical attention immediately.
Product Use:
Laboratory Reagent.
Revision Information:
No Changes.
Disclaimer:
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Prepared by: Environmental Health & Safety
Phone Number: (314) 654-1600 (U.S.A.)

MSDS Number: **W0600** * * * * * *Effective Date:* ***09/09/09*** * * * * * *Supercedes:* ***05/04/07***

24 Hour Emergency Telephone: 908-859-2151
CHEMTREC: 1-800-424-9300

National Response in Canada
CANUTEC: 613-996-6666

Outside U.S. and Canada
Chemtrec: 703-527-3887

MSDS Material Safety Data Sheet

From: Mallinckrodt Baker, Inc.
222 Red School Lane
Phillipsburg, NJ 08865

Mallinckrodt CHEMICALS J.T.Baker

NOTE: CHEMTREC, CANUTEC and National Response Center emergency numbers to be used only in the event of chemical emergencies involving a spill, leak, fire, exposure or accident involving chemicals.

All non-emergency questions should be directed to Customer Service (1-800-582-2537) for assistance.

Water

1. Product Identification

Synonyms: Hydrogen oxide; Dihydrogen oxide; Distilled water
CAS No.: 7732-18-5
Molecular Weight: 18.02
Chemical Formula: H2O
Product Codes:
J.T. Baker: 4022, 4201, 4212, 4216, 4218, 4219, 4221, 6906, 9823, 9831, XL-317
Mallinckrodt: 6795, H453, V564

2. Composition/Information on Ingredients

Ingredient	CAS No	Percent	Hazardous
Water	7732-18-5	100%	No

3. Hazards Identification

Emergency Overview

Not applicable.

SAF-T-DATA(tm) Ratings (Provided here for your convenience)

Health Rating: 0 - None
Flammability Rating: 0 - None
Reactivity Rating: 1 - Slight
Contact Rating: 0 - None
Lab Protective Equip: GOGGLES; LAB COAT
Storage Color Code: Green (General Storage)

Potential Health Effects

Water is non-hazardous.

Inhalation:
Not applicable.
Ingestion:
Not applicable.
Skin Contact:
Not applicable.
Eye Contact:
Not applicable.
Chronic Exposure:
Not applicable.
Aggravation of Pre-existing Conditions:
Not applicable.

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4. First Aid Measures

Inhalation:
Not applicable.
Ingestion:
Not applicable.
Skin Contact:
Not applicable.
Eye Contact:
Not applicable.

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5. Fire Fighting Measures

Fire:
Not applicable.
Explosion:
Not applicable.
Fire Extinguishing Media:
Use extinguishing media appropriate for surrounding fire.
Special Information:
In the event of a fire, wear full protective clothing and NIOSH-approved self-contained breathing apparatus with full facepiece operated in the pressure demand or other positive pressure mode.

---

6. Accidental Release Measures

Non-hazardous material. Clean up of spills requires no special equipment or procedures.

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7. Handling and Storage

Keep container tightly closed. Suitable for any general chemical storage area. Protect from freezing. Water is considered a non-regulated product, but may react vigorously with some specific materials. Avoid contact with all materials until investigation shows substance is compatible.

---

8. Exposure Controls/Personal Protection

Airborne Exposure Limits:
Not applicable.
Ventilation System:
Not applicable.
Personal Respirators (NIOSH Approved):
Not applicable.
Skin Protection:
None required.
Eye Protection:
None required.

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9. Physical and Chemical Properties

Appearance:
Clear, colorless liquid.
Odor:
Odorless.
Solubility:
Complete (100%)
Specific Gravity:
1.00
pH:
7.0
% Volatiles by volume @ 21C (70F):
100
Boiling Point:
100C (212F)
Melting Point:
0C (32F)
Vapor Density (Air=1):
Not applicable.
Vapor Pressure (mm Hg):
17.5 @ 20C (68F)
Evaporation Rate (BuAc=1):
No information found.

---

10. Stability and Reactivity

Stability:
Stable under ordinary conditions of use and storage.
Hazardous Decomposition Products:
Not applicable.
Hazardous Polymerization:
Will not occur.
Incompatibilities:
Strong reducing agents, acid chlorides, phosphorus trichloride, phosphorus pentachloride, phosphorus

oxychloride.
Conditions to Avoid:
No information found.

11. Toxicological Information

For Water: LD50 Oral Rat: >90 ml/Kg. Investigated as a mutagen.

--------\Cancer Lists\--

Ingredient	---NTP Carcinogen--- Known	Anticipated	IARC Category
Water (7732-18-5)	No	No	None

12. Ecological Information

Environmental Fate:
Not applicable.
Environmental Toxicity:
Not applicable.

13. Disposal Considerations

Whatever cannot be saved for recovery or recycling should be flushed to sewer. If material becomes contaminated during use, dispose of accordingly. Dispose of container and unused contents in accordance with federal, state and local requirements.

14. Transport Information

Not regulated.

15. Regulatory Information

--------\Chemical Inventory Status - Part 1\---------------------------------

Ingredient	TSCA	EC	Japan	Australia
Water (7732-18-5)	Yes	Yes	Yes	Yes

--------\Chemical Inventory Status - Part 2\---------------------------------

Ingredient	Korea	--Canada-- DSL	NDSL	Phil.
Water (7732-18-5)	Yes	Yes	No	Yes

--------\Federal, State & International Regulations - Part 1\----------------

Ingredient	-SARA 302- RQ	TPQ	------SARA 313------ List	Chemical Catg.
Water (7732-18-5)	No	No	No	No

--------\Federal, State & International Regulations - Part 2\----------------

Ingredient	CERCLA	-RCRA- 261.33	-TSCA- 8(d)
Water (7732-18-5)	No	No	No

```
Chemical Weapons Convention:  No     TSCA 12(b):  No     CDTA:  No
SARA 311/312:  Acute: No       Chronic: No   Fire: No  Pressure: No
Reactivity: No          (Pure / Liquid)
```

Australian Hazchem Code: None allocated.
Poison Schedule: None allocated.
WHMIS:
This MSDS has been prepared according to the hazard criteria of the Controlled Products Regulations (CPR) and the MSDS contains all of the information required by the CPR.

