Production of Biodiesel from Algae

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June, 2015

Acknowledgement

Commencing with the name of the Al-Mighty, Allah, the All-Knowing, the most beneficent and the most merciful, we thank our Lord for giving us the capabilities, the energies and the strengths to carry out and undertake a project of this nature and importance as our undergraduate thesis. We would sincerely like to thank our supervisors, Dr. Arshad Chughtai and Dr. Habib Nasir, for guiding us, assisting us and mentoring us throughout this project with their valuable insight on the subject, and technical expertise in this area. We will also like to thank Mr.Ameer Ali from Pakistan State Oil (PSO) for guiding us throughout the project Also for being a friend and a confidante throughout the year, boosting our morale when the chips were down and helping us through good times and bad. Our sincere thanks are also devoted to Dr. Arshad Hussain, who allowed us to take up this project despite the concerns of finding a supervisor and choosing a project of this nature. His constant motivation and encouragement has kept us going throughout the term of this project thesis. Last but not the least, we would not hesitate to devote our thanks to Mr. Muneeb Qayyum, Mr. Kamal Mustafa for helping us in queries no one else could at that time, his critical approach and crucial knowledge has also contributed towards the completion of this project.

Dedication

This project and this thesis we will first dedicated firstly to our beloved parents, without whom none of this would have been possible. We also thank SCME and the entire faculty for supporting us throughout the entirety of the project, specially our Supervisors Dr. Arshad Chughtai, Dr.Habib Nasir and HoD, Dr. Arshad Hussain.

Certificate

This is to certify that work in this thesis has been carried out by **Mr. Ahmed Zulfiqar, Mr. Mufaddal Azhar and Ms. Sehr Karimjee** and completed under supervision of **Dr. Arshad Chughtai and Dr. Habib Nasir** in School of Chemical and Materials Engineering (SCME), National University of Sciences and Technology, H-12, Islamabad, Pakistan.

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Nomenclature

[S] solute concentration in solvent

$\propto$ relative volatility

A cross sectional area for membrane/reactor/tanks

A surface area for heat exchangers

aq aqueous solvent

B bottoms

BDC bundle diameter clearance

B_s baffle spacing

C_A^* concentration in solution

c_p specific heat capacity

C_x cell density

d diameter of the plate (sparger)

D diameter or reactor/column/tank

D dilution constant (lipid extraction unit)

D dilution rate (bioreactor)

D distillate (distillation column)

D_b baffle spacing

d_b bubble diameter

d_c column diameter

D_l mass flux

E plate efficiency

F flow rate

F moles of feed

F_T correction factor

g gravitational constant

G mass velocity

h heat transfer coefficient

H height of column

H height of reactor/column/tank

H Henry constant

I light intensity

I pore size

I wall thickness

J mass flux

k_l growth rate

k_la diffusion coefficient

L volumetric flow rate (decanter)

LMTD Log Mean Temperature Difference

m mass flow rate

MR molecular mass

N number of plates

N number of tubes (heat exchanger)

NTU Number of Transfer Units

org organic solvent

P permeability

P pressure

Q heat transfer (heat exchangers)

Q volumetric flow rate

R correction factor

R gas constant (distillation)

r radius

R reflux ratio

S correction factor

T temperature

U overall heat transfer coefficient

u velocity

u_{gs} superficial gas velocity

V volume of reactor

V_t terminal velocity

W weight

x biomass concentration

Y_x yield

ΔH change in enthalpy

ΔT temperature change

ε_g gas hold up

η_l viscosity of liquid

μ viscosity

ρ density

σ surface tension

τ residence time

Abstract

Biodiesel is a diesel fuel derived from animal or plant lipids (oils and fats). With the need to reduce carbon emissions, and the dwindling reserves of crude oil, liquid fuels derived from plant material – biofuels – are an attractive source of energy. Biodiesel produced from algae appears to be a feasible solution today for replacing petro-diesel completely. No other feedstock has the oil yield high enough for it to be in a position to produce such large volumes of oil. Studies have shown that some species of algae can produce 60% or more of their dry weight in the form of oil. Because the cells grow in aqueous suspension, where they have more efficient access to water, CO2 and dissolved nutrients, microalgae are capable of producing large amounts of biomass and usable oil in either high rate algal ponds or photo-bioreactors. This oil can then be turned into biodiesel which could be sold for use in automobiles. Also, regional production of microalgae and processing into biofuels will provide economic benefits to rural communities.

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Introduction

Problem statement:

Due to declining fossil fuel reserves, global climate changes and government regulation regarding Green House Gases; the alternate fuel sources available nowadays have become more and more important. Although there are a number of different fuel sources available but biodiesel is amongst most feasible options available.

PSO has selected algae as a potential oil producing source for bio-diesel because of its unique characteristics and for the same reason substantial research work has been done by countries like India, Malaysia, China, Brazil and USA.

The Government of Pakistan in association with Pakistan State Oil has pledged to replace 5 percent of annual diesel demand by producing biodiesel.

Purpose of the study

Production of biodiesel from algae is being studied all over the world but its practical implementation is hindered by the lack of continuous supply of algae having high lipid content and a development of a process that continuously converts algal oil into biodiesel to make it industrially feasible. The purpose of our study is to devise a process for continuous harvesting of algae and continuous conversion of oil into biodiesel with process efficiency and low residence time

Contribution of the Study

We have successfully addressed the two major problems faced by the algal biodiesel industry which are

- Continuous harvesting of algae

- Efficient conversion of algal oil into biodiesel in minimum time

Using results from our study a continuous process can be established that will allow fast growth of algae in a closed environment and continuous conversion of algal oil into biodiesel. currently the process of converting the oil into biodiesel is a batch one that takes about 2 hours where as our designed process converts the oil into biodiesel in under a minute.

Why biodiesel?

Biodiesel is produced from vegetable oil or animal fat. These resources can be regrown and considered infinite in comparison to crude oil that is limited. When using biodiesel instead of petro diesel, the demand for crude oil products is reduced and as a result their prices decrease.

While using biodiesel the exhaust emission is significantly less polluting. According to studies, there are several parameters like hydrocarbons and carbon dioxide that are reduced when using biodiesel for running diesel engines.

Experience shows that engines running on biodiesel, its lubricity is improved compared to petro diesel fuel. The higher lubricity enables the engine to work in a smoother manner and because it works better it reduce the physical deterioration of the engine.

Each alternative to petro diesel fuel requires adjustments and modifications, sometimes it is in the fuel injection system and sometimes another tank should be installed. The only exception is biodiesel. The majority of diesel engines will work smoothly with biodiesel with no modifications requirements. Another advantage of biodiesel over the other energy alternatives is the mileage per gallon. It is important to mention about the little drop in power when compared to petro diesel. Biodiesel has a drop in power of around 9%, which most drivers not even notice or notice a non-significant drop in power.

Compared to standard diesel fuel, biodiesel fuel has lower energy content per volume. This means that a driver who switches to biodiesel fuel will see a significant loss of engine power, up to 11 percent according to a recent assessment by Land of the Free. To prevent the loss of power, fuel makers often blend biodiesel fuel with conventional diesel, which also dilutes its environmental advantages

While using biodiesel the exhaust emission is significantly less polluting. According to studies, there are several parameters like hydrocarbons and carbon dioxide that are reduced when using biodiesel for running diesel engines. Biodiesel reduces nearly all forms of air pollution. Most importantly, biodiesel reduces air toxics and cancer-causing compounds. Pure biodiesel can reduce the cancer risks by 94%; B20 will reduce that risk by as much as 27%.There is no sulfur in biodiesel, so biodiesel won‘t contribute to sulfur dioxide emissions or poison exhaust catalysts. B20 has 20% of the benefits of pure biodiesel.

Biodiesel emissions are less harmful to the environment compared to petro diesel. The comparison of which is given below.

Explanation of Emission Types

- Particulate Matter (Black Smoke)
  Emissions of particulate matter have been linked to respiratory diseases and are generally considered to be a human health hazard. Emissions of particulate matter are reduced with biodiesel by 47%.

- Carbon Monoxide
  Carbon Monoxide is a poisonous gas. Reduced with biodiesel by 48%.

- Total Unburned Hydrocarbons
  Compounds which contribute to localized formation of smog. Reduced with biodiesel by 67%.

- Nitrogen Oxides
  Compounds which contribute to localized formation of smog. According to NREL (Biodiesel Handling and Use Guidelines 2009 revision) examination of the NOx testing results shows that the effect of biodiesel can vary with engine design, calibration, and test cycle. At this time, the data are insufficient for users to

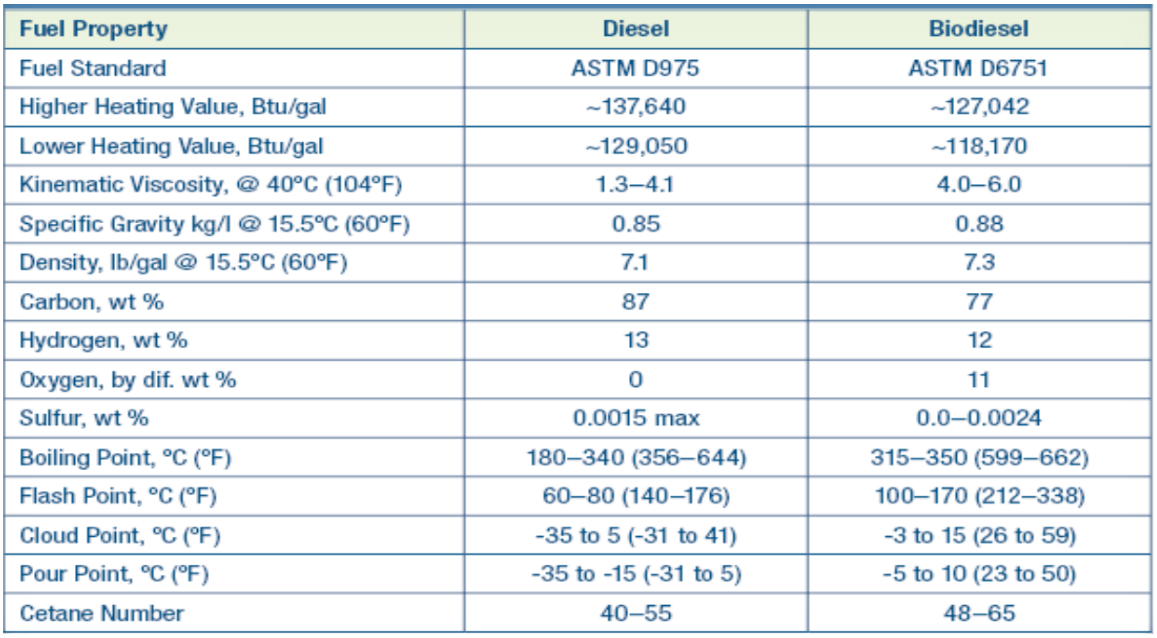
conclude anything about the average effect of B20 on NOx, other than that it is likely very close to zero.

- Sulfates
  Sulfates are major contributors to acid rain. These emissions are practically eliminated when using biodiesel.
- Polycyclic Aromatic Hydrocarbons (PAH and nPAH)
  These compounds have been identified as carcinogenic (cancer causing) compounds. Biodiesel reduces emissions of these compounds by up to 85% for PAH compounds and 90% for nPAH compounds.
- Speciated Hydrocarbons
  These compounds contribute to the formation of localized smog and ozone. The potential for smog formation from speciated hydrocarbons is reduced by 50% when using biodiesel.
- Life Cycle Reduction of CO2
  Biodiesel helps reduce the risk of global warming by reducing net carbon emissions to the atmosphere. When biodiesel is burned, it releases carbon dioxide to the atmosphere, but crops which are used to produce biodiesel take up carbon dioxide from the atmosphere in their growth cycle.

Biodiesel has much higher cetane ratings than today's lower sulfur diesel fuels. It has been claimed biodiesel gives better lubricity and more complete combustion thus increasing the engine energy output and partially compensating for the higher energy density of petro diesel

With the ability to run in any unmodified diesel engine, pure bio diesel can also help increase the life of your diesel engine because it leaves no deposits. The life of your diesel engine is increased by bio diesel due to the fact that it is more lubricating than traditional petroleum diesel products. Fuel consumption, ignition, engine power and torque remain unaffected by the use of bio diesel.

Table 1: Comparison of Petrodiesel and Biodiesel



Fuel Property	Diesel	Biodiesel
Fuel Standard	ASTM D975	ASTM D6751
Higher Heating Value, Btu/gal	~137,640	~127,042
Lower Heating Value, Btu/gal	~129,050	~118,170
Kinematic Viscosity, @ 40°C (104°F)	1.3–4.1	4.0–6.0
Specific Gravity kg/l @ 15.5°C (60°F)	0.85	0.88
Density, lb/gal @ 15.5°C (60°F)	7.1	7.3
Carbon, wt %	87	77
Hydrogen, wt %	13	12
Oxygen, by dif. wt %	0	11
Sulfur, wt %	0.0015 max	0.0–0.0024
Boiling Point, °C (°F)	180–340 (356–644)	315–350 (599–662)
Flash Point, °C (°F)	60–80 (140–176)	100–170 (212–338)
Cloud Point, °C (°F)	-35 to 5 (-31 to 41)	-3 to 15 (26 to 59)
Pour Point, °C (°F)	-35 to -15 (-31 to 5)	-5 to 10 (23 to 50)
Cetane Number	40–55	48–65

Current Situation

Energy Demand

Renewable energy is generally defined as energy that comes from resources that are not significantly depleted by their use, such as sunlight, wind, rain, tides, waves and geothermal heat. Renewable energy is gradually replacing conventional fuels in four distinct areas: electricity generation, hot water/space heating, motor fuels, and rural (off-grid) energy services.

Based on REN21's (Renewable Energy Policy Network for the 21st Century) 2014 report, renewables contributed 19 percent to our energy consumption and 22 percent to our electricity generation in 2012 and 2013, respectively. Both, modern renewables, such as hydro, wind and biofuels, as well as traditional biomass, contributed in about equal amount to the global energy supply. Worldwide investments in renewable technologies amounted to more than US$214 billion in 2013, with countries like China and the United States heavily investing in wind, hydro, solar and biofuels. Renewable energy resources exist over wide geographical areas, in contrast to other energy sources, which are concentrated in a limited number of countries. Rapid deployment of renewable energy and energy efficiency is resulting in significant energy security, climate change mitigation, and economic benefits. [51] In international public opinion surveys there is strong support for promoting renewable sources such as solar power and wind power

World final energy consumption refers to the fraction of the energy that the world consumes that is used by humanity. In 2012, world energy consumption amounted to 155,505 terawatt-hours (TWh), while the world final energy consumption was 104,426 TWh or about 32% less than the world energy consumption. World electricity generation refers to the fraction of the final world energy consumption that is used as electricity. The world's electricity generation is smaller than the final energy consumed because energy is used for many other things then electricity. For instance, energy must be used in industry to power machines that manufacture goods, and energy must be used in the form of gasoline to power vehicles. In 2012, the world's electricity generation was 19,422 TW*h. In 2008, total worldwide energy consumption was 132,000 terawatt-hours (TWh) or 474 exajoules (EJ). This is corresponds to an average global power demand of 15 terawatts (TW). In 2012, energy demand increased to 158,000 TWh (567 EJ), equivalent to an average power use of 18.0 TW. Based upon some attempted estimates, making strong assumptions, the annual potentials for renewable fuels are of the order of:

solar energy 1575 - 49387 EJ (437,500 - 13,718,611 TW*h),
wind power 640 EJ (180,000 TWh),
geothermal energy 5,000 EJ (1,400,000 TWh),
biomass 276 EJ (77,000 TWh),
hydropower 50 EJ (14,000 TWh) and
ocean energy 1 EJ (280 TWh) (*World Energy Assessment-Newyork 2008*)

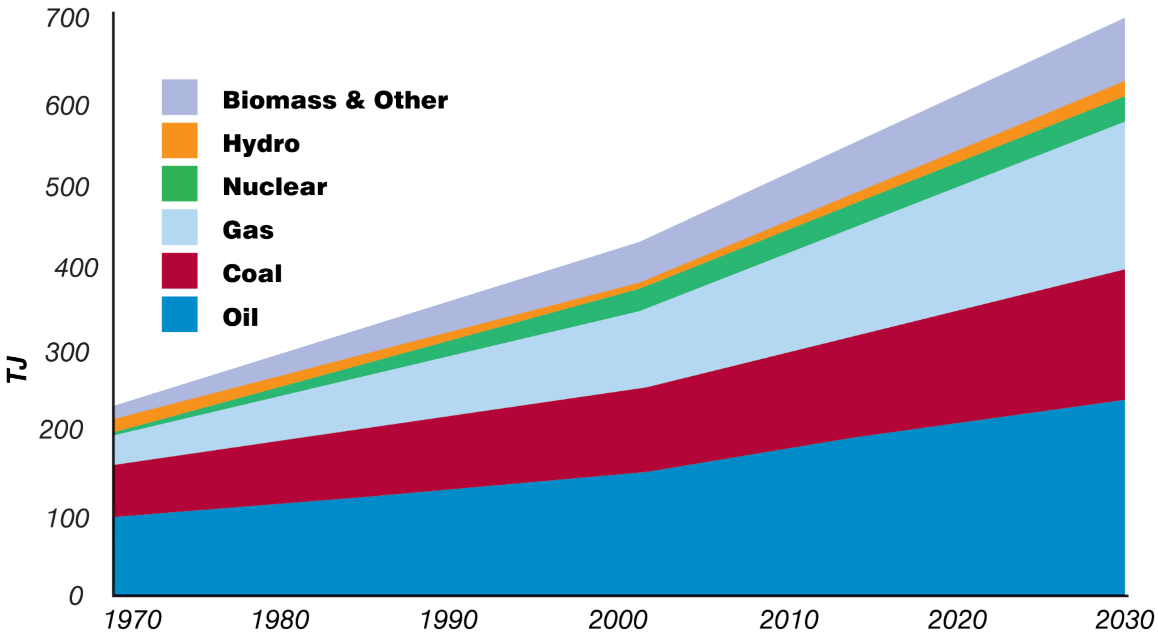


Figure 1: Energy Reserves

(*Primary Energy Demand Projection Report*)

World Oil Reserves by Region

Energy demand with respect to fuel

Catalysis for renewable

At present much of the energy requirement is fulfilled by fossil fuel (oil). This is the situation of world oil reserves by region. Pakistan lies in the region having only 3% of the total world‘s oil reserves. Middle east as is popular dominates the current world oil reserves with almost 4 times as much as North America‘s where Venezuela is extremely rich in oil.