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16. Other Information

NFPA Ratings: Health: **0** Flammability: **0** Reactivity: **0**
Label Hazard Warning:
Not applicable.
Label Precautions:
Keep in tightly closed container.
Label First Aid:
Not applicable.
Product Use:
Laboratory Reagent.
Revision Information:
No Changes.
Disclaimer:

Prepared by: Environmental Health & Safety
Phone Number: (314) 654-1600 (U.S.A.)

SAMPLE MATERIAL
SAFETY DATA SHEET

1. CHEMICAL PRODUCT

General Product Name:	**Biodiesel (B100)**
Synonyms:	Methyl Soyate, Rapeseed Methyl Ester (RME)
Product Description:	Methyl esters from lipid sources
CAS Number:	Methyl Soyate: 67784-80-9; RME: 73891-99-3;

2. COMPOSITION/INFORMATION ON INGREDIENTS

This product contains no hazardous materials.

3. HAZARDS IDENTIFICATION

Potential Health Effects:

INHALATION:

Negligible unless heated to produce vapors. Vapors or finely misted materials may irritate the mucous membranes and cause irritation, dizziness, and nausea. Remove to fresh air.

EYE CONTACT:

May cause irritation. Irrigate eye with water for at least 15 to 20 minutes. Seek medical attention if symptoms persist.

SKIN CONTACT:

Prolonged or repeated contact is not likely to cause significant skin irritation. Material is sometimes encountered at elevated temperatures. Thermal burns are possible.

INGESTION:

No hazards anticipated from ingestion incidental to industrial exposure.

4. FIRST AID MEASURES

EYES:

Irrigate eyes with a heavy stream of water for at least 15 to 20 minutes.

SKIN:

Wash exposed areas of the body with soap and water.

INHALATION:

Remove from area of exposure; seek medical attention if symptoms persist.

INGESTION:

Give one or two glasses of water to drink. If gastro-intestinal symptoms develop, consult medical personnel. (Never give anything by mouth to an unconscious person.)

5. FIRE FIGHTING MEASURES

Flash Point (Method Used): 130.0 C or 266.0 F min (ASTM 93)

Flammability Limits: None known

EXTINGUISHING MEDIA:

Dry chemical, foam, halon (may not be permissible in some countries), CO_2, water spray (fog). Water stream may splash the burning liquid and spread fire.

SPECIAL FIRE FIGHTING PROCEDURES:

Use water spray to cool drums exposed to fire.

UNUSUAL FIRE AND EXPLOSION HAZARDS:

Biodiesel soaked rags or spill absorbents (i.e. oil dry, polypropylene socks, sand, etc.) can cause spontaneous combustion if stored near combustibles and not handled properly. Store biodiesel soaked rags or spill absorbents in approved safety containers and dispose of properly. Oil soaked rags may be washed with soap and water and allowed to dry in well ventilated area. Firefighters should use self-contained breathing apparatus to avoid exposure to smoke and vapor.

6. ACCIDENTAL RELEASE MEASURES SPILL CLEAN-UP PROCEDURES

Remove sources of ignition, contain spill to smallest area possible. Stop leak if possible. Pick up small spills with absorbent materials and dispose of properly to avoid spontaneous combustion (see unusual fire and explosion hazards above).

Recover large spills for salvage or disposal. Wash hard surfaces with safety solvent or detergent to remove remaining oil film. Greasy nature will result in a slippery surface.

7. HANDLING AND STORAGE

Store in closed containers between 50°F and 120°F.
Keep away from oxidizing agents, excessive heat, and ignition sources.
Store and use in well ventilated areas.
Do not store or use near heat, spark, or flame, store out of sun.
Do not puncture, drag, or slide this container.
Drum is not a pressure vessel; never use pressure to empty.

8. EXPOSURE CONTROL /PERSONAL PROTECTION

RESPIRATORY PROTECTION:

If vapors or mists are generated, wear a NIOSH approved organic vapor/mist respirator.

PROTECTIVE CLOTHING:

Safety glasses, goggles, or face shield recommended to protect eyes from mists or splashing. PVC coated gloves recommended to prevent skin contact.

OTHER PROTECTIVE MEASURES:

Employees must practice good personal hygiene, washing exposed areas of skin several times daily and laundering contaminated clothing before re-use.

9. PHYSICAL AND CHEMICAL PROPERTIES

Boiling Point, 760 mm Hg:>200°C	Volatiles, % by Volume: <2
Specific Gravity (H_2O=1): 0.88	Solubility in H_2O, % by Volume: insoluble
Vapor Pressure, mm Hg: <2	Evaporation Rate, Butyl Acetate=1: <1
Vapor Density, Air=1:>1	

Appearance and Odor: pale yellow liquid, mild odor

10. STABILITY AND REACTIVITY

GENERAL:

This product is stable and hazardous polymerization will not occur.

INCOMPATIBLE MATERIALS AND CONDITIONS TO AVOID:

Strong oxidizing agents

HAZARDOUS DECOMPOSITION PRODUCTS:

Combustion produces carbon monoxide, carbon dioxide along with thick smoke.

11. DISPOSAL CONSIDERATIONS

WASTE DISPOSAL:

Waste may be disposed of by a licensed waste disposal company. Contaminated absorbent material may be disposed of in an approved landfill. Follow local, state and federal disposal regulations.

12. TRANSPORT INFORMATION

UN HAZARD CLASS: N/A

NMFC (National Motor Freight Classification):

PROPER SHIPPING NAME: Fatty acid ester
IDENTIFICATION NUMBER: 144920
SHIPPING CLASSIFICATION: 65

13. REGULATORY INFORMATION:

OSHA STATUS:

This product is not hazardous under the criteria of the Federal OSHA Hazard Communication Standard 29 CFR 1910.1200. However, thermal processing and decomposition fumes from this product may be hazardous as noted in Sections 2 and 3.

TSCA STATUS:

This product is listed on TSCA.

CERCLA (Comprehensive Response Compensation and Liability Act):

NOT reportable.

SARA TITLE III (Superfund Amendments and Reauthorization Act):

Section 312 Extremely Hazardous Substances:

None

Section 311/312 Hazard Categories:

Non-hazardous under Section 311/312

Section 313 Toxic Chemicals:

None

RCRA STATUS:

If discarded in its purchased form, this product would not be a hazardous waste either by listing or by characteristic. However, under RCRA, it is the responsibility of the product user to determine at the time of disposal, whether a material containing the product or derived from the product should be classified as a hazardous waste,
(40 CFR 261.20-24)

CALIFORNIA PROPOSITION 65:

The following statement is made in order to comply with the California Safe Drinking Water and Toxic Enforcement Act of 1986. This product contains no chemicals known to the state of California to cause cancer.

14. OTHER INFORMATION:

This information relates only to the specific material designated and may not be valid for such material used in combination with any other materials or in any other process. Such information is to the best of the company's knowledge and believed accurate and reliable as of the date indicated. However, no representation, warranty or guarantee of any kind, express or implied, is made as to its accuracy, reliability or completeness and we assume no responsibility for any loss, damage or expense, direct or consequential, arising out of use. It is the user's responsibility to satisfy himself as to the suitableness and completeness of such information for his own particular use.

MSDS Number: **G4774** * * * * * *Effective Date:* ***02/15/08*** * * * * * *Supercedes:* ***05/25/05***

MSDS Material Safety Data Sheet

From: Mallinckrodt Baker, Inc.
222 Red School Lane
Phillipsburg, NJ 08865

Mallinckrodt CHEMICALS
J.T.Baker

24 Hour Emergency Telephone: 908-859-2151
CHEMTREC: 1-800-424-9300

National Response in Canada
CANUTEC: 613-996-6666

Outside U.S. and Canada
Chemtrec: 703-527-3887

NOTE: CHEMTREC, CANUTEC and National Response Center emergency numbers to be used only in the event of chemical emergencies involving a spill, leak, fire, exposure or accident involving chemicals.

All non-emergency questions should be directed to Customer Service (1-800-582-2537) for assistance.

GLYCEROL

1. Product Identification

Synonyms: 1,2,3-propanetriol; glycerin; glycol alcohol; glycerol, anhydrous
CAS No.: 56-81-5
Molecular Weight: 92.10
Chemical Formula: C3H5(OH)3
Product Codes:
J.T. Baker: 2135, 2136, 2140, 2142, 2143, 2988, 4043, 5093, 72138, M778
Mallinckrodt: 0564, 5092, 5093, 5100

2. Composition/Information on Ingredients

Ingredient	CAS No	Percent	Hazardous
Glycerin	56-81-5	90 - 100%	Yes

3. Hazards Identification

Emergency Overview

CAUTION! MAY CAUSE IRRITATION TO SKIN, EYES, AND RESPIRATORY TRACT. MAY AFFECT KIDNEYS.

SAF-T-DATA(tm) Ratings (Provided here for your convenience)

Health Rating: 2 - Moderate (Life)
Flammability Rating: 1 - Slight
Reactivity Rating: 0 - None
Contact Rating: 1 - Slight
Lab Protective Equip: GOGGLES; LAB COAT; VENT HOOD; PROPER GLOVES

Storage Color Code: Green (General Storage)

Potential Health Effects

Inhalation:
Due to the low vapor pressure, inhalation of the vapors at room temperatures is unlikely. Inhalation of mist may cause irritation of respiratory tract.
Ingestion:
Low toxicity. May cause nausea, headache, diarrhea.
Skin Contact:
May cause irritation.
Eye Contact:
May cause irritation.
Chronic Exposure:
May cause kidney injury.
Aggravation of Pre-existing Conditions:
Persons with pre-existing skin disorders or eye problems or impaired liver or kidney function may be more susceptible to the effects of the substance.

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4. First Aid Measures

Inhalation:
Remove to fresh air. Get medical attention for any breathing difficulty.
Ingestion:
Induce vomiting immediately as directed by medical personnel. Never give anything by mouth to an unconscious person. Get medical attention.
Skin Contact:
Immediately flush skin with plenty of water for at least 15 minutes. Remove contaminated clothing and shoes. Wash clothing before reuse. Thoroughly clean shoes before reuse. Get medical attention if irritation develops.
Eye Contact:
Immediately flush eyes with plenty of water for at least 15 minutes, lifting upper and lower eyelids occasionally. Get medical attention if irritation persists.

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5. Fire Fighting Measures

Fire:
Flash point: 199C (390F) CC
Autoignition temperature: 370C (698F)
Slight fire hazard when exposed to heat or flame. Slight fire hazard when exposed to heat or flame.
Explosion:
Above flash point, vapor-air mixtures may cause flash fire.
Fire Extinguishing Media:
Use any means suitable for extinguishing surrounding fire. Water spray may be used to extinguish surrounding fire and cool exposed containers. Water spray will also reduce fume and irritant gases.
Special Information:
In the event of a fire, wear full protective clothing and NIOSH-approved self-contained breathing apparatus with full facepiece operated in the pressure demand or other positive pressure mode.

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6. Accidental Release Measures

Ventilate area of leak or spill. Wear appropriate personal protective equipment as specified in Section 8. Contain and recover liquid when possible. Collect liquid in an appropriate container or absorb with an inert material (e. g., vermiculite, dry sand, earth), and place in a chemical waste container. Do not use combustible materials, such as saw dust. Do not flush to sewer!