Oil Peak

The Oil Peak, according to a recent prognosis of the Association for the Study of Peak Oil says that US, Russian and European oil supplies have 'peaked' already years ago. After 2008, the global crisis kicks in.

The real problem, of course, is that our society is addicted to cheap, abundant oil. Our cars, airplanes and ships run on oil. Our electricity is generated in oil-fueled power plants. The stuff you buy in the stores is brought there by trucks that run on oil. In the end, literally everything is to some degree made of oil! So when oil gets expensive, so does our world. Even seemingly innocent things like socks, drinking water and bread will become very expensive. Factories and businesses will go bankrupt. Unemployment will explode, pushing up the state deficit and deepening the crisis even more. Banks will shut down, thereby killing the savings of their clients. In the end, the Oil Peak will send massive shockwaves through the world‘s economies.

Growth of Industrial Economy-increase in oil demand

On the other hand, as economies of the developing countries grow, the thrust for oil will grow more. This thrust means an increase in demand of oil.

World energy demand-related carbon dioxide emissions

World energy demand-related carbon dioxide emissions rise from 29.7 billion metric tons in 2007 to 33.8 billion metric tons in 2020 and 42.4 billion metric tons in 2035—an increase of 43 percent over the projection period. With strong economic growth and continued heavy reliance on fossil fuels expected for most non- OECD economies under current policies, much of the projected increase in carbon dioxide emissions occurs among the developing non-OECD nations. In 2007, non-OECD emissions exceeded OECD emissions by 17 percent; in 2035, they are projected to be double OECD emissions.

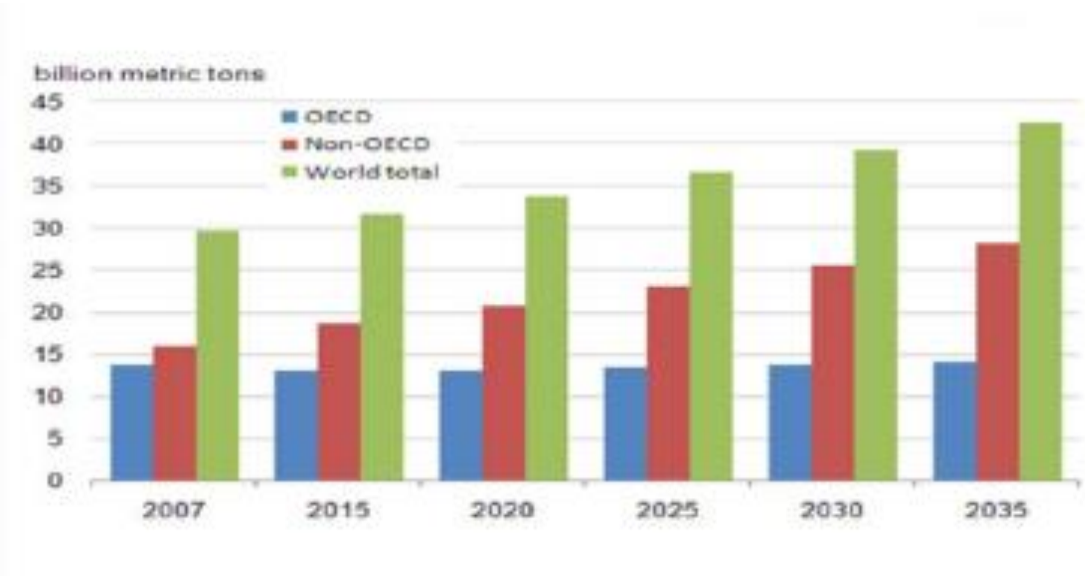


Figure 2: Carbon Dioxide Emissions

Three major problems to be encountered:

- Depletion of world oil reserves
- Continuous hike in international oil price
- Threatening Carbon dioxide emissions

Along with that, the following factors have paved the way for extensive research and development for alternate energy sources:

- The reserves of fossil fuels are declining
- Concerns over security of assured supplies
- Ever-increasing prices of petroleum products
- Environment effects

Types of biofuels

- Bio alcohols
- Biodiesel
- Vegetable oil
- Biogas
- Solid biofuels

Bio alcohols: Biologically produced alcohols; most common is Ethanol; Alcohol fuels are produced by fermentation of sugars derived from wheat, corn, sugar beets, sugar cane, molasses and any sugar or starch that alcoholic beverages can be made from (like potato and fruit waste, etc.).

Biodiesel: It is produced from oils or fats using transesterification and is a liquid similar in composition to fossil/mineral diesel. Chemically, it consists mostly of fatty acid methyl (or ethyl) esters (FAMEs).

Vegetable Oil: Used vegetable oil is increasingly being processed into biodiesel, or (more rarely) cleaned of water and particulates and used as a fuel.

Biogas: Biogas is methane produced by the process of anaerobic digestion of organic material by anaerobic organism.

Solid biofuels include wood, saw dust, grass cuttings, agricultural waste, charcoal, dried manure etc.

Why Algae?

The algae that are used in biodiesel production are usually aquatic unicellular green algae. This type of algae is a photosynthetic eukaryote characterized by high growth rates and high population densities. Under good conditions, green algae can double its biomass in less than 24 hours. Additionally, green algae can have huge lipid contents, frequently over 50%. This high yield, high density biomass is ideal for intensive agriculture and may be an excellent source for biodiesel production.

The annual productivity and oil content of algae is far greater than seed crops. Soybean can only produce about 450 l of oil per hectare. Canola can produce 1200 l per hectare, and palm can produce 6000 l. Now, compare that to algae which can yield 90,000 l per hectare. It is possible that the US demand for liquid fuel could be achieved by cultivating algae in one tenth of the area currently devoted to soybean cultivation. (*Richard Gordon 2004*)

The process for producing micro algal oils consists of a micro algal biomass production step that requires light, carbon dioxide, water and inorganic nutrients. The latter are mainly nitrates, phosphates, iron and some trace elements. Approximately half of the dry weight of the micro algal biomass is carbon, which is typically derived from carbon dioxide. Therefore, producing 100 tons of algal biomass fixes roughly 183 tons of carbon dioxide. This carbon dioxide must be fed continually during daylight hours. This carbon dioxide is often available at little. Optimal temperature for growing many microalgae is between 293 and 303 K. A temperature outside this range could kill or otherwise damage the cells. This makes the harvesting process cost efficient. (*Ayhan Demirbas 2006*)

Biodiesel from palm oil costs roughly $0.66/L or 35% more than petro diesel. This suggests that the process of converting palm oil to biodiesel adds about $0.14/L to the price of oil. For palm oil sourced biodiesel to be competitive with petro diesel, the price of palm oil should not exceed $0.48/L, assuming an absence of tax on biodiesel. Using the same analogy, a reasonable target price for micro algal oil is $0.48/L for algal diesel to be cost competitive with petro diesel. (*Yusuf Chisti 2007*)

Producing biodiesel from algae has been touted as the most efficient way to make biodiesel fuel. Algal-oil processes into biodiesel as easily as oil derived from land-based crops. The difficulties in efficient biodiesel production from algae lie not in the extraction of the oil, but in finding an algal strain with a high lipid content and fast growth rate that is not too difficult to harvest, and a cost-effective cultivation system (i.e. type of photo-bioreactor) that is best suited to that strain. Algae are the fastest-growing plants in the world. Microalgae have much faster growth-rates than terrestrial crops. The per unit area yield of oil from algae is estimated to be from 18,927 to 75,708 liters per acre, per year; this is 7–31 times greater than the next best crop, palm oil 2404 liters.(*Biofuel from algae 2012*)

Major advantages of using algae as oil source:

Algae Grow Fast

Algae can double their numbers every few hours, can be harvested daily, and have the potential to produce a volume of biomass and biofuel many times greater than that of our most productive crops.

Algae Can Have High Biofuel Yields

Algae store energy in the form of oils and carbohydrates, which, combined with their high productivity, means they can produce from 2,000 to as many as 5,000 gallons of biofuels per acre per year.

Algae Consume CO2

Like any other plant, algae, when grown using sunlight, consume (or absorb) carbon dioxide (CO2) as they grow, releasing oxygen (O2) for the rest of us to breathe. For high productivity, algae require more CO2, which can be supplied by emissions sources such as power plants, ethanol facilities, and other sources.

Algae Do Not Compete With Agriculture

Algae cultivation uses both land that in many cases is unsuitable for traditional agriculture, as well as water sources that are not useable for other crops, such as sea-, brackish- and wastewater. As such, algae-based fuels complement biofuels made from traditional agricultural processes.

Micro algal Biomass Can Be Used for Fuel, Feed and Food

Microalgae can be cultivated to have a high protein and oil content, for example, which can be used to produce either biofuels or animal feeds, or both. In addition, micro algal biomass, which is rich in micronutrients, is already used for dietary supplements to advance human health.

Macro algae Can Be Grown in the Sea

Macro algae (seaweeds) are grown in the sea, or even on land with seawater, and their sugars can be converted into biofuels and chemicals.

Algae Can Purify Wastewaters

Algae thrive in nutrient-rich waters like municipal waste waters (sewage), animal wastes and some industrial effluents, at the same time purifying these wastes while producing a biomass suitable for biofuels production.

Algal Biomass Can Be Used as an Energy Source

After oil extraction, the remaining algal biomass can be dried and "pelletized" and used as fuel that is burned in industrial boilers and other power generation sources.

Algae Can Be Used to Produce Many Useful Products

Algae can be cultivated to produce a variety of products for large to small markets: plastics, chemical feedstock, lubricants, fertilizers, and even cosmetics.

Microalgae Comparison to Terrestrial Crops

There is a general consensus that the production of biofuels from oilseed crops and waste /fats cannot meet the worldwide demand for transportation fuels 10 (Tyson et al., 2004). In 2009, the U.S. used approximately 206 billion liters of petroleum diesel (http://eia.doe.gov). In 2009, U.S. biodiesel production decreased from to 2.56 billion liters to 1.85 billion liters (Biodiesel Magazine). Even though soy oil production for 2008-2009 totaled 9.3 billion liters, more than 90 percent of this oil is used in the food products market thereby severely limiting its use as a biofuel feedstock (Iowa State University). As of 2000, the U.S. produced in excess of 10.2 billion liters of waste vegetable oil that could be converted to transportation fuels. Combined with other U.S.-produced vegetable oils, along with fats and greases, these feed stocks, however, would only have the potential of replacing approximately five percent of the total petroleum diesel usage (Figure 1-5). One of the main drivers in the development of micro algal diesel fuels is the higher photosynthetic efficiency of microalgae when compared to conventional crops, and hence the potentially higher productivities per unit area. Depending on productivity, algae productivity could range from 65% of oil palm (6000 L/ha/yr.) to surpassing that crop by nearly an order of magnitude. Using the higher oil content scenario (but demonstrated productivity), if the same amount of land currently devoted to the US soybean crop (75 million acres), microalgae could produce more than enough feedstock for biodiesel or green diesel to meet the current U.S. diesel fuel usage . Potential of Various Oil-Based Feed stocks Sources: Energy Information Administration

(EIA), National Biodiesel Board, and Corn Refiners Association. Demonstrated at Roswell Higher oil content Higher productivity g/m2 /day 10 10 50 lipid content 15 40 40 days of operation per year 330 330 330 percent land devoted to ponds 70 70 70 L/ha/yr. 3,800 10,200 50,800 Table 1-1 Microalgae Potential Yields 12 Crop Oil Yield (Liters/ha/yr.) Soybean 450 Camelina 560 Sunflower 955 Jatropha 1,890 Oil palm 5,940 Algae 3,800-50,800a an estimated yields, this report Table 1-2 Comparison of Oil Yields from Biomass

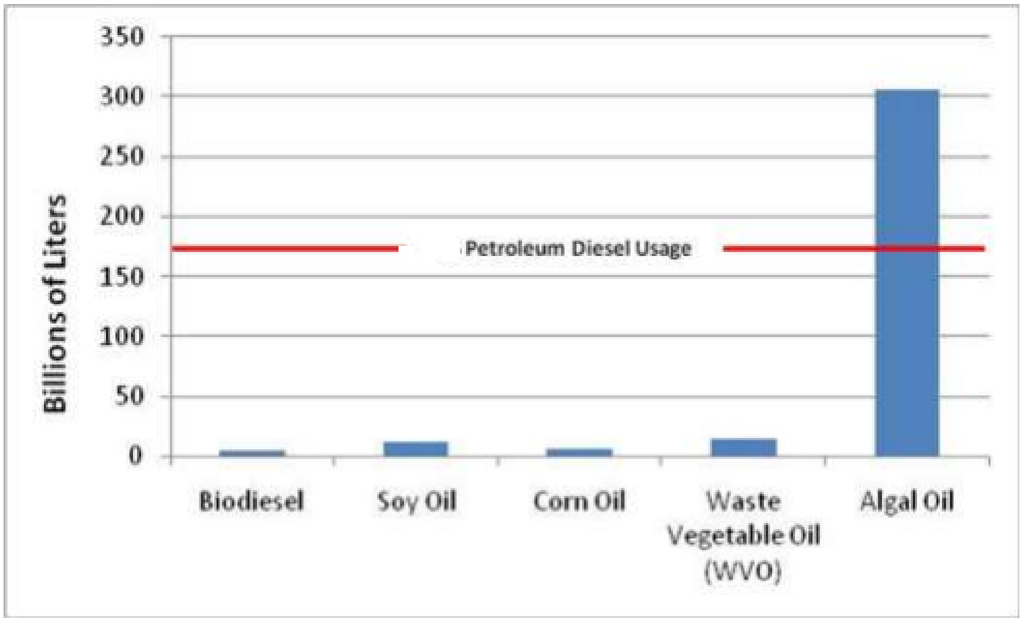
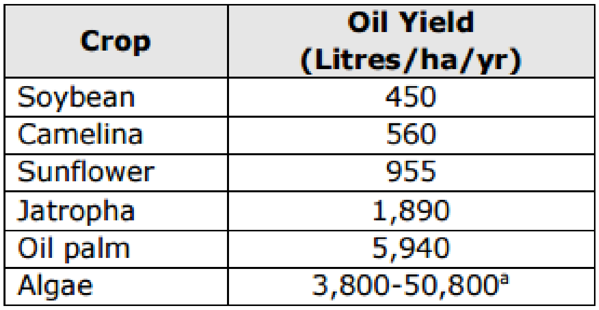


Figure 3: Biodiesel Sources

Table 2: Comparison of Algae Oil yield to Other Sources



Crop	Oil Yield (Litres/ha/yr)
Soybean	450
Camelina	560
Sunflower	955
Jatropha	1,890
Oil palm	5,940
Algae	3,800-50,800[a]

Literature Review

Algae Species

Microalgae are a promising alternative source of lipid for biodiesel production. One of the most important decisions is the choice of species to use. High lipid productivity is a key desirable characteristic of a species for biodiesel production. This paper reviews information available in the literature on micro algal growth rates, lipid content and lipid productivities for 55 species of microalgae, including 17 Chlorophyta, 11 Bacillariophyta and five Cyanobacteria as well as other taxa. The data available in the literature are far from complete and rigorous comparison across experiments carried out under different conditions is not possible. However, the collated information provides a framework for decision-making and a starting point for further investigation of species selection.

More than 10,000 living diatom species are known, there are over 8,000 species of Algae, there are 1500 species of Blue Green algae or Cyanobacteria, there are 6000 species of red algae or Rhodophyta. Algae can be found in a variety of environments, soil, fresh or saltwater. They come in various sizes and shapes. They can be single celled, have flagella and be motile, they can be branched. They can be red, green, brown or blue-green. They are usually common in a bio towers or a trickling filter. They can be found in many clarifiers, and in many lagoon or pond systems. They are often found in ornamental ponds or fountains. (*www.environmentalleverage.com*)

Some of the species used worldwide for biodiesel production are:

Amphora
Anabaena cylindrica
Ankistrodesmus falcatus
Chaetoceros muelleri
Chlorella pyrenoidosa
Chlorella vulgaris
Cyclotella cryptica
Dunaliella salina

Isochrysis galbana

Monodopsis subterranea

Monoraphidium minutum

Nannochloropsis

Nannochloropsis salina

Phaeodactylum tricornutum

Porphyridium purpureum

Scenedesmus obliquus

Spirulina maxima

Spirulina platensis

Tetraselmis suecica

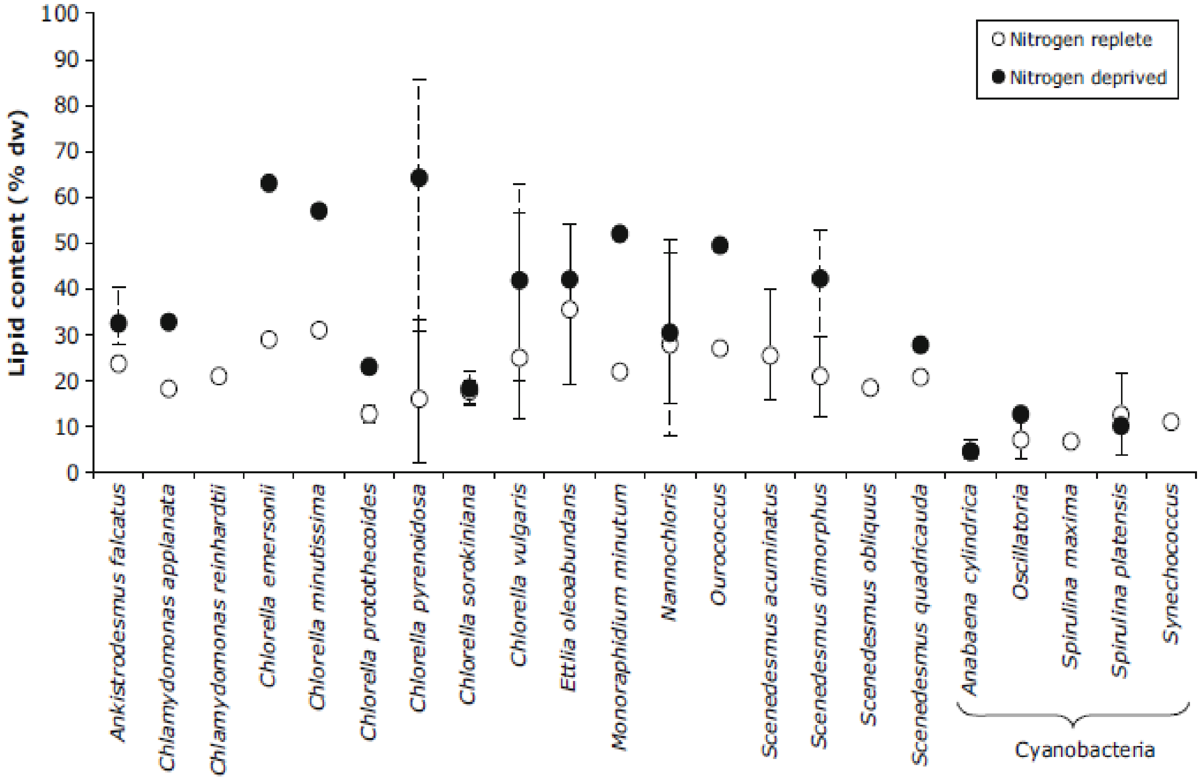


Figure 4: A comparision of lipid content in different algal species

(*Susan T. L. Harrison 2009*)

On basis of the above comparison we have chosen “Chlorella pyrenoidosa” as our algal specie. Chlorella pyrenoidosa can have up to 80% lipid content which gives highest yield of algal oil.

Chlorella pyrenoidosa is green algae that grow in fresh water, and Chlorella pyrenoidosa is an Algae supplement that might contain a combination of green algae, blue-green algae and spiraling. It can be grown in closed environment by providing necessary nutrients (Potassium, Iron, Nitrogen and Phosphorous) to have high lipid content at a high growth rate.

Sustainability

Sustainability is the subject of much discussion at international scientific and governmental forums on biofuels. Emerging from this discussion is a consensus that sustainability is of foremost importance as an overarching principle for the development of biomass-to-energy agro-industrial enterprises. While sustainability criteria that are agreeable to all nations are still being expounded, the generally accepted principles of sustainability include that:

- The greenhouse gas balance of the production chain is positive
- The biomass production is not at the expense of carbon sinks in existing vegetation and soil
- The biomass production does not endanger the food supply and existing local business activity (i.e. local supply of energy, medicines and building materials)
- The biomass production has no impact on biodiversity (protected or vulnerable biodiversity is not affected or if possible strengthened)
- Soil and soil quality are retained or improved
- Ground water and surface water are not depleted and water quality is maintained or improved;
- Air quality is maintained or improved; and

- The production and processing of biomass contributes to local prosperity and to the social well-being of employees and the local population.

It is self-evident that where there is a natural abundance of freshwater, it is likely on arable land (that may be under agriculture and may have multiple competing uses for the water resource), or on land in its natural state with considerable biodiversity value. With few exceptions where the abundance of freshwater is the consequence of human intervention, the water has multiple competing uses. Consequently, from the perspective of sustainability it seems obvious that algal production systems should target water resources other than freshwater. In fact, the proponents of algal biofuel claim that the production system is superior to biofuels based on terrestrial biomass because it can utilize non-arable land and waste water resources.

Types of cultivation

Open Ponds

Open pond systems are the most common system of algae cultivation, already used commercially in the United States to produce nutritional products and treat wastewater. In recent years many companies have begun focusing on this growth model for biofuels and chemicals. In fact, open ponds were the focus of the Department of Energy's Aquatic Species Program.

Open pond systems use shallow (typically one-foot deep) ponds, from about one acre to several acres in size, in which the algae are exposed to natural solar radiation (sunlight) which they convert into biomass. Typically the ponds are called raceway ponds because their shape resembles a race track. They often use paddle wheels or other water moving devices to keep the algae circulating.

The harvesting method is often a two stage process based on the particular properties of the algae and requirements of the process. A fraction of the pond water is generally harvested every day, and the algal biomass within the water is concentrated. The biomass is then processed further, for example to extract the oil for to conversion into biodiesel, jet fuel, or some other oil-based product. The residues, or even the entire biomass, can also be dried and used for animal feeds.

Closed Systems

A photo bioreactor (PBR) differs from an open pond in that the algae enclosed in a transparent vessel, which can be as simple as a greenhouse, but, more generally, is a tubular, bag-type or panel design, in many shapes and sizes, oriented vertically or horizontally. Some systems even use additional artificial light to help boost production, and a few rely exclusively on artificial lights.

One of main advantages of PBRs is that they can better match the ideal conditions and growth requirements of particular types of algae not easily grown in open ponds. They can also prevent, or at least reduce, invasion by weed algae, zooplankton grazers and other organisms that could affect the cultures.

Hybrid Systems

While most companies have chosen to focus exclusively on open pond systems, others on PBRs, and a few on fermentation, some companies are seeking to leverage the advantages of each system by using ‘hybrid’ processes that combine two or more of the above methods. These can be small PBRs that inoculate large ponds, larger PBRs used in combination with ponds, or ponds and fermenters used sequentially. The objective of hybrid systems is to maximize the individual advantages of each process. Phial, an Ohio-based company with operations in Hawaii, uses this type of system

Integrated Systems

This category leverages the ability of algae to treat wastewater by absorbing nutrients and CO2, while breaking down and removing unwanted, even toxic materials. The algal biomass produced during such wastewater treatment processes can then be used to generate methane, produce fertilizers, and can also yield oil and other liquid fuels.

The common thread in any integrated system is that the waste—whether it is dirty water or even polluted air—acts as an input during the growth phase. In other words, the contaminants that algae are able to clean from the air and water serve as ‘food’ for the algae in their growth cycle. By removing nutrients, something that algae can do very efficiently, algal blooms in lakes and coastal waters can be prevented or minimized, solving a major environmental issue.

Excretion Processes

Many companies are working to modify algae to produce biofuels through “excretion.” That is, instead of storing oils in the biomass, the algae would excrete useful chemical into the culture, or medium in which they are grown. This approach has the advantage of requiring only an initial, and small, amount of algal biomass that would continually produce oils, avoiding the need for harvesting and processing which can add significantly to the overall cost. Ethanol, butanol, fatty acids, hydrocarbons, gaseous fuels, and many more useful products all can be excreted by modifying the algae. San Diego-based Synthetic Genomics is working with Exxon on this type of approach.

Tubular Reactors

Most configurations of tubular reactors (TR) are one of the following three types:

(i) Simple airlift and bubble column, which is composed of vertical tubing (in the form of a vertical tubular reactor) that is transparent so as to allow for light penetration and where CO2 is supplied via bubbling

(ii) Horizontal tubular reactor, which is composed of horizontal transparent tubing, usually bearing gas transfer systems attached to the connections

(iii) Helical tubular reactor, which is composed of a flexible plastic tube coiled in a circular framework.

Vertical Tubular Reactor

The airlift and bubble column reactors are examples of vertical tubular reactors (VTR), regularly composed of polyethylene or glass tubes (Figure 1), which are sufficiently transparent to allow good light penetration but are manufactured with sufficiently common materials so as to be not expensive. Air is bubbled at the bottom-a strategy that provides good overall mixing, sufficient supply of CO2, and efficient removal of O2. More recently, pilot-scale photo-bioreactor that uses sunlight and flue gas, and consists of a vertical tubular part (kept in the dark) and a horizontal tubular part (subject to sunlight). Although cultivation of microalgae in the above systems is simple and hence widely employed (including in hatcheries), the corresponding technology is somehow primitive, with obvious constraints derived from the high fragility and the low versatility of the material in stake. Furthermore, scale-up of these systems was initially thought to be easy, but accumulated experience has indicated that increases in culture volume decrease bag productivity.

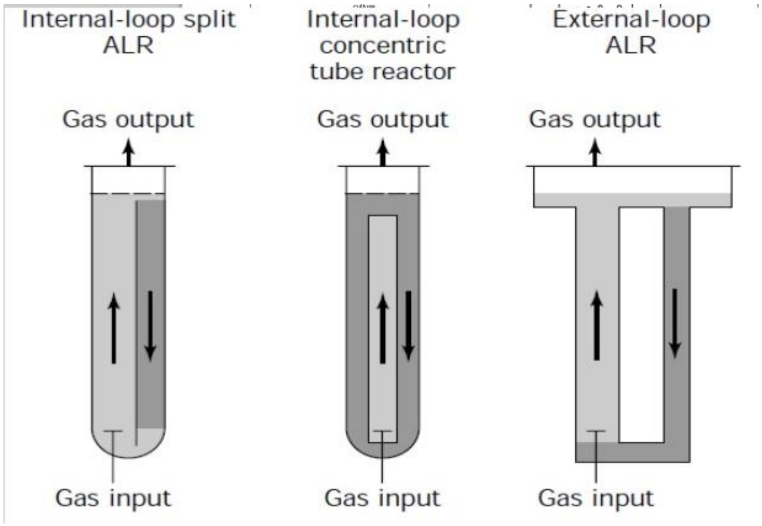


Figure 5: Schematic of types of Airlift Photo bioreactors

Horizontal Tubular Reactor

Horizontal tubular reactors (HTR) have been on the rise; gas transfer takes place in the tube connection or via a dedicated gas-exchange unit, and the angle toward sunlight is particularly adequate for efficient light harvesting. Such systems can handle large working volumes, because they are not susceptible to contamination. On the other hand, they may generate considerable amounts of heat, which may attain temperature amplitudes of 20 °C within a single day if (costly) temperature control systems are not provided; thus, it will likely pose a problem for regular operation.

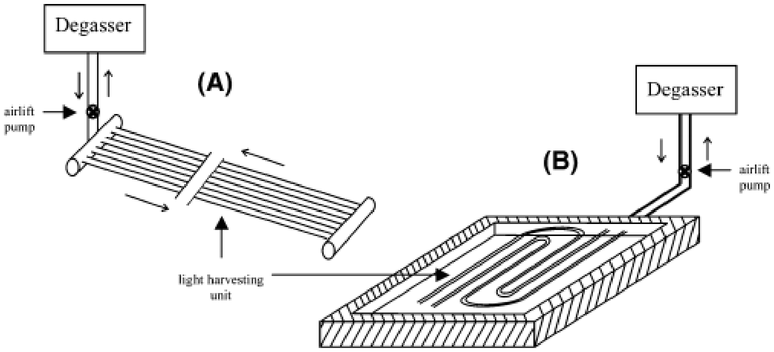


Figure 6: Schematic representation of horizontal tubular reactor with a degassing unit and a light harvesting unit, composed of parallel sets of tubes (A) or a loop tube (B).

Helical Tubular Reactors

Helical tubular reactors (HeTR) are a suitable alternative to straight TR. The most frequently used layout is the Bio coil, initially proposed by Robinson and currently traded by Biotechna (Melbourne, Australia). This reactor is composed of a set of polyethylene tubes (3.0 cm of inner diameter) coiled in an open circular framework, coupled with a gas exchange tower and a heat exchange system; a centrifugal pump drives the culture broth through the long tube to the gas exchange tower (see Figure 4A). A few authors experimented with and eventually improved such a design, so it ranks at present among the most effective ones as a result of its high area-to-volume ratio on the one hand, and the requirement of only a small land area for relatively large volumes on the other. (*F. Xavier Malcata*)

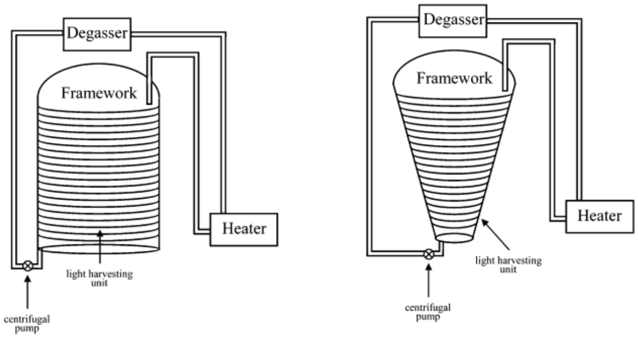


Figure 7: Schematic representation of helical tubular reactors: Bio coil (A) and conical framework (B).

Alpha-Shaped Reactors

Lee described a 300-Litre alpha-shaped tubular reactor, with sets of 2.5 cm × 25 m transparent tubes made of PVC; it used an airlift pump to promote an ascending/descending trajectory, with several CO2 injection points along its path. This configuration presents several advantages from the engineering point of view. For example, the fluid is pumped in a single direction (except in the airlift tubes), so a high flow rate is possible at the expense of relatively low air supply rates in the rising tubes. The ascending and descending trajectories are placed at a 45° angle toward sunlight, so light harvesting is rather efficient. (*F. Xavier Malcata*)

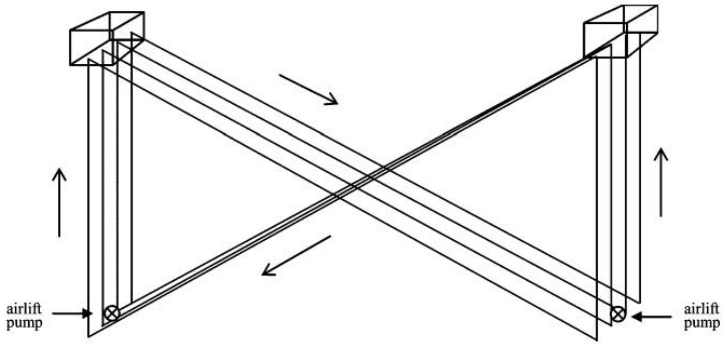


Figure 8: Schematic representation of alpha-shaped reactor

Flat plate reactors

(FPR) are conceptually designed to make efficient use of sunlight; hence, narrow panels are usually built so as to attain high area-to volume ratios (see Figure 6A). In the early 1980s, FPR were considered expensive and were even claimed to exhibit deficiencies in culture flow control. A 500-Liter FPR was developed, in which the culture was circulated from an open gas exchange unit through several parallel panels placed horizontally. The culture flew at a high linear speed (1.2 m s-1), but hydrodynamic parameters usually lay in a safe operating range for the sake of cell integrity. The greatest advantage of this system is its provision of an open gas transfer unit, which has proven

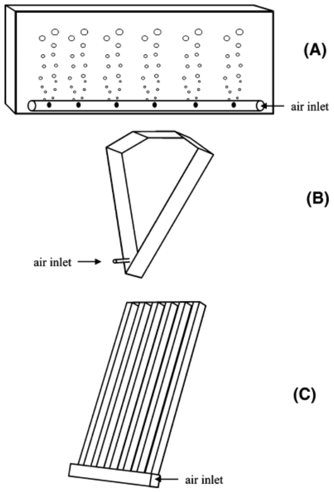


Figure 9: Schematic representation of flat panel reactors: flat panel bubbled in the bottom (A), V-shaped panel (B) and alveolar panel (C)

efficient in overcoming the problem of oxygen buildup; however, such an open zone restricts effectiveness of contamination control, as compared with completely closed reactors. (*Luı´s A. Meireles*)

Fermenter-Type Reactors

The least expanded systems for microalga cultivation are conventional fermenter-type reactors (FTR). This type of reactor presents indeed an intrinsic disadvantage: the area-to-volume ratio is quite low, so sunlight harvesting efficiency is poor. To overcome this nuclear drawback, sophisticated systems of internal illumination were developed, which are able to provide a more homogeneous distribution of light. When possible, microalgae may be heterotrophically cultivated in FTR, using appropriate organic carbon sources. A 250-L FTR was built by Pohl (61) with stainless steel and illuminated internally by fluorescent lamps placed inside narrow glass (or Plexiglas) tubes; CO2-enriched air was bubbled at the bottom, through a V-shaped (i.e., low shear stress) stirrer. Such a system was operated batch-, semi continuous and continuous-wise. The operation parameters could be fully controlled, so axenic cultures were maintained for long periods, as considered crucial for production of certain high-value metabolites. Although biomass productivity was quite low, there is still plenty of room for enhancement of growth parameters with this reactor configuration.