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7. Handling and Storage

Keep in a tightly closed container, stored in a cool, dry, ventilated area. Protect against physical damage. Isolate from incompatible substances. Containers of this material may be hazardous when empty since they retain product residues (vapors, liquid); observe all warnings and precautions listed for the product.

---

8. Exposure Controls/Personal Protection

Airborne Exposure Limits:
For Glycerin Mist:
- OSHA Permissible Exposure Limit (PEL):
Total Dust: 15 mg/m3 (TWA);
Respirable Fraction: 5 mg/m3(TWA).
- ACGIH Threshold Limit Value (TLV):
10 mg/m3
Ventilation System:
A system of local and/or general exhaust is recommended to keep employee exposures below the Airborne Exposure Limits. Local exhaust ventilation is generally preferred because it can control the emissions of the contaminant at its source, preventing dispersion of it into the general work area. Please refer to the ACGIH document, *Industrial Ventilation, A Manual of Recommended Practices*, most recent edition, for details.
Personal Respirators (NIOSH Approved):
If the exposure limit is exceeded and engineering controls are not feasible, a half facepiece particulate respirator (NIOSH type P95 or R95 filters) may be worn for up to ten times the exposure limit or the maximum use concentration specified by the appropriate regulatory agency or respirator supplier, whichever is lowest.. A full-face piece particulate respirator (NIOSH type P100 or R100 filters) may be worn up to 50 times the exposure limit, or the maximum use concentration specified by the appropriate regulatory agency, or respirator supplier, whichever is lowest. Please note that N filters are not recommended for this material. For emergencies or instances where the exposure levels are not known, use a full-facepiece positive-pressure, air-supplied respirator. WARNING: Air-purifying respirators do not protect workers in oxygen-deficient atmospheres.
Skin Protection:
Wear protective gloves and clean body-covering clothing.
Eye Protection:
Use chemical safety goggles. Maintain eye wash fountain and quick-drench facilities in work area.

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9. Physical and Chemical Properties

Appearance:
Clear oily liquid.
Odor:
Odorless.
Solubility:
Miscible in water.
Specific Gravity:
1.26 @ 20C/4C
pH:

(neutral to litmus)
% Volatiles by volume @ 21C (70F):
0
Boiling Point:
290C (554F)
Melting Point:
18C (64F)
Vapor Density (Air=1):
3.17
Vapor Pressure (mm Hg):
0.0025 @ 50C (122F)
Evaporation Rate (BuAc=1):
No information found.

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10. Stability and Reactivity

Stability:
Stable under ordinary conditions of use and storage.
Hazardous Decomposition Products:
Toxic gases and vapors may be released if involved in a fire. Glycerin decomposes upon heating above 290C, forming corrosive gas (acrolein).
Hazardous Polymerization:
Will not occur.
Incompatibilities:
Strong oxidizers. Can react violently with acetic anhydride, calcium oxychloride, chromium oxides and alkali metal hydrides.
Conditions to Avoid:
Heat, flames, ignition sources and incompatibles.

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11. Toxicological Information

Oral rat LD50: 12,600 mg/kg. Investigated as a mutagen, reproductive effector.

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--------\Cancer Lists\------------------------------------------------------
                                      ---NTP Carcinogen---
Ingredient                            Known    Anticipated    IARC Category
------------------------------------  -----    -----------    -------------
Glycerin (56-81-5)                     No          No             None
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12. Ecological Information

Environmental Fate:
When released into the soil, this material is expected to readily biodegrade. When released into the soil, this material is not expected to evaporate significantly. When released into water, this material is expected to readily biodegrade. This material is not expected to significantly bioaccumulate. When released into the air, this material may be moderately degraded by reaction with photochemically produced hydroxyl radicals. When released into the air, this material may be removed from the atmosphere to a moderate extent by wet deposition.
Environmental Toxicity:
This material is not expected to be toxic to aquatic life.

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13. Disposal Considerations

Whatever cannot be saved for recovery or recycling should be managed in an appropriate and approved waste disposal facility. Processing, use or contamination of this product may change the waste management options. State and local disposal regulations may differ from federal disposal regulations. Dispose of container and unused contents in accordance with federal, state and local requirements.

14. Transport Information

Not regulated.

15. Regulatory Information

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 --------\Chemical Inventory Status - Part 1\---------------------------------
 Ingredient                                       TSCA  EC   Japan  Australia
 -----------------------------------------------  ----  ---  -----  ---------
 Glycerin (56-81-5)                                Yes  Yes   Yes      Yes

 --------\Chemical Inventory Status - Part 2\---------------------------------
                                                        --Canada--
 Ingredient                                       Korea  DSL   NDSL  Phil.
 -----------------------------------------------  -----  ---   ----  -----
 Glycerin (56-81-5)                                Yes   Yes   No     Yes

 --------\Federal, State & International Regulations - Part 1\----------------
                                            -SARA 302-    ------SARA 313------
 Ingredient                                 RQ    TPQ     List  Chemical Catg.
 -----------------------------------------  ---   -----   ----  --------------
 Glycerin (56-81-5)                         No    No      No          No

 --------\Federal, State & International Regulations - Part 2\----------------
                                                       -RCRA-    -TSCA-
 Ingredient                                 CERCLA     261.33     8(d)
 -----------------------------------------  ------     ------    ------
 Glycerin (56-81-5)                         No         No         No


Chemical Weapons Convention:  No     TSCA 12(b):  No     CDTA:  No
SARA 311/312:  Acute: Yes      Chronic: Yes  Fire: No  Pressure: No
Reactivity: No          (Pure / Liquid)
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Australian Hazchem Code: None allocated.
Poison Schedule: None allocated.
WHMIS:
This MSDS has been prepared according to the hazard criteria of the Controlled Products Regulations (CPR) and the MSDS contains all of the information required by the CPR.

16. Other Information

NFPA Ratings: Health: **1** Flammability: **1** Reactivity: **0**
Label Hazard Warning:
CAUTION! MAY CAUSE IRRITATION TO SKIN, EYES, AND RESPIRATORY TRACT. MAY AFFECT KIDNEYS.
Label Precautions:
Avoid breathing mist.
Avoid contact with eyes, skin and clothing.
Keep container closed.
Use with adequate ventilation.

Wash thoroughly after handling.
Label First Aid:
If inhaled, remove to fresh air. Get medical attention for any breathing difficulty. In case of contact, immediately flush eyes or skin with plenty of water for at least 15 minutes. Get medical attention if irritation develops or persists.
Product Use:
Laboratory Reagent.
Revision Information:
No Changes.
Disclaimer:

Prepared by: Environmental Health & Safety
Phone Number: (314) 654-1600 (U.S.A.)

VIII. REFERENCES

[1] "CEP Magazine." SBE Supplement, Algal Biofuels: The Backstory, March, 2011 Issue. *AIChE Home Page*. Web. 10 Apr. 2011. <http://www.aiche.org/cep/>.

[2] "CEP Magazine." SBE Supplement, Algal Biofuels: The Process, March, 2011 Issue. *AIChE Home Page*. Web. 10 Apr. 2011. <http://www.aiche.org/cep/>.

[3] Rosenthal, Elisabeth. "Rush to Use Crops as Fuel Raises Food Prices and Hunger Fears." *The New York Times*. 6 Apr. 2011. Web. 8 Apr. 2011. <http://www.nytimes.com/2011/04/07/science/earth/07cassava.html >.

[4] CEP Magazine." SBE Supplement, Algal Biofuels: The Research, March, 2011 Issue. *AIChE Home Page*. Web. 10 Apr. 2011. <http://www.aiche.org/cep/>.

[5] Vaccari, D., "Phosphorous: A Looming Crisis*", Scientific American*, 300 (6) pp. 54-59 (June 2009).

[6] Carlson, L.S., Lee, M.Y., Oje, C.A.E., Xu, A., "*Algae to Alkanes*". University of Pennsylvania, Department of Chemical & Biomolecular Engineering, 2010.

[7] Carbon Dioxide Emissions from the Generation of Electric Power in the United States." *The Department of Energy* and *Environmental Protection Agency.* July 2000. Web. 5 April 2010.

[8] Yoo, C., Jun, S.Y., Lee, J.Y., Ahn, C.Y., Oh, H.M., 2009. Selection of microalgae for lipid production under high levels carbon dioxide. Bioresource Technology 101 (Suppl. 1), S71–S74.

[9] Wei Xiong, Chunfang Gao, Dong Yan, Chao Wu, Qingyu Wu. "*Double CO2 fixation in photosynthesis–fermentation model enhances algal lipid synthesis for biodiesel production*". Department of Biological Sciences and Biotechnology, Tsinghua University, Beijing 100084, PR China

[10] Pyper, P. G. and Weber, J. W., "*Culturing of Algae in Controlled Light-Dark Detention Systems: Effects of Culture Density*". Engineering Extension Series., Purdue University, 0079-8088

[11] Miao, X.L., Wu, Q.Y., 2006. Biodiesel production from heterotrophic microalgal oil. Bioresource Technology 97, 841–846.

[12] Van Baalen, C., Hoare, S.D., Brandt, E., "*Heterotrophic Growth of Blue-Green Algae in Dim Light*" The University of Texas, Marine Science Institute, Port Aransas, Texas 78373, and Microbiology Department

[13] Samejima, H., and J. Myers. 1958. On the heterotrophic growth of Chlorella pyrenoidosa. J. Gen. Microbiol. 18:107-117.

[14] Holmgren, Jennifer, Rich Marinangeli, and Terry Marker. "A New Development In Renewable Fuels: Green Diesel." (2007): 1-12. Print.

[15] "Eldorado Biofuels." Interview by David Schwartz. *Algae Industry Magazine* 29 Nov. 2010. Print.

[16] NREL/TP-580-24190: "*A Look Back at the U.S. Department of Energy's Aquatic Species Program: Biodiesel from Algae*". (July, 1998)

[17] Grobbelaar, U. Johan (1994), "Turbulence in mass algal cultures and the role of light/dark fluctuations", Department of Botany and Genetics, University of the OFS, Bloemfontein 9300, South Africa.

[18] http "Biofuels." *Home - BRAE Department - Cal Poly*. Web. 09 Apr. 2011. <http://brae.calpoly.edu/CEAE/biofuels.html>.

[19] Tredici, M,R, (1999), "Photobioreactors, In", Encyclopedia of Bioprocess Technology: Fermentation, Biocatalysis and Bioseparation (ed, M,C. Flickinger & S,W, Drew), pp, 395-419, j. Wiley & Sons, New York.

[20] Pienkos P. (2007),"The Potential for Biofuels from Algae", Algae Biomass Summit, National Renewable Energy Laboratory.

[21] Ugwu, C. U., H. Aoyagi, and H. Uchiyama. "Photobioreactors for Mass Cultivation of Algae." (2007)

[22] "Vertigro Algae Bio-Fuel Oil/C02 Sequestration System." *The Energy Blog*. Web. 03 Apr. 2011. <http://thefraserdomain.typepad.com/energy/2006/10/vertigro_algae_.html>.

[23] "EARTH Magazine: As Green As It Gets: Algae Biofuels." *EARTH Magazine - Voices: What Does "The Nation's Report Card" Tell Us about Science Education?* Web. 03 Apr. 2011. <http://www.earthmagazine.org/earth/article/1d6-7d9-2-d>.

[24] *Vertigro: Generating Biomass Rapidly*. Web. 03 Apr. 2011. <http://vertigro2day.blogspot.com/>

[25] "Green Is the New Yellow - Vertigro's Algae Farm." *JUICE: Alternate Fuels World*. Web. 03 Apr. 2011. <http://www.alternatefuelsworld.com/green-is-the-new-yellow>.