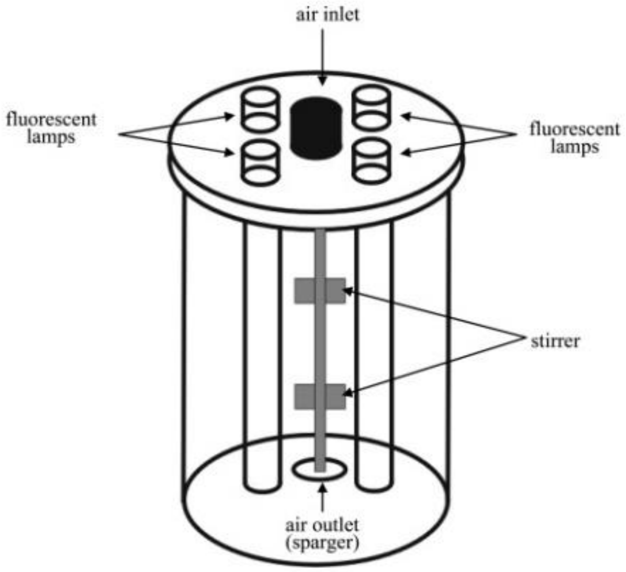


Figure 10: Schematic representation of fermenter-type bioreactor

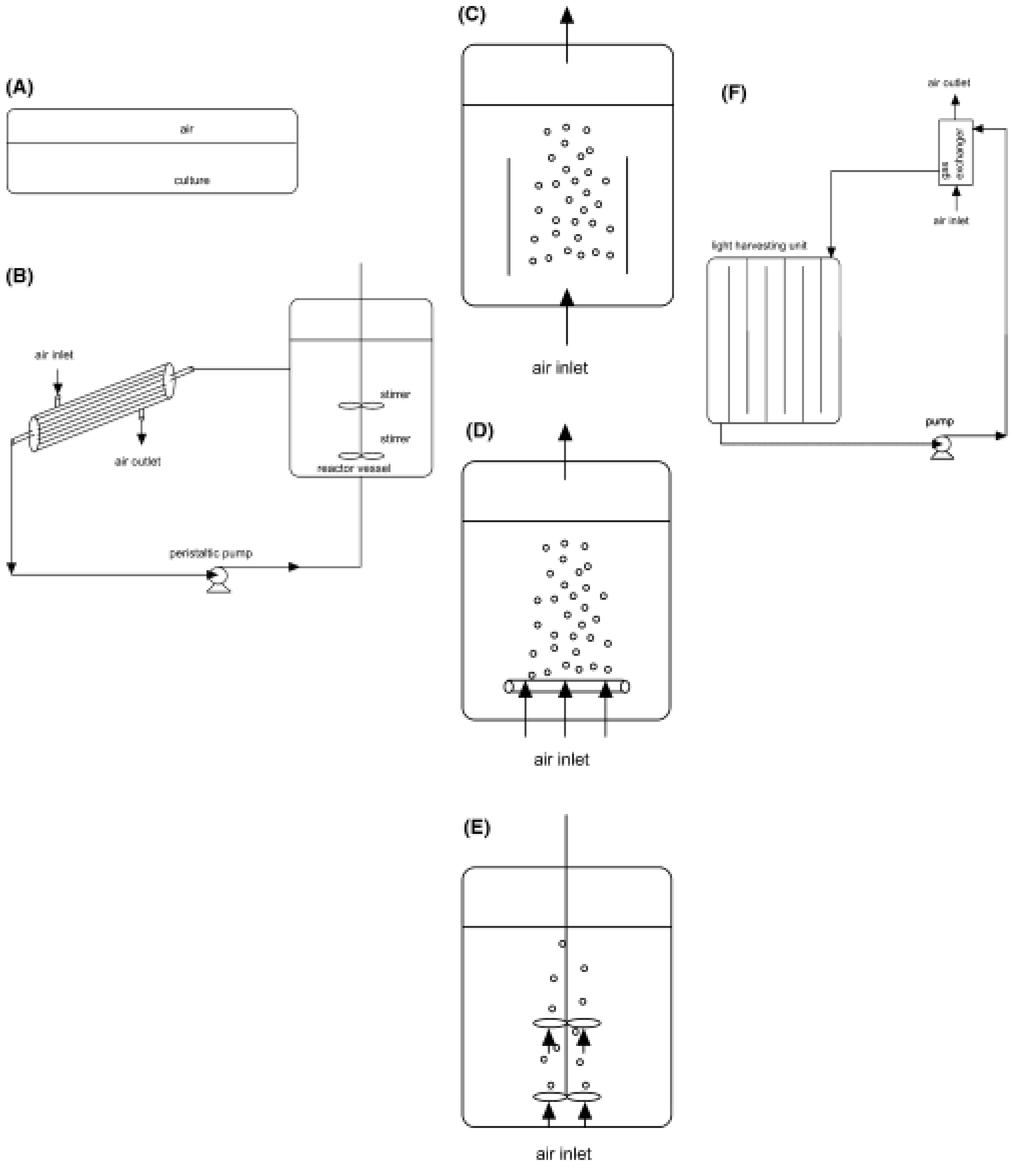


Figure 11 Schematic representation of major methods of gas exchange in micro algal reactors: surface driven aeration (A), microporous hollow fiber membranes (B), airlift loop (C), bubble column (D), stirrer blade bubbling (E), and gas exchanger system (F)

Separation Techniques

Centrifugation

Centrifugation is a process which involves the use of the centrifugal force for the sedimentation of heterogeneous mixtures with a centrifuge, used in industry and in laboratory settings. This process is used to separate two immiscible liquids. More-dense components of the mixture migrate away from the axis of the centrifuge, while less-dense components of the mixture migrate towards the axis. Chemists and biologists may increase the effective gravitational force on a test tube so as to more rapidly and completely cause the precipitate ("pellet") to gather on the bottom of the tube. The remaining solution is properly called the "supernate" or "supernatant liquid". The supernatant liquid is then quickly decanted from the tube without disturbing the precipitate. The rate of centrifugation is specified by the angular velocity measured in revolutions per minute (RPM), or acceleration expressed as g. The conversion factor between RPM and g depends on the radius of the sample in the centrifuge rotor. The particles' settling velocity in centrifugation is a function of their size and shape, centrifugal acceleration, the volume fraction of solids present, the density difference between the particle and the liquid, and the viscosity

Sedimentation

Sedimentation is the tendency for particles in suspension to settle out of the fluid in which they are entrained, and come to rest against a barrier. This is due to their motion through the fluid in response to the forces acting on them: these forces can be due to gravity, centrifugal acceleration electromagnetism. In geology sedimentation is often used as the polar opposite of erosion, i.e., the terminal end of sediment transport. In that sense it includes the termination of transport by saltation or true bed load transport. Settling is the falling of suspended particles through the liquid, whereas sedimentation is the termination of the settling process. Sedimentation of algae is a slow process and makes it unfavorable for a continuous process.

Chemical flocculation

Metal salts such as alum and ferric chloride are widely used for flocculation in industries such as water treatment and mining. Although metal salts are being applied for harvesting microalgae, their use results in high concentrations of metals in the harvested biomass. These metals remain in the biomass residue after extraction of lipids or carotenoid. The metals in the biomass residue may however interfere with the use of the protein fraction in this residue as animal feed. The valorization of the protein fraction as animal feed is said to be important for making micro algal biofuels economically viable. Despite this shortcoming, metal coagulants provide a good model system to study the interaction between flocculants and micro algal cells because their properties are well understood.

Other commonly used chemical flocculants in other industries are synthetic polyacrylamide polymers. These may however contain traces of toxic acrylamide and thus also contaminate the micro algal biomass. Flocculants based on natural biopolymers are therefore a safer alternative. To be able to interact with the negative surface charge of micro algal cells, these biopolymers should be positively charged, which is rare in nature. A well-known positively charged biopolymer is chitosan, which is derived from chitin, a waste product from shellfish production. Chitosan is a very efficient flocculent but it works only at low pH, but pH in micro algal cultures is relatively high. An alternative to chitosan is cationic starch, which is prepared from starch by addition of quaternary ammonium groups. The charge of those quaternary ammonium groups is independent of pH and therefore cationic starch works over a broader pH range than chitosan. Other examples of biopolymers than can be used to flocculate microalgae are poly-γ glutamic acid (an extracellular polymer produced by Bacillus subtilis) or polymers present in flour from Moringa oleifera seeds. A general problem of polymer flocculants is that they undergo coiling at high ionic strengths and become ineffective. Therefore, they are less suitable for harvesting microalgae cultivated in seawater. (*Imogen Foubert 2013*)

Bio flocculation

In natural blooms of microalgae occurring in lakes or rivers, flocculation sometimes occurs spontaneously. This spontaneous flocculation is assumed to be caused by extracellular polymer substances in the medium and is called bio flocculation. Bio flocculation is often successfully used for harvesting microalgae in facilities where microalgae are used in wastewater treatment. The underlying mechanism, however, is poorly understood and deserves further research because it may lead to a chemical-free method for flocculating microalgae. Some micro algal species flocculate more readily than others and such naturally bio flocculating microalgae can be mixed with other species to induce flocculation. There are indications that bio flocculation may be initiated by info chemicals. Recently, an info chemical isolated from a senescent and flocculating culture of a Skeletonema species was found to be capable of inducing flocculation in a culture of another species of microalgae.

Bacteria or fungi can also induce bio flocculation of microalgae. Some fungi, for instance, have positively charged hyphae that can interact with the negatively charged micro algal cell surface and cause flocculation. Specific consortia of bacteria can also induce flocculation of microalgae. These flocculating fungi or bacteria can be cultivated separately or in combination with the microalgae. Cultivating bacteria or fungi in combination with microalgae requires a carbon source in the medium. In wastewater, a carbon source is usually present and this allows cultivation of microalgae and bacteria. This results in a culture of mixed algal–bacterial flocs that can be easily harvested. The use of bacteria or fungi as a flocculating agent avoids chemical contamination of the biomass but results in microbiological contamination, which may also interfere with food or feed applications of the micro algal biomass.

Auto flocculation

Flocculation often occurs spontaneously in micro algal cultures when pH increases above 9. This type of flocculation is often referred to as auto flocculation because it occurs spontaneously in micro algal cultures as a result of a pH increase due to photosynthetic CO2 depletion. Auto flocculation is associated with the formation of calcium or magnesium precipitates. Depending on the conditions, these precipitates carry positive surface charges and can induce flocculation through charge neutralization and/or sweeping flocculation. Calcium phosphate precipitates are positively charged when calcium ions are in excess of phosphate ions and interact with the negative surface charge of micro algal cells. High phosphate concentrations are required for this type of flocculation to occur. As a result of the declining phosphate reserves and increasing prices of phosphate, flocculation by calcium phosphate precipitation is unsustainable, except perhaps in applications where microalgae are used for wastewater treatment and excess phosphate needs to be removed. Magnesium hydroxide or brucite also precipitates at high pH. These precipitates are positively charged up to pH 12 and can therefore also interact with the micro algal cell surface to cause flocculation. Most waters contain sufficiently high background concentrations of magnesium for this process to occur. Calcium carbonate or calcite also precipitates at high pH but whether it can induce flocculation of microalgae remains to be demonstrated. Flocculation at high pH is caused by formation of inorganic precipitates and not by pH as such, therefore, the harvested biomass contains high concentrations of minerals. Although these have a low toxicity, it is nevertheless preferable to remove them from the biomass.

Sand Filtration

The sand filter can be used for separation of algae from water. Sand filters are widely used in filtration to eliminate different kinds of compounds. A lot of work has been performed to study the various parameters that can play an important role in the process of granular filtration. The complexity of the phenomena that occur during the filtration process (mechanical, physicochemical and biological) and the diversity of parameters involved require that each study be conducted under specific conditions. It was demonstrated experimentally in a study 1 with sand of average size ranging between 0.064 and 0.335 mm that all media with median sand sizes of 0.200 mm or below gave consistently high micro-algae removal rates (the average removal was 97.3%). From the simulation results based on six different pore-size distribution functions 2, it was confirmed that pore-size distribution plays an important role in the removal efficiency as well as the head loss inside the sand bed. (R.F *Malone 2003*)

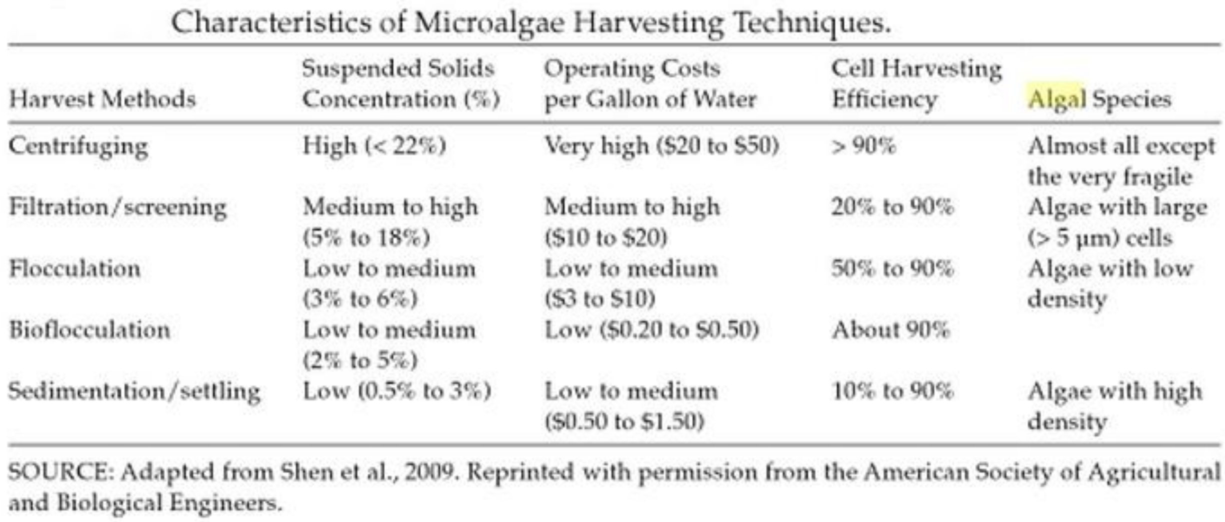
Coagulation

Chemical coagulation is commonly used in the treatment of water. The effective reduction of clay, silt, organic matter, algae, and bacteria in surface waters by coagulation and settling is demonstrated daily in water works operation. Even though the coagulation of clays and other inorganic sols has been reported in detail, similar attention has not been given to the coagulation of algae. Coagulation agents are used to separate algae from the dilute water solution before sending it for oil extraction.

Membrane

Membrane-based separations are very attractive as these offer several advantages over traditional separation technologies including high reliability and direct scalability along with superior chemical, mechanical and thermal properties. It is believed that biofuels from algal or cellulosic feed stocks can be more cost-effective through the generation of value added co-products. Membrane-based dewatering offers the potential to concentrate biomass to recover valuable co-products such as protein and for more efficient downstream processing to produce algal oils, biodiesel, or jet fuels. For cost-effective processing, hollow fiber membranes can be utilized to recover 90–95% media for recycle. Tubular membranes can provide additional media and water recovery to achieve target final concentrations. Based on the operating conditions used in this study and taking into scale-up considerations, an integrated hollow fiber-tubular membrane system can process micro algal biomass with at least 80% lower energy requirement compared to traditional processes. Back pulsing was found to be an effective flux maintenance strategy to minimize flux decline at high biomass concentration. (*Lawrence Powell 2012*)

Table 3: Characteristics of harvesting techniques



Characteristics of Microalgae Harvesting Techniques.

Harvest Methods	Suspended Solids Concentration (%)	Operating Costs per Gallon of Water	Cell Harvesting Efficiency	Algal Species
Centrifuging	High (< 22%)	Very high ($20 to $50)	> 90%	Almost all except the very fragile
Filtration/screening	Medium to high (5% to 18%)	Medium to high ($10 to $20)	20% to 90%	Algae with large (> 5 μm) cells
Flocculation	Low to medium (3% to 6%)	Low to medium ($3 to $10)	50% to 90%	Algae with low density
Bioflocculation	Low to medium (2% to 5%)	Low ($0.20 to $0.50)	About 90%	
Sedimentation/settling	Low (0.5% to 3%)	Low to medium ($0.50 to $1.50)	10% to 90%	Algae with high density

SOURCE: Adapted from Shen et al., 2009. Reprinted with permission from the American Society of Agricultural and Biological Engineers.

Types of Extraction

Hexane Solvent Extraction

Many plants and plant-material, such as algae, oil-seeds, cereal brans, beans, nuts, and microbial organisms, contain oils that can be useful for many commercial products. These oils are used in cooking, processing foods, cosmetics, lubricants, and a host of other useful products. Because of this high commercial demand, much Work had been done in an attempt to improve oil extraction processes to make them more efficient and more suitable for mass extraction. Numerous processes for the extraction of oil are known in the art. The most commonly used process is solvent extraction from a dried plant material. To use the conventional process, the plant material must already be dry. The plant material may be pretreated, for example, by flaking to facilitate penetration of the plant structure by a solvent, such as hexane, without creating fine particles. The dried, lipid containing plant material is then contacted with the solvent that will dissolve the oil or other valuable lipids and extract them out of the material. Contact time is provided with the solvent typically by means of counter-current washing. The resulting mixture of solvent and lipid material (the miscella) is separated from the extracted plant material and fractionated to remove the solvent, leaving the lipid. This process is problematic when applied to oil containing microbial mass. To remove oils from the microbial biomass, the biomass must first be dried, e.g., by spray drying, then slurried in the solvent. Biomass is produced in relatively dilute aqueous slurry (fermentation culture), which means drying is an expensive process. Additionally, the temperature profile during drying must be such that oil quality is not compromised. Conventional extract equipment, which relies on coarse screens to retain the oil-bearing material, is not designed to handle the particles produced by such means. Second, the cells may need to be disrupted to permit adequate contact with the solvent. This cell disruption step generates a significant amount of fines which tend to be carried along With the product in the solvent. Consequently, before further processing, these fines must be removed by filtration, centrifugation, or a combination thereof. The fines clog equipment used in downstream processing steps and make extraction more difficult.