[26] Teymour, Fouad, Said Al-Hallaj, Aly-Eldeen ElTayeb, and Omar Khalil." "The Emerald Forest " – An Integrated Approach for Sustainable Community Development and Bio-derived Energy Generation." *Department of Chemical and Biological Engineering Illinois Institute of Technology*: 1-10. Print.

[27] Miao, X.L., Wu, Q.Y., 2004. High yield bio-oil production from fast pyrolysis by metabolic controlling of Chlorella protothecoides. Journal of Biotechnology 110, 85–93

[28] Xiufeng L., Xu H., Wu Q., "Large-Scale Biodiesel Production From Microalga Chlorella protothecoides Through Heterotrophic Cultivation in Bioreactors", Teymour, Fouad, Said Al-Hallaj, Aly-Eldeen ElTayeb, and Omar Khalil." "The Emerald Forest " – An Integrated Approach for Sustainable Community Development and Bio-derived Energy Generation." *Department of Chemical and Biological Engineering Illinois Institute of Technology*: 1-10. Print.

[29] *Diversified Energy*®. Web. 10 Apr. 2011. <http://www.diversified-energy.com/>.

[30] Simgae Algal Biomass Production System, http://www.diversified-energy.com/index.cfm?s_webAction=simgae

[31] Wikipedia contributors. "Greater Houston." *Wikipedia, The Free Encyclopedia*. Wikipedia, The Free Encyclopedia, 6 Apr. 2011. Web. 7 Apr. 2011.

[32] Aquifers of Texas (2001) Bureau of economic geology, John A. and Katherine G. Jackson school of Geosciences, The university of Texas at Austin, University Station.

[33] "Hydrogeology of the Edwards Aquifer." *The Edwards Aquifer Website*. Web. 09 Apr. 2011. <http://www.edwardsaquifer.net/geology.html>.

[34] *The Edwards Aquifer Website*. Web. 10 Apr. 2011. <http://www.edwardsaquifer.net/>.

[35] "Desalination Plant." *El Paso Water Utilities - Public Service Board*. Web. 03 Apr. 2011. <http://www.epwu.org/water/desal_info.html#>.

[36] "Infrastructure - Transportation - Gulf Coast - Texas In Focus." *Susan Combs - Texas Comptroller of Public Accounts*. Web. 07 Apr. 2011. <http://www.window.state.tx.us/specialrpt/tif/gulf/transportation.php>.

[37] "Final Texas Rail Plan." *Texas Department of Transportation Home Page*. Web. 10 Apr. 2011. <http://www.txdot.gov/public_involvement/rail_plan/trp.htm>.

[38] "W. A. Parish Electric Generating Station." *Texas-Flyer*. Web. 03 Apr. 2011. <http://www.texas-flyer.com/Fly-In-EngineOut/powerplant.htm>.

[39] Olaizola, M. et al. "Microalgal removal of CO2 from flue gases: CO2 capture from a coal combustor." Web. 20 Mar. 2010. <http://www.netl.doe.gov/publications/proceedings/04/carbon-seq/123.pdf>

[40] Solar Radiation Data Manual for Flat-Plate and Concentrating Collectors. NREL. Web. 4 Apr. 2010. <http://rredc.nrel.gov/solar/pubs/redbook/PDFs/TX.PDF>.

[41] Tilzer,M.M., "The Importance of Fractional Light Absorption by Photosynthetic Pigments for Phytoplankton Productivity in Lake Constance". The American Society of Limnology and Oceanography, 1983.

[42] Sirnoval, C. and Kandler, O. "Photoxidation processes in normal green *Chlorella* cells: I. The bleaching process". Botanical Institute of the University, Munich, Germany, December 2002.

[43] *Diversified Energy*®. Web. 03 Apr. 2011. <http://www.diversified-energy.com/index.cfm?s_webAction=simgae>.

[44] Cloud, G. et al., US2008/0311649 A1. "Pressurized Flexible Tubing System for Producing Algae." 18 Dec 2008. XL Renewables

[45] "Thermal Conductivity of Some Common Materials." Web. <http://www.engineeringtoolbox.com/thermal-conductivity-d_429.html>.

[46] Sirnoval, C. and Kandler, O. "Photoxidation processes in normal green *Chlorella* cells: I. The bleaching process". Botanical Institute of the University, Munich, Germany, December 2002.

[47] Ugwa, C.U., Aoyagi, H., Uchiyama, H. "Photobioreactors for mass cultivation of algae". Institute of Life Sciences and Bioengineering, University of Tsukuba, 1-1-1 Tennodai, Tsukuba City, Ibaraki 305-8572, Japan.

[48] Janssen, Marcel. "Cultivation of Microalgae: Effect of Light/dark Cycles on Biomass Yield." Web. <http://www.bpe.wur.nl/NR/rdonlyres/450C58DF-A0D2-4954-BB77-D0677325E981/65655/proefscriftMJanssen.pdf>.

[49] Andersen, Robert Arthur. "Algal Culturing Techniques." *Google Books*. Web. 10 Apr. 2011. <http://books.google.com/books?id=9NADUHyFZaEC>.

[50] Tamarys Heredia-Arroyo & Wei Wei & Bo Hu. Oil Accumulation via Heterotrophic/Mixotrophic Chlorella protothecoides, Appl Biochem Biotechnol (2010)

[51] Campbell, D. "Intracellular Calcium Regulation Edited by Bader. J., Geitzen, K., Rosenthal, J., Rudel, R. and Wolf, H.U. (editors) 1986 Manchester University Press." *Cell Calcium* 8.2 (1987): 185-86. Print.

[52] Blanch, Harvey W. and Clark, Douglas S. Biochemical Engineering. 1st ed. New York, Marcel Dekker, 1997. 720 p. ISBN0-82-47 0099-6.

[53] Andersen, Robert A. *Algal Culturing Techniques*. Burlington, MA: Elsevier/Academic, 2005. Print.

[54] Blanch, Harvey W. and Clark, Douglas S. Biochemical Engineering. 1st ed. New York, Marcel Dekker, 1997. 720 p. ISBN0-82-47 0099-6.

[55] "Solubility of Gases in Water." Web. <http://www.engineeringtoolbox.com/gases-solubility-water-d_1148.html>.

[56] Corman et al. 1957, Brown 1970, Wang et al. 1979, Moo-Young and Blanch 1987, Blanch and Clark 1997)

[57] Web. 03 Apr. 2011. <http://www.ncbi.nlm.nih.gov/pubmed/2178781>.

[58] Tamarys Heredia-Arroyo & Wei Wei & Bo Hu, "Oil Accumulation via Heterotrophic/Mixotrophic Chlorella protothecoides". Appl Biochem Biotechnol (2010)

[59] Stensel, David H., and Franklin L. Burton. *Wastewater Engineering; Treatment and Reuse.* Boston: McGraw-Hill, 2003. Print"

[60] Buehner, Michael R., et al."*Microalgae Growth Modeling and Control for a Vertical Flat Panel Photobioreactor.*"2009 American Control Conference. 20 Mar. 2010

[61] "Henry's Law - Encyclopedia Article - Citizendium." *Welcome to Citizendium - Citizendium*. Web. 09 Apr. 2011. <http://en.citizendium.org/wiki/Henry's_law>.

[62]"W. A. Parish Electric Generating Station." *Texas-Flyer*. Web. 03 Apr. 2011. <http://www.texas-flyer.com/Fly-In-EngineOut/powerplant.htm>.

[63] "Growing MicroAlgae - NRAC." *Learning, Discovery, Service | in the College of Agriculture*. Web. 03 Apr. 2011. <http://www.ca.uky.edu/wkrec/AlgaeGrowNRAC-160.htm>.

[64] Tamarys Heredia-Arroyo & Wei Wei & Bo Hu. Oil Accumulation via Heterotrophic/Mixotrophic Chlorella protothecoides, Appl Biochem Biotechnol (2010)

[65] http://rothfus.cheme.cmu.edu/tlab/ferm/projects/t6_s99/t6_s99_manual.PDF

[66] "Alcoholic Fermentation." Web. <http://www.nysaes.cornell.edu/fst/faculty/acree/fs430/notes_thk/pdf/fermentation.pdf>.

[67] Seider, Warren D., Daniel R. Lewin, and J. D. Seader. *Process Design Principles: Synthesis, Analysis and Evaluation*. New York , Wiley, 1999. Print.

[68] "OriginOil CEO Addresses Algae Fuels Conference at World Biofuels Markets | OriginOil." *OriginOil - Breakthrough Algae Oil Technology*. Web. 10 Apr. 2011. <http://www.originoil.com/company-news/originoil-ceo-addresses-algae-fuels-conference-at-world-biofuels-markets.html>.

[69] *Flocculants | Polymers | Coagulants*. Web. 10 Apr. 2011. <http://www.flocculants.info/>.

[70] "Hydrocyclone Separator." *Industrial Air, Wastewater & Sewage Treatment Systems – Ecologix Environmental Systems*. Web. 10 Apr. 2011. <http://www.ecologixsystems.com/hydrocyclone_separator.php>.

[71] "Fluid Bed Drying - Fundamentals." *GEA Process Engineering Inc. - A.k.a. Niro Inc.* Web. 10 Apr. 2011. <http://www.niroinc.com/drying_dairy_food/fluid_bed_technology.asp>.

[72] *Industrial Microwave Systems, Inc. - Microwave Heating, Drying, Bonding,curing, Cooking, Pasteurizing and Sterilizing Equipment for Continuous Industrialprocessing of Wide Webs, Conveyed Materials and Materials Pumped through a Pipe*. Web. 10 Apr. 2011. <http://www.industrialmicrowave.com/>.

[73] "Expeller Press - Definition, Glossary, Details - Oilgae." *Biodiesel from Algae Oil - Oilgae - Information, News, Links for Algal Fuel, Alga Bio-diesel, Biofuels, Algae Biofuel, Energy - Oilgae.com*. Web. 10 Apr. 2011. <http://www.oilgae.com/ref/glos/expeller_press.html>

[74] "Solvent Extraction - Solvent Extractors." *Methanol Recovery, Biodiesel Distillation, Glycerin Purification, High FFA Pretreatment, Reactor*. Web. 10 Apr. 2011. <http://www.srsbiodiesel.com/SolventExtraction.aspx>.

[75] Mollah, M.Y.A., et al., "Fundamentals, Present and Future Perspectives of Electrocoagulation,' *Journal of Hazardous Materials*, 114, pp. 199-210 (2004)

[76] Andrich, G., et al., "Supercritical Fluid Extraction of Bioactive Lipids from the Microalga *Nannochloropsis sp*., *European Journal of Lipid Science and Technology,* 107, pp.31-36 (2005)

[77] Jessop, P.G., et al., "Reversilbe Non Polar-to-Polar Solvent," *Nature,* 426 , p.1102 (2005)

[78] "Single-Step Extraction." *OriginOil - Breakthrough Algae Oil Technology*. Web. 10 Apr. 2011. <http://www.originoil.com/technology/single-step-extraction.html>.

[79] Eckelberry, N. and Eckelberry, R.T., "Algae Growth Systems for Oil Production" (Quantum Fracturing Patent Application), (2009).

[80] Lawrence K. Wang, Nazih K. Shammas, and Yung-Tse Hung , Biosolid Treatment Processes, Volume 6, Handbook of Environmental Engineering, A Humana Press product.

[81] OriginOil presentation at the World Biofuel Market Conference, 15-17 March 2010.

[82]"EIA - Biofuels in the US Transportation Sector." Web. 10 Apr. 2011. <http://www.eia.doe.gov/oiaf/analysispaper/biomass.html>.