Third, the extracted biomass carries 10—50% hexane by Weight With it. This hexane Will contain some product, Which is now lost. Additionally, the hexane must be substantially removed before the dilapidated biomass can be disposed. Extraction of oil from high moisture materials, including
animal products, such as eggs, and microbial biomass, have been described using polar solvents that are partly or completely miscible With Water. In a separate and distinct technology, addition of polymers to Water to create two immiscible phases, between Which Water soluble substances may be partitioned. However, these processes are not fully satisfactory for efficient extraction of non-polar lipids, such as triglyceride oils, on a commercial scale.

In accordance With a principle aspect of the present invention, it is a technical advantage of this invention to provide a novel method for separating edible oil from biological material. Another technical advantage of this invention is that it provides a novel method for extraction of lipids, specifically edible oil, from microbial biomass. The invention uses an appropriate solvent to extract oil from relatively fine particles in aqueous slurry without the need to dry the slurry or reform the material to create larger-sized particles. Another technical advantage of this invention is that it provides a novel method for separating edible oil from biological material that overcomes the problems of conventional methods. When disrupting the biomass in an aqueous phase and extracting without further drying, the fines stay in the aqueous phase and do not contaminate the solvent. Therefore, additional treatment of the solvent to remove the fines may be avoided. Moreover, hexane can be more easily removed from the aqueous liquid. Although hexane is soluble in Water up to 3%, this hexane removed by heating the aqueous liquid. (*United States Patent*)

Soxlet Extraction

Extraction refers to an operation in which one or more components of a liquid or a solid phase are transferred to another liquid phase. Extraction utilizes the differences in the solubility of components. In liquid-liquid extraction or solvent extraction, a solute in a liquid solution is removed by contacting with another liquid solvent, which is relatively immiscible with the solution. The liquid, which is added to solution to bring about the extraction, is known as solvent. This solvent layer is called extract and the other layer, the raffinate. Leaching involves treating a finely divided solid with a liquid that dissolves and removes a solute contained in the solid. Ex. oilseed extraction .The solvent extraction method recovers almost all the oils and leaves behind only 0.5% to 0.7% residual oil in the raw material. The solvent extraction method can be applied to any low oil content materials. It can also be used for pre-pressed oil cakes obtained from high oil content materials. Because of the high percentage of recovered oil, solvent extraction has become most popular method of extraction of oils and fats. Mechanical pressing methods include the expeller method, hydraulic presses etc.

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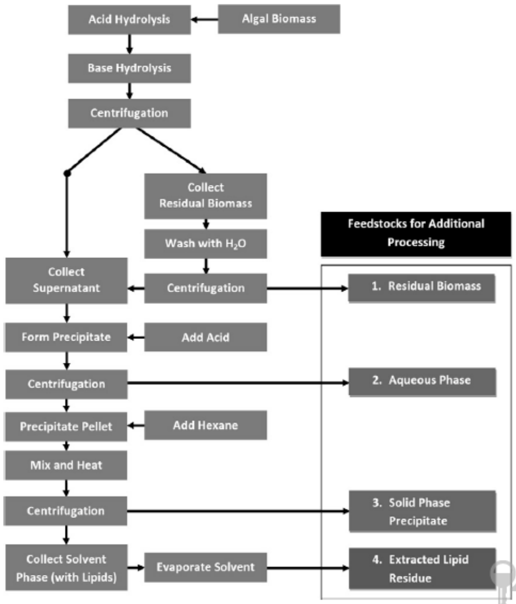
Super Critical Carbon Dioxide Solvent Extraction

Supercritical fluid extraction (SFE) holds greater potential for the extraction of algae lipids in higher concentration and purities than traditional organic separation methods. SFT Inc. has developed a method to extract algae oil utilizing supercritical carbon dioxide. it utilizes pressurized carbon dioxide, allowing extraction to take place in room temperature environments, so purer, less thermally decomposed extract is generated. The carbon dioxide is tunable, meaning certain conditions such as flow rate, temperature, and pressure can yield products that are lipid specific. Supercritical carbon dioxide also separates the oils in a clean, energy efficient fashion. From a large scale perspective, the carbon dioxide can be recycled from supercritical extraction into a necessary component for algal growth. To determine which parameters would be most advantageous for future production, several varieties were extracted via supercritical carbon dioxide. GC (Gas Chromatography)/MS (Mass Spectroscopy) analysis is used to determine which alga produces the highest and purest yield of oil. Three types of algae samples were utilized in the experiment: untreated fresh Schizochytrium, spray dried Nannochloropsis, and freeze dried Nannochloropsis. The fresh algae sample was newly cultivated algae frozen and stored in the dark. Freeze dried algae was frozen in the first step of storage. Following the freeze, the algae are depressurized to 0.01 psi. The ice is removed via a cold condenser and vacuum chamber. The final algae product is left dry with its original components except for water. Spray dried algae is prepared by flash drying the cultivated sample. Air is thrust into a combustion chamber of a drying unit with fuel until it explodes. This controlled explosion creates 3 psi pressurized air. The dryer controls the explosion with air to control the heat that comes into the presence of the algae. Combustion occurs in a controlled series to yield algae with original nutrients. (*Yusuf Chishti 2007*)

Wet Lipid Extraction

Microalgae are a source of renewable oil for liquid fuels. However, costs for dewatering/drying, extraction, and processing have limited commercial scale production of biodiesel from algal biomass. A wet lipid extraction procedure was developed that was capable of extracting 79% of transesterifiable lipids from wet algal biomass (84% moisture) via acid and base hydrolysis (90 C and ambient pressures),and 76% of those extracted lipids were isolated, by further processing, and converted to FAMEs. Furthermore, the procedure was capable of removing chlorophyll contamination of the algal lipid extract through precipitation. In addition, the procedure generated side streams that serve as feed stocks for microbial conversion to additional bio products. The capability of the procedure to extract lipids from wet algal biomass, to reduce/remove chlorophyll contamination, to potentially reduce organic solvent demand, and to generate feed stocks for high-value bio products presents opportunities to reduce costs of scaling up algal lipid extraction for biodiesel production. (*Satish A 2012*)



Wet lipid extraction procedure block diagram (WLEP showing side streams).

Figure 12: Wet Lipid Extraction

Mechanical Extraction

Ultrasonic Extraction

Intense sonication of liquids generates sound waves that propagate into the liquid media resulting in alternating high-pressure and low-pressure cycles. During the low-pressure cycle, high-intensity small vacuum bubbles are created in the liquid. When the bubbles attain a certain size, they collapse violently during a high-pressure cycle. This is called cavitation. During the implosion very high pressures and high speed liquid jets are produced locally. The resulting shear forces break the cell structure mechanically and improve material transfer. This effect supports the extraction of lipids from algae.

Ultrasonic Solvent Extraction

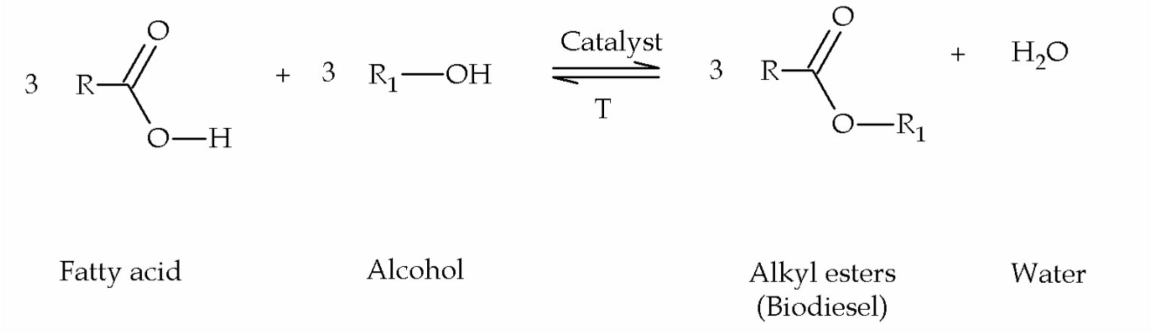
The high pressure cycles of the ultrasonic waves support the diffusion of solvents, such as hexane into the cell structure. As ultrasound breaks the cell wall mechanically by the cavitation shear forces, it facilitates the transfer of lipids from the cell into the solvent. After the oil dissolved in the cyclohexane the pulp/tissue is filtered out. The solution is distilled to separate the oil from the hexane.

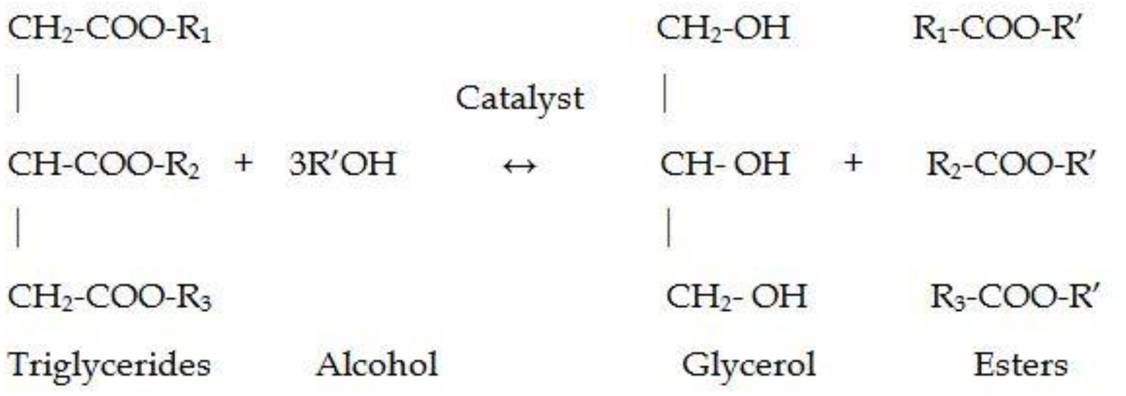
Expellers

This is a mechanical method that uses expeller to press the algae. Similar methods are screw expeller method, mechanical pressing method (by piston) and osmotic shock method. In the osmotic shock method the osmotic pressure is suddenly reduced. We have carried out our work on screw expeller. The raw materials are squeezed under high pressure in a single step. Expeller presses can recover 75% of the oil from algae. The alga was obtained from the open pond system. It was dried by exposure to atmosphere. In an expeller press, as the raw material is pressed, friction causes it to heat up; in some cases, the temperatures may exceed 120o F. The expeller we used was a screw type machine that presses oil seeds through a caged barrel like cavity. Algae entered the expeller press on one side of the press and products exit was on other side of the press. The machine used friction and continuous pressure from the screw drive to compress the filamentous algae. The algae were green in long strands like fiber. Initially, the algae did not move easily into the screw. Its surface had to be wetted with water for easy movement through the caged barrel. The oil seeps through small openings that do not allow the other components to seep through. Afterwards, the pressed algae almost form cake, was removed from the machine. Pressures involved in expeller pressing create heat in the range of 140-210o F. Expeller processing cannot remove the last trace of oil from algae. A significant amount of oil was left in the cake formed. The cake formed was in large quantity. It was not subjected to solvent extraction, since the quantity of solvent required would have been much greater. (*Niraj S. Toparea 2011*)

Reactors for Biodiesel

The reactions through which biodiesel (esters) are produced are esterification and transesterification.





$$\begin{array}{llcll} CH_2\text{-}COO\text{-}R_1 & & & CH_2\text{-}OH & R_1\text{-}COO\text{-}R' \\ | & & \text{Catalyst} & | & \\ CH\text{-}COO\text{-}R_2 & +\ 3R'OH & \leftrightarrow & CH\text{-}OH\ + & R_2\text{-}COO\text{-}R' \\ | & & & | & \\ CH_2\text{-}COO\text{-}R_3 & & & CH_2\text{-}OH & R_3\text{-}COO\text{-}R' \\ \text{Triglycerides} & \text{Alcohol} & & \text{Glycerol} & \text{Esters} \end{array}$$

Batch Reactors

The batch reactor can be simply a tank that is equipped with some type of agitation. The tank is filled with the reactants for the process (in this case oil, alcohol, and catalyst), and then the agitator is operated for some period of time. After the required time has elapsed, the contents of the reactor are drained out and further processed.

The main characteristic of a batch reactor is that it starts with unreacted material, causes it to react, and then at a later time ends up with reacted material. That is, a batch reactor contains different types of material depending on which time one happens to look at it. Batch reactors are generally used in small biodiesel production plants. A disadvantage of batch processes is that, to increase production, it might be necessary to increase the physical size of the plant by a proportional amount (by buying another reactor, for example). In contrast, when using a continuous flow process, it is usually possible to increase the plant’s production capacity by increasing the feed rate or shortening the reaction time

Continuous-Flow Reactors

The most common continuous-flow system in biodiesel production is the continuous stirred-tank reactor (CSTR). At first glance a CSTR appears to be identical to a batch reactor. In fact, often the actual reactor may be the same, but additional controls are needed to set the reactor up in a continuous-flow system. Some continuous-flow plants may be able to operate in either batch or continuous mode.

In a CSTR, the reactants are continuously added and the product (mixture of different chemicals, including unreacted reactants) continuously withdrawn. Adequate agitation is required to ensure uniform chemical composition and temperature. The continuous-flow process typically requires intricate process controls and online monitoring of product quality.

When a CSTR is operated continuously at a steady state, ideally the concentration of any chemical involved should be approximately constant anywhere in the reactor and at all times. In reality, this ideal state is rarely achieved; thus, adjustments need to be made to operating parameters to ensure complete reaction.

Sometimes more than one reactor is used. In this system, approximately 80 percent of the alcohol and catalyst are added to the oil in a first-stage CSTR. Then, the reacted stream goes through a glycerol removal step before entering a second CSTR. The remaining of the alcohol and catalyst are added to this reactor. This system provides a very complete reaction with the potential of using less alcohol than single-step systems.

Ultrasonic Biodiesel Reactors

Ultrasound is a useful tool to mix liquids that tend to separate. In biodiesel production, adequate mixing is required to create sufficient contact between the vegetable oil/animal fat and alcohol, especially at the beginning of the reaction. Ultrasonic waves cause intense mixing so that the reaction can proceed at a much faster rate.

Ultrasound transfers energy into fluid and creates violent vibrations, which form cavitation bubbles. As the bubbles burst, a sudden contraction of the fluid occurs, and the ingredients are mixed in the area of the bubbles. Such a high-energy action in liquid can considerably increase the reactivity of the reactant mixture and shorten the reaction time without involving elevated temperatures. In fact, this reaction can be achieved at or slightly above ambient temperature. Because there is no need to heat the mixture, energy may be saved. The ultrasound processing results in similar yields of biodiesel with a much shortened reaction time compared to the conventional stirred-tank procedure.

Ultrasound can be a good choice for small producers, who may only need one or two ultrasound probes per reactor vessel. However, using ultrasound in large-scale processing may be challenging because many ultrasound probes would be needed to reach every area of the reactant mixture.

Supercritical Reactors

Traditional biodiesel production requires a catalyst (usually sodium or potassium hydroxide) to complete the trans-esterification reaction. After the reaction, the catalyst has to be removed to ensure fuel quality. This can sometimes be problematic. To avoid the catalyst requirement, trans-esterification can be achieved in a catalyst-free manner by using a "supercritical" process. A critical point of a fluid is defined by its critical temperature and critical pressure, "the highest temperature and highest pressure at which a pure chemical species is observed to exist in vapor/liquid equilibrium at the supercritical state, the phase boundary between liquid and vapor starts to disappear, and the substance has qualities of both a liquid and a vapor.

When transesterification occurs during the supercritical state of methanol (typically high temperature and high pressure), the vegetable oil or animal fat dissolves in methanol to form a single phase. The reaction then occurs to reach completion in a few minutes without any catalysts.

The supercritical process tolerates water and free fatty acids in the system, and the soap formation that is common in the traditional process is eliminated. Since the supercritical state demands very high temperature and pressure, the process can be expensive. Nevertheless, large biodiesel producers may find this process to be cost effective because, since the reaction happens so quickly, producers can make a large quantity with a relatively small reactor and limited space.

Static Mixers as Biodiesel Reactors

Static mixers are simple devices consisting of spiral-shaped internal parts within an enclosure, such as a tube or pipe that promote turbulent flow. They have no moving parts, are easy to use and maintain, and are very effective at mixing liquids that are not readily miscible under normal conditions.

Biodiesel production from vegetable oils and alcohols is limited initially by the solubility of alcohol in vegetable oils. Static mixers can be used to mix the reactants before they enter the reactor vessel. A study was conducted at the University of Idaho to explore the possibility of using a static mixer as a continuous-flow reactor for biodiesel production.

The results showed that the static mixer reactor was effective for biodiesel production, and products meeting the ASTM D6584 specification were obtained. As with other reactor configurations, temperature and catalyst concentration influenced the product yield significantly. The most favorable conditions for complete trans-esterification were 60°C and 1.5% catalyst for 30 minutes. It is feasible, therefore, to use a static mixer alone as the reactor for biodiesel preparation from vegetable oils and alcohols. A similar process is sometimes used commercially, but the use of a large static mixer as the biodiesel processor has not been commercialized. (*Reactors for biodiesel*)

Fixed Bed Reactors

A novel continuous fixed bed reactor process has been developed for the production of biodiesel using a
metal oxide-based catalyst. Porous zirconia, titania and alumina micro-particulate heterogeneous
catalysts are shown to be capable of continuous rapid esterification and trans-esterification reactions
under high pressure and elevated temperature (300–450 8C). The continuous trans-esterification of triglycerides and simultaneous esterification of free fatty acids with residence times as low as 5.4 s is described. Biodiesel produced from soybean oil,

acidulated soap stock, tall oil, algae oil, and corn oil with different alcohols to make different alkyl esters using this new process pass all current ASTM testing specifications.

Furthermore, the economics of this novel process is much more cost competitive due to the use of inexpensive lipid feed stocks that often contain high levels of free fatty acids. The process has been shown to easily scale up a factor of 49 for more than 115 h of continuous operation without loss of conversion efficiency. The increased use of biodiesel world-wide could help reduce the emission of greenhouse gases that are linked to the progression of global warming. (*Thomas R. Hoye 2008*)

Catalyst Selection for Reactor

Zirconia Dioxide

Preparation of hydroxide modified zirconia, titania and alumina particles Hydroxide-modified zirconia, titania and alumina particles were synthesized using identical procedures. Unmodified zirconia, titania or alumina particles (50 g) were suspended in 300 mL of 2.0 M sodium hydroxide and sonicated under house vacuum (200 mbar) to allow the liquid to penetrate the pores while with gentle swirling by rotating the flask for 2 h at ambient temperature. The particles were then allowed to settle and the supernatant liquid was decanted. HPLC-grade water (300 mL) was added to the flask and the particles were suspended by swirling, allowed to settle, and again isolated by decantation. This washing process was repeated two more times. The particles were then re-suspended in 300 mL of HPLC-grade water and collected by vacuum filtration on a 0.45 mm nylon membrane (Waters, Milford, MA, USA) HPLC-solvent filter. The collected particles were washed with 2 aliquots (200 mL) of HPLC-grade water followed by 3 aliquots (200 mL) of HPLC-grade methanol. Finally, air was drawn through the particle cake for about 8 h until the particles were dry and free flowing. (*Thomas R. Hoye 2008*)

solid catalysts, both acid and basic (Santacesaria et al., 2007). In fact, heterogeneous catalysts are either easily separated or retained in the reactor and may result into higher quality of both product (esters) and byproduct (glycerol). In fact pure high grade glycerol can be obtained, in this case, without expensive refining operations (Bourney et al., 2005). For these reasons, many heterogeneous catalysts, based on both basic and acid solids, have recently been proposed in the literature (Santacesaria et al., 2006; Furuta *et al*., 2004; Furuta *et al*., 2006). In particular solid super acids appear to be well suited for promoting the transesterification of vegetable oils as well as the esterification of free fatty acids (Furuta *et al*., 2004; Lopez et al. (2007), Lopez et al.2008).Titanium doped zirconia employed as the solid catalyst by Furuta et al (2004) was tested for the transesterification of rapeseed oil with *methanol. (F. Scargiali 2008)*

Zirconia oxide properties:

- Use temperatures up to 2400°C
- High density
- Low thermal conductivity (20% that of alumina)
- Chemical inertness
- Resistance to molten metals

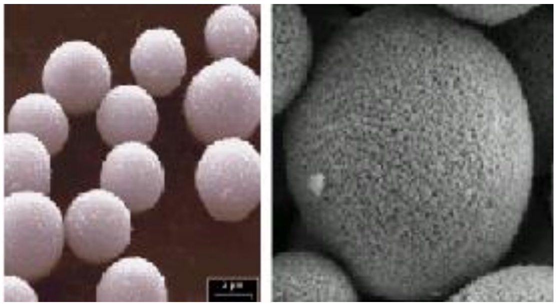


Figure 13: Catalyst

- Ionic electrical conduction
- Wear resistance
- High fracture toughness
- High hardness

Safety and Handling

Stability

Oxidative stability is a major industry issue for diesel and biodiesel fuels. Some biodiesels are more stable than others and some unstable biodiesel contain stability additives that perform very well. The tendency of a fuel to be unstable can be predicted by the Iodine number but the test method may not pick up the presence of stability additives. Iodine number actually measures the presence of C=C bonds that are prone to oxidation. The general rule of thumb is that instability increases by a factor of 1 for every C=C bond on the fatty acid chain.

Poor stability can lead to increasingly high acid numbers, increasing viscosity, and the formation of gums and sediments that can clog filters. Comparing the fuel‘s acid number and viscosity over time can provide some idea about whether or not the fuel is oxidizing, but you need to take a sample at the beginning when the fuel is fresh and then sample on a regular basis after that.

Material compatibility

Plastics: High density polyethylene (HDPE) is compatible but polyvinyl chloride (PVC) is slowly degraded. Polystyrenes are dissolved on contact with biodiesel.

Metals: Biodiesel has an effect on copper-based materials (e.g. brass), and it also affects zinc, tin, lead, and cast iron. Stainless steels (316 and 304) and aluminum are unaffected.

Rubber: Biodiesel also affects types of natural rubbers found in some older engine components. Studies have also found that fluorinated elastomers (FKM) cured with peroxide and base-metal oxides can be degraded when biodiesel loses its stability caused by oxidation. However testing with FKM- GBL-S and FKM- GF-S were found to be the toughest elastomer to handle biodiesel in all conditions

Flash point

A minimum flash point for diesel fuel is required for fire safety. B100‘s flash point is required to be at least 93ºC (200ºF) to ensure it is classified as nonhazardous under the National Fire Protection Association (NFPA) code. It is critical to ensure that the manufacturer has removed excess methanol used in the manufacturing process. Residual methanol in the fuel is a safety issue, because even very small amounts reduce the flash point; can affect fuel pumps, seals, and elastomers; and can result in poor combustion properties. The intent of the alcohol control requirement is to limit methanol to less than

0.2 wt. %. This can be accomplished by meeting a higher flash point requirement of 130ºC (266ºF); or by measuring methanol content by gas chromatography.

Viscosity

A minimum viscosity is required for some engines because of the potential for power loss caused by injection pump and injector leakage. This is not an issue for B100, and the minimum is set at the same level as for petroleum diesel. The maximum viscosity is limited by the design of engine fuel injection systems. Higher viscosity fuels can cause poor fuel combustion that leads to deposit formation as well as higher in-cylinder penetration of the fuel spray, which can result in elevated engine oil dilution with fuel. The maximum allowable viscosity in ASTM D975 for No. 2 diesel is 4.1 mm2/s at 104ºF (40ºC). ASTM D6751 allows for slightly higher viscosity than D975, primarily because that is where the normal viscosity of B100 lies. Biodiesel blends of 20 vol % or lower should have a viscosity within the range allowed by D975.

Sulfur

This is limited to reduce sulfate and sulfuric acid pollutant emissions and to protect exhaust catalyst systems when they are deployed on diesel engines in the future. Sulfur content of 15 ppm or lower is also required for proper functioning of diesel particle filters. Biodiesel generally contains less than 15 ppm sulfur.

Acid number

The acid number for biodiesel is primarily an indicator of free fatty acids (natural degradation products of fats and oils) and can be elevated if a fuel is not properly manufactured or has undergone oxidative degradation.

Contamination by water

Biodiesel may contain small but problematic quantities of water. Although it is not miscible with water, it is, like ethanol, hygroscopic (absorbs water from atmospheric moisture). One of the reasons biodiesel can absorb water is the persistence of mono and diglycerides left over from an incomplete reaction. These molecules can act as an emulsifier, allowing water to mix with the biodiesel. In addition, there may be water that is residual to processing or resulting from storage tank condensation. The presence of water is a problem because:

- Water reduces the heat of combustion of the bulk fuel. This means more smoke, harder starting, less power
.
- Water causes corrosion of vital fuel system components: fuel pumps, injector pumps, fuel lines, etc.
- Water & microbes cause the paper element filters in the system to fail (rot), which in turn results in premature failure of the fuel pump due to ingestion of large particles.
- Water freezes to form ice crystals near 0 °C (32 °F). These crystals provide sites for nucleation and accelerate the gelling of the residual fuel.
- Water accelerates the growth of microbe colonies, which can plug up a fuel system.

Biodiesel users who have heated fuel tanks therefore face a year-round microbe problem.

NOx Emission

NOx emission control use feed stocks that are not used as food like Waste Vegetable Oil, Jatropha or oil from Algae. Because biodiesel contains no sulfur you can use NOx Control Technologies which is impossible to use with petro diesel because of the sulfur content. Some companies have also successfully developed additives that can be used to reduce biodiesel's NOx Emissions.

Eye Contact

May cause irritation. Irrigate eye with water for at least 15 to 20 minutes. Seek medical attention if symptoms persist. Safety glasses, goggles, or face shield are recommended to protect eyes from mists or splashing.

Acute Oral Toxicity/Rates

Biodiesel is nontoxic and there are no hazards anticipated from Ingestion incidental to industrial exposure. If ingested, give one or two glasses of water to drink. If gastro-intestinal symptoms develop, consult medical personnel.

Skin Irritation – Humans

A 24-hr. human patch test indicated that undiluted biodiesel produced very mild irritation. The irritation was less than the result produced by 4 percent soap and water solution. Wash with soap and water. Avoid contact with hot fuel as it may cause burns. PVC coated gloves are recommended to prevent skin contact. May soften skin.

Storage and Gelling

Biodiesel storage:

Finished biodiesel fuel that has been washed and dried has a long shelf life when properly stored. Biodiesel should be stored in closed containers or cans that are completely filled to the top. Storing biodiesel in partially filled containers is not recommended because condensation can build up on the inside of the container and cause the biodiesel to deteriorate over a long period of time and can possibly cause the growth of bacteria and mold.

When storing biodiesel containers or cans, it's important to keep them away from direct sunlight and insulated from low temperatures. Biodiesel will gel at a higher temperature than petroleum-based diesel fuel, but it will return to its liquid state when it is heated or allowed to gradually warm up. Biodiesel containers should also be protected from precipitation and excess moisture because of the aforementioned issues with condensation.

The current industry recommendation is that biodiesel be used within six months, or reanalyzed after six months to ensure the fuel meets ASTM D6751 specifications. A longer shelf life is possible depending on the fuel composition and the use of storage-enhancing additives. The biodiesel should be stored in a cool and dry place.

Gelling

When biodiesel is cooled below a certain point, some of the molecules aggregate and form crystals. The fuel starts to appear cloudy once the crystals become larger than one quarter of the wavelengths of visible light - this is the cloud point (CP). As the fuel is cooled further these crystals become larger. The lowest temperature at which fuel can pass through a 45 micron filter is the cold filter plugging point (CFPP). As biodiesel is cooled further it will gel and then solidify. Within Europe, there are differences in the

CFPP requirements between countries. This is reflected in the different national standards of those countries. The temperature, at which pure (B100) biodiesel starts to gel, varies significantly and depends upon the mix of esters and therefore the feedstock oil used to produce the biodiesel. There are a number of commercially available additives that will significantly lower the pour point and cold filter plugging point of pure biodiesel. Winter operation is also possible by blending biodiesel with other fuel oils including low sulfur diesel fuel and diesel / kerosene

.

Another approach to facilitate the use of biodiesel in cold conditions is by employing a second fuel tank for biodiesel in addition to the standard diesel fuel tank. The second fuel tank can be insulated and a heating coil using engine coolant is run through the tank. The fuel tanks can be switched over when the fuel is sufficiently warm. A similar method can be used to operate diesel vehicles using straight vegetable oil.

Plant Design

Process

Process description

Water and nutrients are mixed and stored in the material tank. The nutrient solution is dozed to the Inner loop airlift photo bioreactor through a cooler; the cooler is used to make sure that the water is entered at the optimum temperature required for the bioreactor. In the bioreactor carbon dioxide gas is sparged from the bottom and continuous light source provides lights for photosynthesis. In the photo bioreactor algae and a diluted algal solution is passed through the ceramic membrane. The membrane concentrates the algal solution by removing water which is recycled to the nutrient tank. The concentrated algal solution then goes through the lysis heater which is a plate and frame heat exchanger at 90˚C, to ensure cell walls are rupture. Sufficient holding time is provided in the holding tank to ensure maximum rupturing of cell walls happens. The concentrated algal solution is then sent to the hexane solvent extraction where the hexane and oil stream are mixed, oil are dissolved hexane and move out from the top whereas the aqueous biomass solution exits from the bottom. The biomass solution is further sent to the gasifier where natural gas is produced and stored in tanks. The hexane-lipid solution is then heated to 85˚C through a heat exchanger and is sent to the flash column where oil and hexane are separated. Hexane is recycled back to the hexane tank, fresh hexane is also added to the tank and the resulting stream in continuously fed in the solvent extraction column. Methanol from the methanol tank and oil are fed in the Mcgyan reactor where simultaneous esterification and transesterification takes place at 300˚C. The exit stream from the reactor includes streams of biodiesel, glycerol, water, unreacted oil and unreacted methanol. The exit stream is flashed at a temperature of 175˚C, at which water and methanol are separated from rest of the streams. Water and methanol are further flashed at 75˚C for separation of components; methanol is recycled to the methanol tank. The solution of biodiesel, unreacted oil and glycerol are sent to a decanter. The heavy dispersed phase of glycerol is separated at the bottom and the continuous light phase of biodiesel and unreacted oil are fed to the distillation column. Distillation column is operated at vacuum conditions at 270˚C to produce ASTM standard biodiesel from the top, lipid are recycled back to the oil tank.

Process Flow Diagram

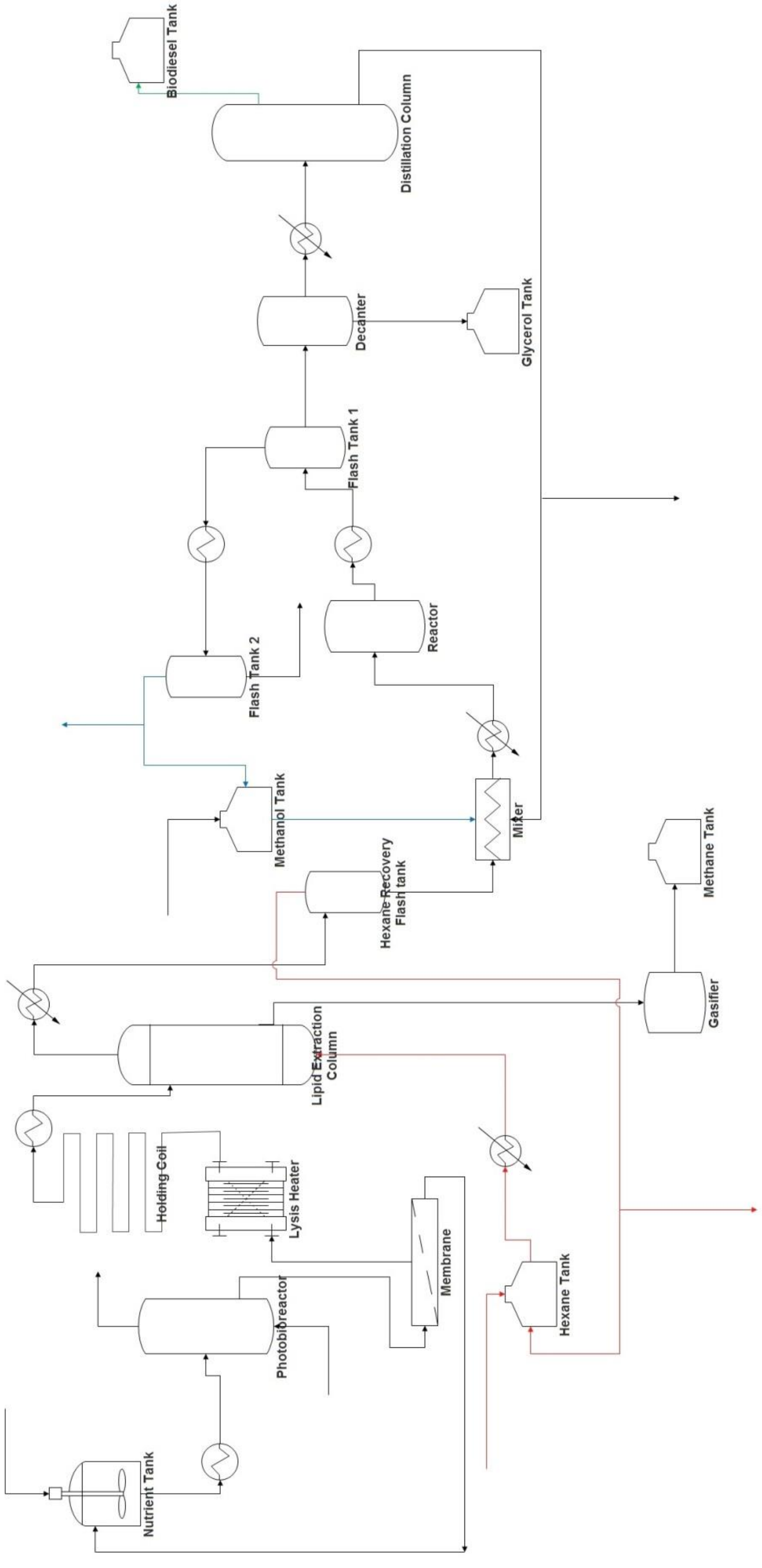


Figure 14: Process Flow Diagram

Instrumentation and Process Control

Nutrient Tank

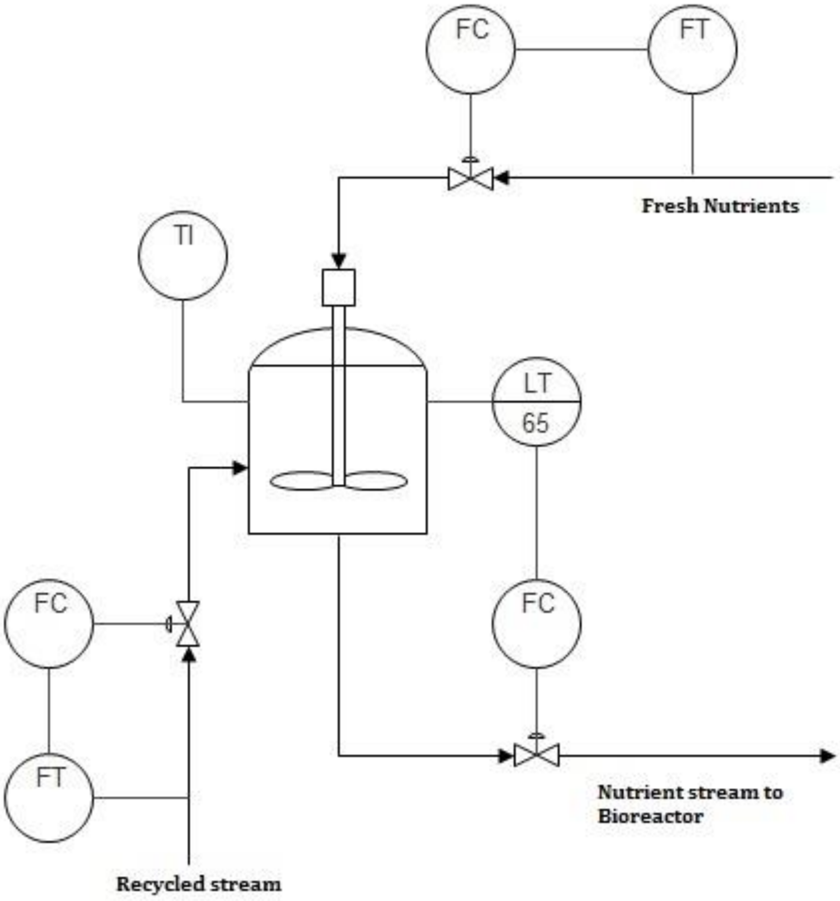


Figure 15: IPC of Nutrient Tank

Parameters being controlled are level of the tank and flow rate. Parameters being monitored are temperature