[83] “What Do We Feed to Food-Production Animals? A Review of Animal Feed Ingredients and Their Potential Impacts on Human Health” Amy R. Sapkota, Lisa Y. Lefferts, Shawn McKenzie, and Polly Walker

[84] Knothe, Gerhard, Jurgen Krahl, and Gerpen Jon Harlan. Van. *The Biodiesel Handbook*. Urbana, IL: AOCS, 2009. Print.

[85] "Glycerol (chemical Compound) -- Britannica Online Encyclopedia." *Encyclopedia - Britannica Online Encyclopedia*. Web. 04 Apr. 2011. <http://www.britannica.com/EBchecked/topic/236029/glycerol>.

[86] Huber, George W., Paul O’Connor, and Avelino Corma. "Processing Biomass in Conventional Oil Refineries: Production of High Quality Diesel by Hydrotreating Vegetable Oils in Heavy Vacuum Oil Mixtures." *Applied Catalysis A: General* 329 (2007): 120-29. Print.

[87] Tamarys Heredia-Arroyo & Wei Wei & Bo Hu. Oil Accumulation via Heterotrophic/Mixotrophic Chlorella protothecoides, Appl Biochem Biotechnol (2010)

[88] Leevijit, T., C. Tongurai, G. Prateepchaikul, and W. Wisutmethangoon. "Performance Test of a 6-stage Continuous Reactor for Palm Methyl Ester Production 2606." *Bioresource Technology* 99.1 (2008): 214-21. Print.

[89] Chang, Ai-Fu and Y.A. Liu. “Integrated Process Modeling and Product Design of Biodiesel Manufacturing.” *Ind. Eng. Chem. Res.* 49. (2010): 1197-1213. Print.

[90] Stiefel, Scott and Gustavo Dassori. “ Simulation of Biodiesel Production Through Transesterification of Vegetable Oils.: *Ind. Eng. Chem. Res.* 48 (2009): 1068-1071. Print.

[91] Pokoo-Aikins, Grace, Ahmed Nadim, Mahmou M. El-Halwagi and Vladimir Mahalec. “Design and analysis of biodiesel production from algae grown through carbon sequestration.” *Clean Techn Environ Policy*. 12 (2010): 239-254. Print.

[92] “Properties of Saturated Steam – Imperial Units.” *Engineering ToolBox*. Web 04 Apr. 2011. <http://www.engineeringtoolbox.com/saturated-steam-properties-d_273.html>

[93] "Biodiesel (USA) Price Report - Chemical Pricing Information - ICIS Pricing." *Chemical Prices and Chemical Industry Trends from ICIS Pricing*. Web. 04 Apr. 2011. <http://www.icispricing.com/il_shared/Samples/SubPage10100131.asp>.

[94] “Methanol – Material Safety Data Sheet.” *MIDI-Inc.* Web 04 Apr. 2011. <*www.midi-inc.com/pdf/MSDS_Methanol.pdf*>

[95] “Potassium Hydroxide – Material Safety Data Sheet.” *JT Baker* Web 04 Apr. 2011. <*www.jtbaker.com/msds/englishhtml/p5887.htm*>

[96] “Hydrochloric Acid – Material Safety Data Sheet.” *JT Baker* Web 04 Apr. 2011. <*www.jtbaker.com/msds/englishhtml/h3880.htm*>

[97] *CCX - Chicago Climate Exchange*. 08 Apr. 2011. Web. 10 Apr. 2011. <http://www.chicagoclimatex.com/>.

[98] "Carbon Market News." *Albert Holmes Limited / Ethical Investment Consultants* Issue 1 (20 Aug. 2010): Http://www.albertholmes.com. Print.

[99] "The Material Balance for Chemical Reactors." Web. <http://jbrwww.che.wisc.edu/home/jbraw/chemreacfun/ch4/slides-matbal.pdf>.

[100] "The Material Balance for Chemical Reactors." Lecture. Copyright C 2011 by Nob Hill Publishing, LLC. Web. <http://http://jbrwww.che.wisc.edu/home/jbraw/chemreacfun/ch4/slides-matbal.pdf>.

[101] Frank P. Incropera, David P. DeWitt - John Wiley. "Fundamentals of heat and mass transfer (2007)

[102] Solar Radiation Data Manual for Flat-Plate and Concentrating Collectors. NREL. Web. 4 Apr.

[103] "Reform of Biodiesel Tax Incentives." US Congress. Web 04 Apr 2011. <http://www.biodiesel.org/news/taxincentive/Biodiesel%20Tax%20Incentive%20Reform%20and%20Extension%20Act%20of%202009.pdf>

[104] *Google Maps*. Web. 11 Apr. 2011. <http://maps.google.com/maps?hl=en>.

[105] O' Grady, John and John A Morgan. "Heterotrophic Growth and Lipid Production of Chlorella protothecoides on Glycerol." Bioprocess Biosyst Eng (2011) 34:121–125

[106] Eckelberry, N. and Eckelberry, R.T., "Systems, Apparaturs and Methods for Obtaining Intracellular Products and Cellular Mass and Debris from Algae and Derivative Products and Process of Use Thereof" (Single Step Extraction Patent Application), (2010).

[107] "Hydrochloric Acid Pricing." *Alibaba.com* Web. 04 Apr 2011. <http://www.alibaba.com/showroom/hydrochloric-acid-34%2525.html

[108] "Methanex - Methanol Price." *Methanex - the Global Leader in Methanol Production and Marketing*. Web. 11 Apr. 2011. <http://www.methanex.com/products/methanolprice.html>.

[109] "Potassium Hydroxide Pricing." *Alibaba.com* Web. 04 Apr 2011. <http://www.alibaba.com/showroom/potassium-hydroxide-45%2525.html>

[110]"Biodiesel.org - Tax Credit Extension Action Center." *National Biodiesel Board - Www.biodiesel.org - Www.nbb.org*. Web. 11 Apr. 2011. <http://www.biodiesel.org/news/taxcredit/default.shtm>.