Bioreactor

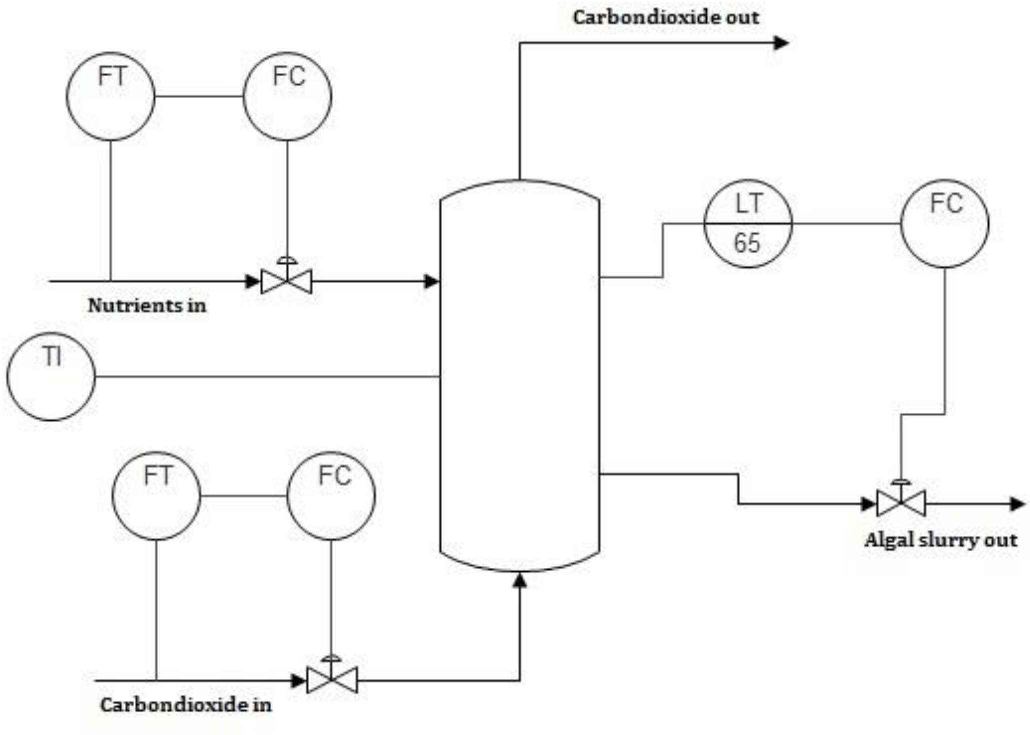


Figure 16: IPC of Bioreactor

Parameters being controlled are level of the tank and flow rate. Parameters being monitored are temperature

Cell Lysis Heater

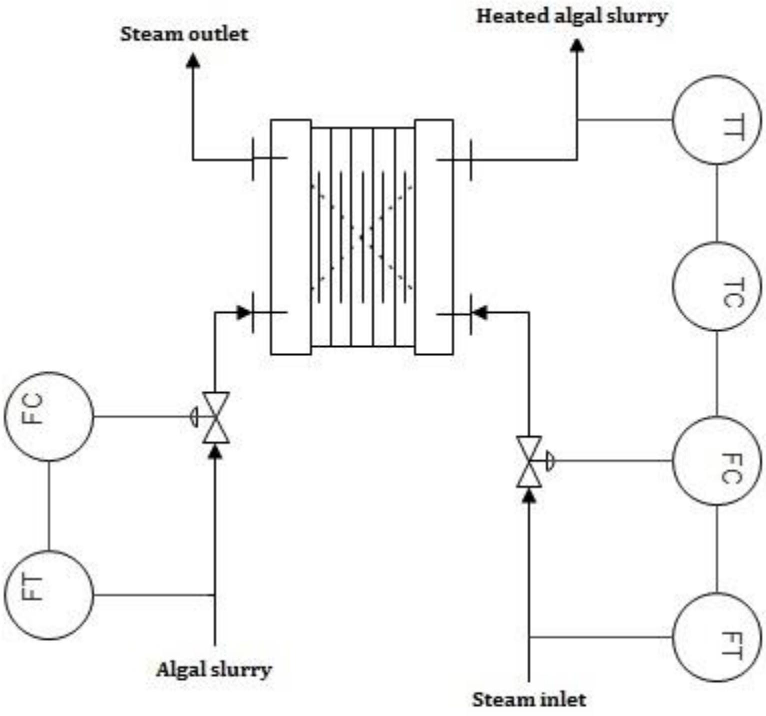


Figure 17: IPC of Cell Lysis Heater

Parameters being maintained are temperature and flow

Hexane Extraction Column

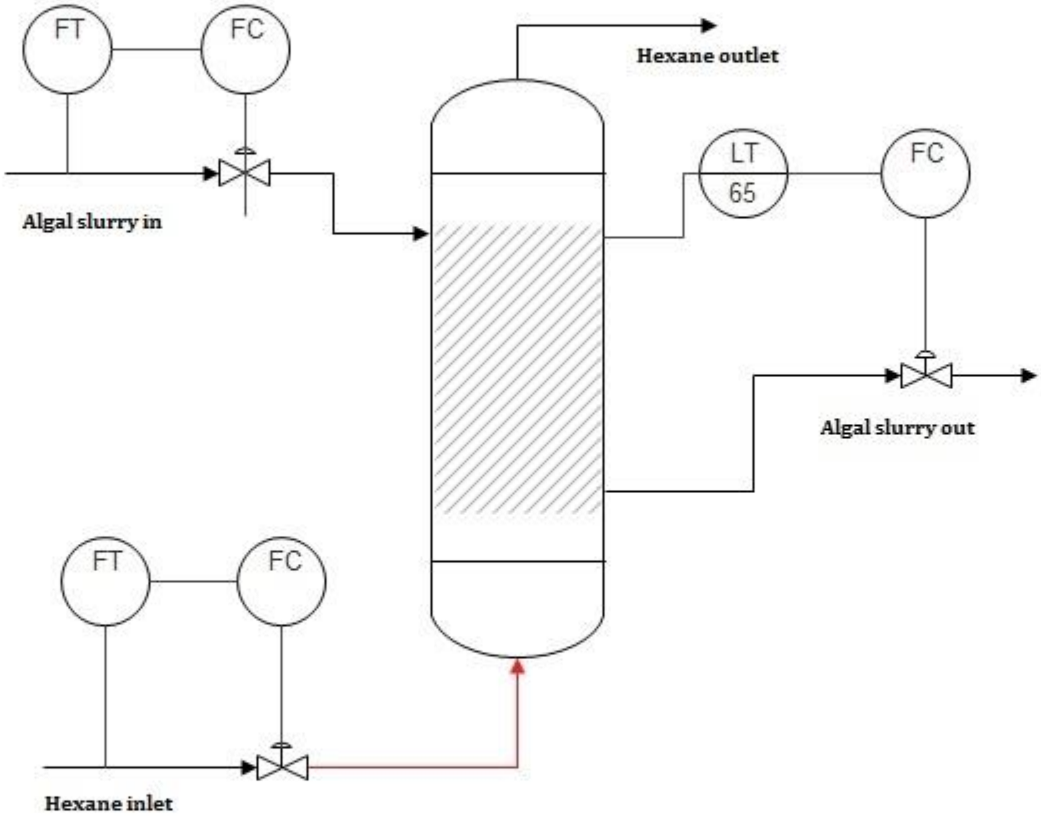


Figure 18: IPC of Hexane Extraction Column

Parameters being controlled are flow rate and level

Mcgyan Reactor

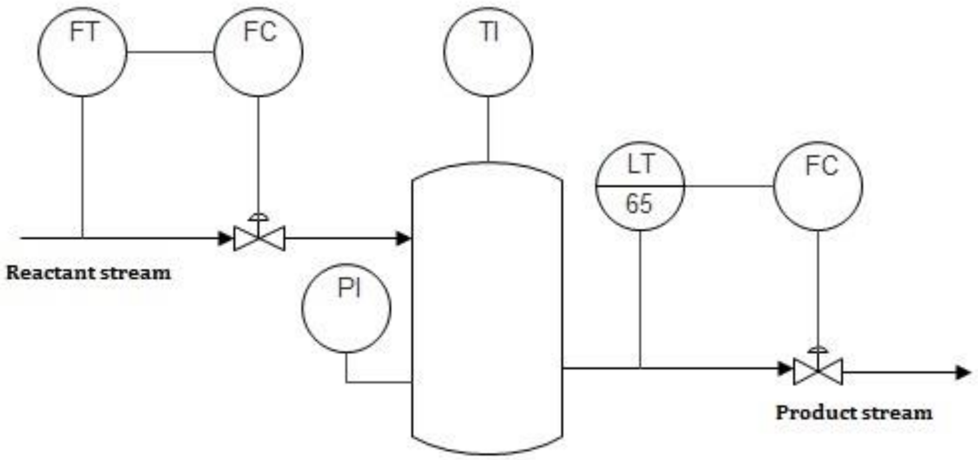


Figure 19: IPC of Mcgyan Reactor

Parameters being controlled are your flow rate. Parameters being monitored are temperature and pressure

Decanter

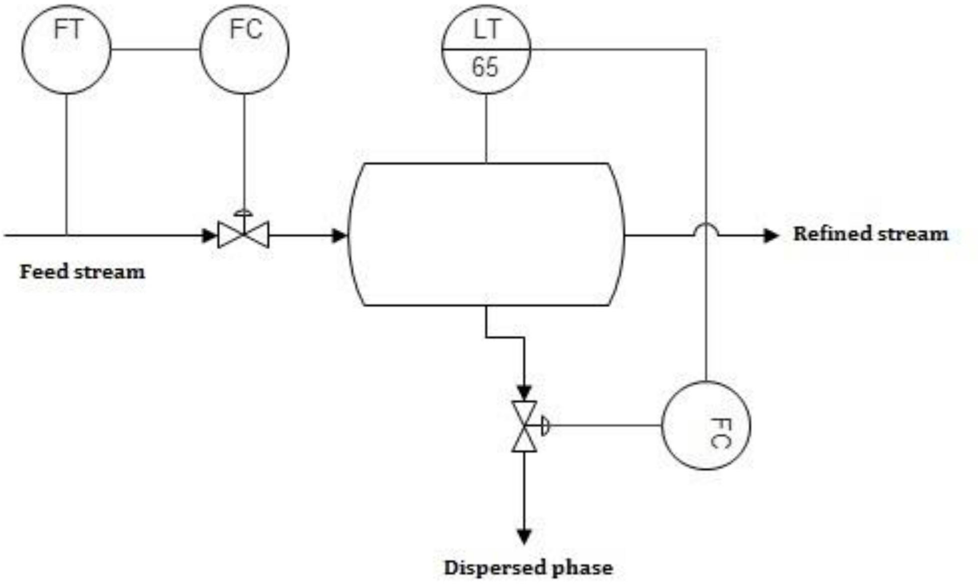


Figure 20: IPC of Decanter

Parameters being controlled are flow rate and level.

Distillation Column

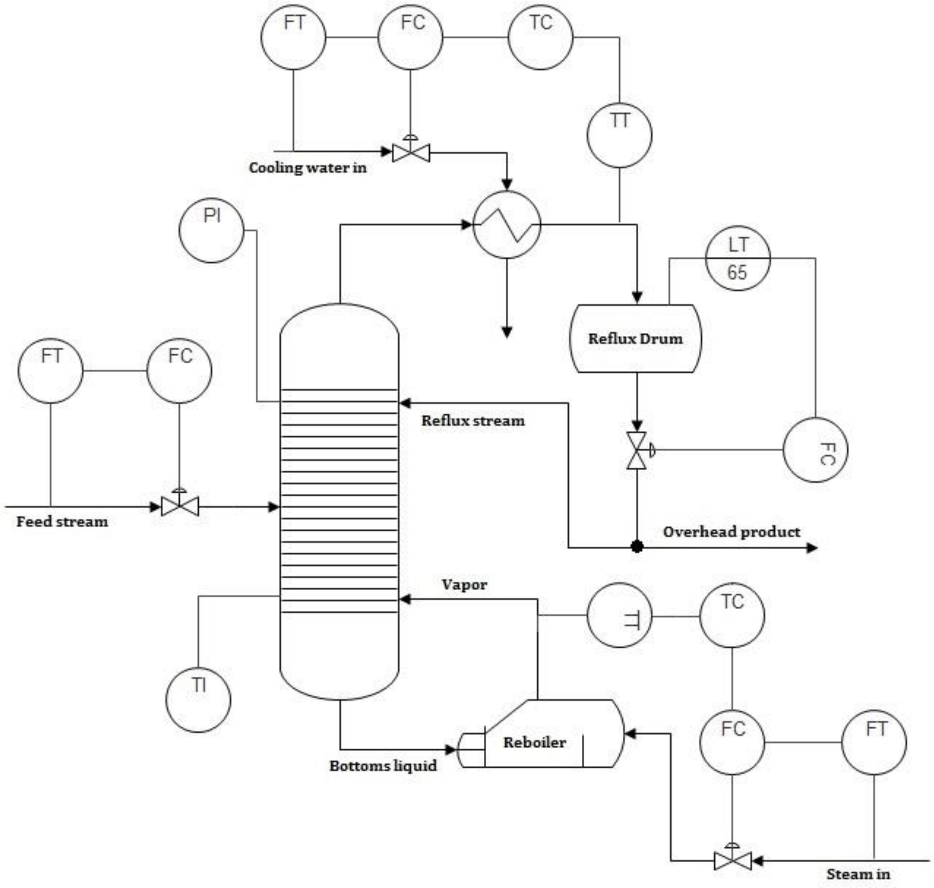


Figure 21: IPC of Distillation Column

Parameters being controlled are reflux drum level, temperature of column and flow in column. Parameter being monitored is pressure.

Energy and Material Balance

Nutrient Tank

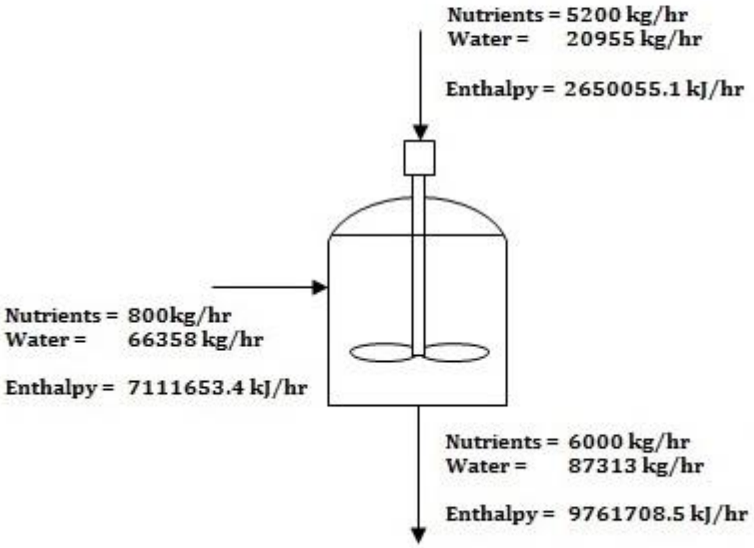


Figure 22: Nutrient Tank

Table 4: Nutrient Tank

Components	**Specific Heat Capacity kJ/kg**	**Fresh Nutrients (25˚C) kJ/hr**	**Recycled Nutrients (25˚C) kJ/hr**	**To Bioreactor (25˚C) kJ/hr**
Water	4.18	2189797.5	6934411	9124208.5
Nutrients		460257.6	177242.4	637500
Total		**2650055.1**	**7111653.4**	**9761708.5**

Specific Data:

Temperature of Nutrient tank = 25˚C

Formula: $\dot{m}C_p\Delta T$

Bioreactor

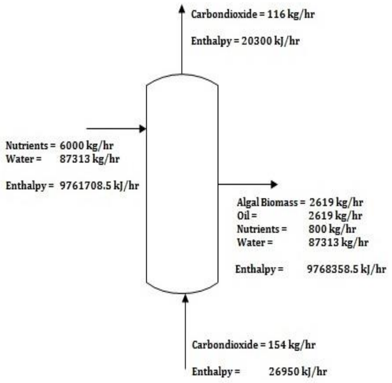


Figure 23: Bioreactor

Table 5: Bioreactor

Components	**Specific Heat Capacity kJ/kg**	**CO_2 In (35˚C) kJ/hr**	**CO_2 Out (35˚C) kJ/hr**	**From Nutrient tank (25˚C) kJ/hr**	**To Membrane (25˚C) kJ/hr**
Algal Biomass	4.25	-	-	-	278268.8
Oil	2.88	-	-	-	188568
Water	4.18	-	-	9124208.5	9124208.5
Nutrients		-	-	637500	177313.2
Carbon dioxide	5.0	26950	20300	-	-
Total		**26950**	**20300**	**9761708.5**	**9768358.5**

Specific Data:

Temperature of Bioreactor = 25˚C **Formula: $\dot{m}C_p\Delta T$**

Membrane

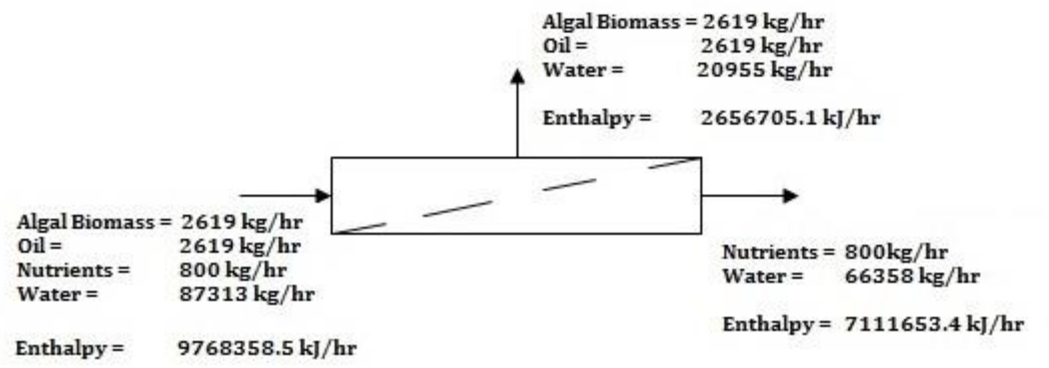


Figure 24: Membrane

Table 6: Membrane

Components	Specific Heat Capacity kJ/kg	From Bioreactor (25˚C) kJ/hr	To Cell Lysis Heater (25˚C) kJ/hr	Recycle to Nutrient tank (25˚C) kJ/hr
Algal Biomass	4.25	278268.8	278268.8	-
Oil	2.88	188568	188568	-
Water	4.18	9124208.7	2189797.3	6934411
Nutrients		-	-	177242.4
Total		**9768358.5**	**2656705.1**	**7111653.4**

Specific Data:

Temperature of Membrane = 25˚C **Formula: $\dot{m}C_p\Delta T$**

Lipid extraction column

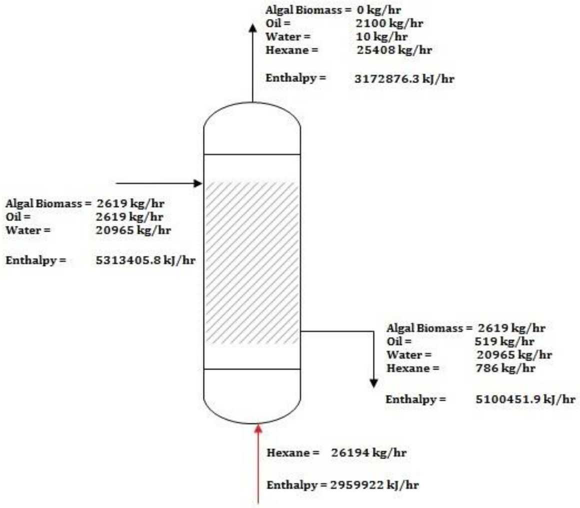


Figure 25: Lipid Extraction

Table 7: Lipid Extraction

Components	Specific Heat Capacity kJ/kg	Pre-cooler to Column (50˚C) kJ/hr	Hexane tank to Column (˚C) kJ/hr	Column to Flash tank (50˚C) kJ/hr	Waste (50˚C) kJ/hr
Algal Biomass	4.25	556617.2	-	-	556617.2
Oil	2.88	377193.6	-	301752	75442
Water	4.18	4379595	-	595	437900
Hexane	2.26	-	2959922	2871124.3	88797.7
Total		**5313405.8**	**2959922**	**3172876.3**	**5100451.9**

Specific Data:

Temperature of Hexane Column = 50˚C **Formula: $\dot{m}C_p\Delta T$**

Hexane Recovery Flash Tank

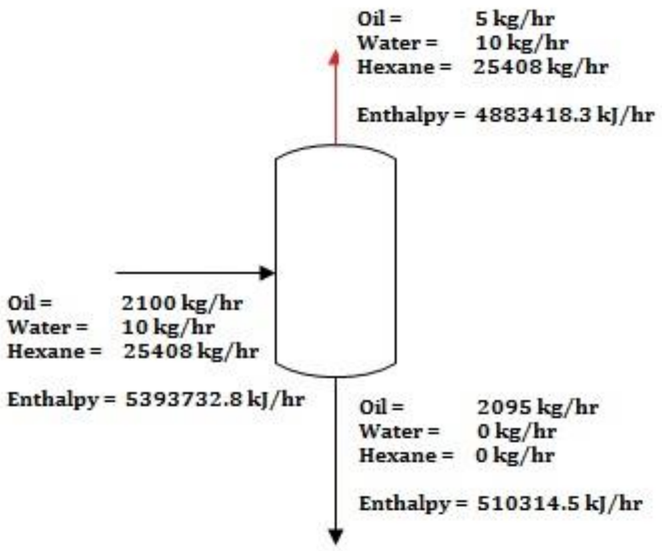


Figure 26: Hexane Recovery Flash Tank

Table 8: Hexane Recovery Flash Tank

Components	Specific Heat Capacity kJ/kg	From Recovery Pre-Heater (85˚C) kJ/hr	To Hexane tank (˚C) kJ/hr	To Mixer (85˚C) kJ/hr
Oil	2.88	512856	1224	512856
Water	4.18	2320	3553	-
Hexane	2.26	4880876.8	4880876.8	-
Total		**5399605.8**	**4884429.8**	**512856**

Specific Data:

Temperature of flash tank = 85˚C **Formula: $\dot{m}C_p\Delta T$**

Hexane Tank

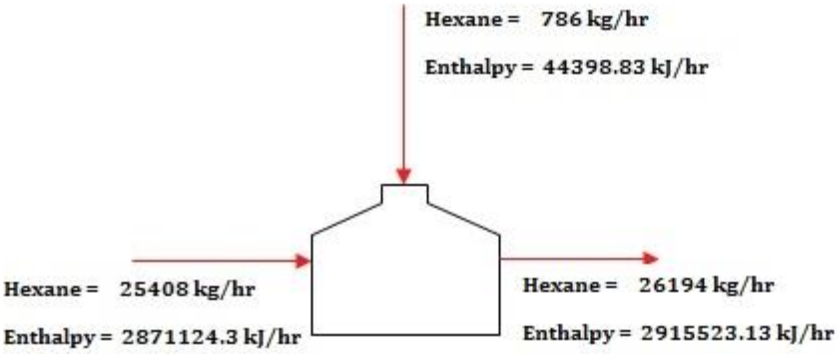


Figure 27: Hexane Tank

Table 9: Flash Tank

Components	**Specific Heat Capacity kJ/kg**	**From Cooler (50˚C) kJ/hr**	**Fresh Hexane (25˚C) kJ/hr**	**To Hexane Extraction Column (50˚C) kJ/hr**
Hexane	2.26	2871124.3	44398.83	2915523.13
Total		**2871124.3**	**44398.83**	**2915523.13**

Specific Data:

Temperature of Hexane tank = 50˚C

Formula: $\dot{m}C_p\Delta T$

Mixer

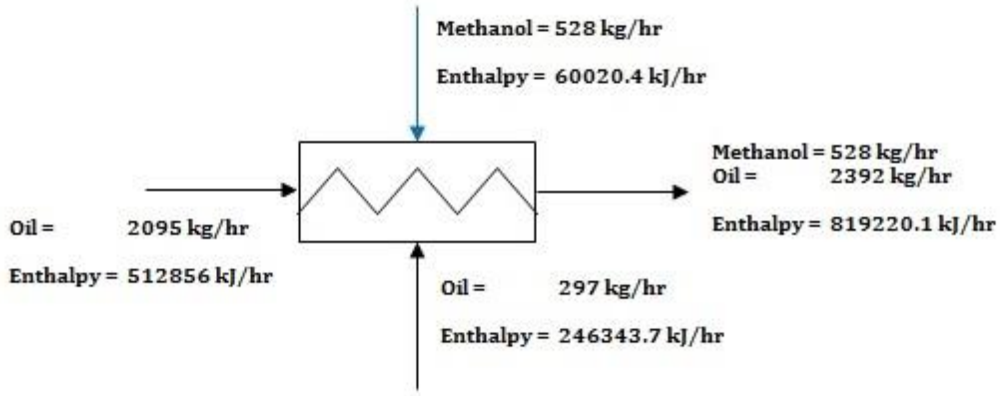


Figure 28: Mixer

Table 10: Mixer

Components	Specific Heat Capacity kJ/kg	From Hexane Recovery (85˚C) kJ/hr	Methanol tank to Mixer (˚C) kJ/hr	Distillation column recycle (288˚C) kJ/hr	Waste (50˚C) kJ/hr
Oil	2.88	512856	-	246343.7	759199.7
Methanol	1.92	-	60020.4	-	60020.4
Total		**512856**	**60020.4**	**246343.7**	**819220.1**

Specific Data:

Temperature of Mixer = 115˚C

Formula: $\dot{m}C_p\Delta T$

Mcgyan Reactor

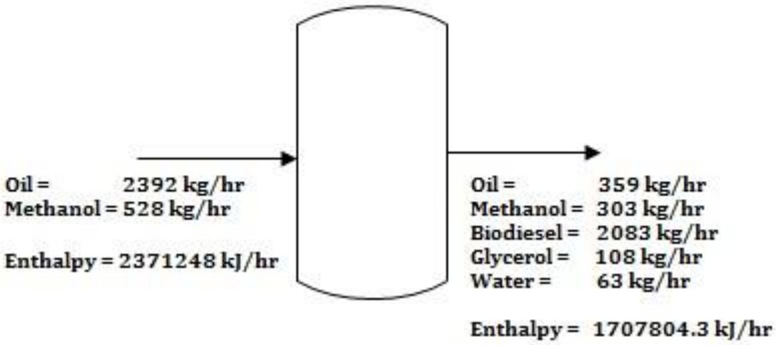


Figure 29: Mcgyan Reactor

Table 11: Mcgyan Reactor

Components	**Specific Heat Capacity kJ/kg**	**Pre-heater to Reactor (300˚C) kJ/hr**	**Reactor to Flash tank (290˚C) kJ/hr**
Oil	2.88	2066688	299836.8
Methanol	1.92	304560	168710.4
Water	4.18	-	76368.6
Biodiesel	1.8	-	1087326
Glycerol		-	75562.5
Total		**2371248**	**1707804.3**

Specific Data:

Temperature of Reactor = 300˚C

Formula: $\dot{m}C_p\Delta T$

Flash Tank 1

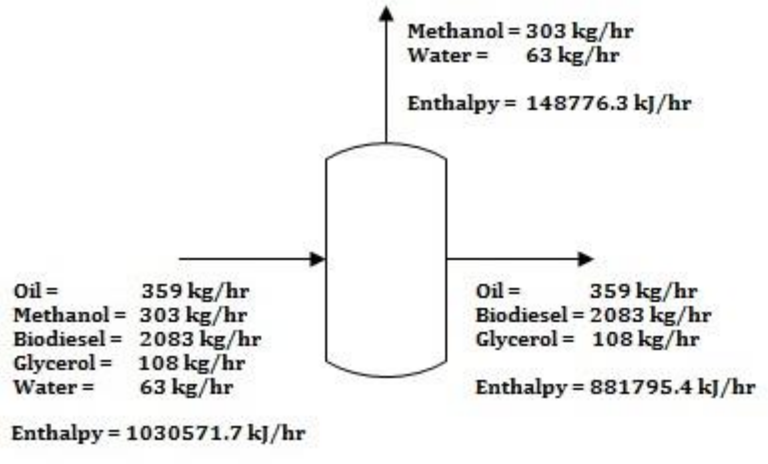


Figure 30: Flash Tank 1

Table 12: Flash Tank 1

Components	**Specific Heat Capacity kJ/kg**	**Let down valve to Flash tank (175˚C) kJ/hr**	**To Methanol recovery (175˚C) kJ/hr**	**To Decanter (175˚C) kJ/hr**
Oil	2.88	180936	-	180936
Methanol	1.92	10808	142808	-
Water	4.18	46084.5	46084.3	-
Biodiesel	1.8	656145	-	656145
Glycerol		45598.2	-	45598.4
Total		**1030571.7**	**148776.3**	**881795.4**

Specific Data:

Temperature of Flash tank = 175˚C **Formula: $\dot{m}C_p\Delta T$**

Flash Tank 2

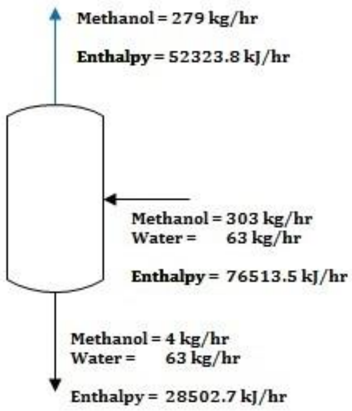


Figure 31: Flash Tank 2

Table 13: Flash Tank 2

Components	Specific Heat Capacity kJ/kg	From Flash tank (90˚C) kJ/hr	To Methanol tank (90˚C) kJ/hr	To Waste (90˚C) kJ/hr
Methanol	1.92	52358.4	52323.8	691.2
Water	4.18	24155.1	-	23700.6
Total		**76513.5**	**52323.8**	**28502.8**

Specific Data:

Temperature of Flash tank = 90˚C

Formula: $\dot{m}C_p\Delta T$

Methanol Tank

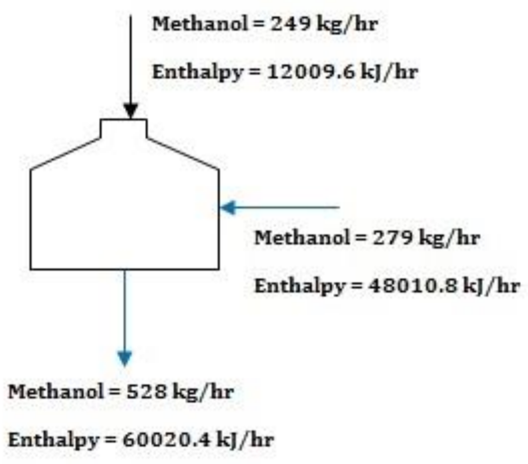


Figure 32: Methanol Tank

Table 14: Methanol Tank

Components	**Specific Heat Capacity kJ/kg**	**From Flash tank (90˚C) kJ/hr**	**Fresh Methanol (25˚C) kJ/hr**	**To Mixer (59˚C) kJ/hr**
Methanol	1.92	48010.8	12009.6	60020.4
Total		**48010.8**	**12009.6**	**60020.4**

Specific Data:

Temperature of Methanol tank = 59˚C **Formula: $\dot{m}C_p\Delta T$**

Decanter

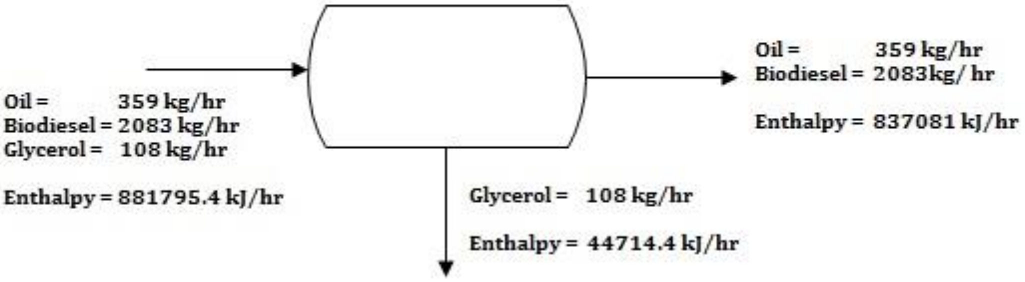


Figure 33: Decanter

Table 15: Decanter

Components	Specific Heat Capacity kJ/kg	From Flash tank (175˚C) kJ/hr	To Distillation Column (175˚C) kJ/hr	To Glycerol tank (175˚C) kJ/hr
Oil	2.88	180936	180936	-
Biodiesel	1.8	656145	656145	-
Glycerol		44714.4	-	44714.4
Total		**881795.4**	**837081**	**44714.4**

Specific Data:

Temperature of Decanter = 175˚C **Formula: $\dot{m}C_p\Delta T$**

Distillation Column

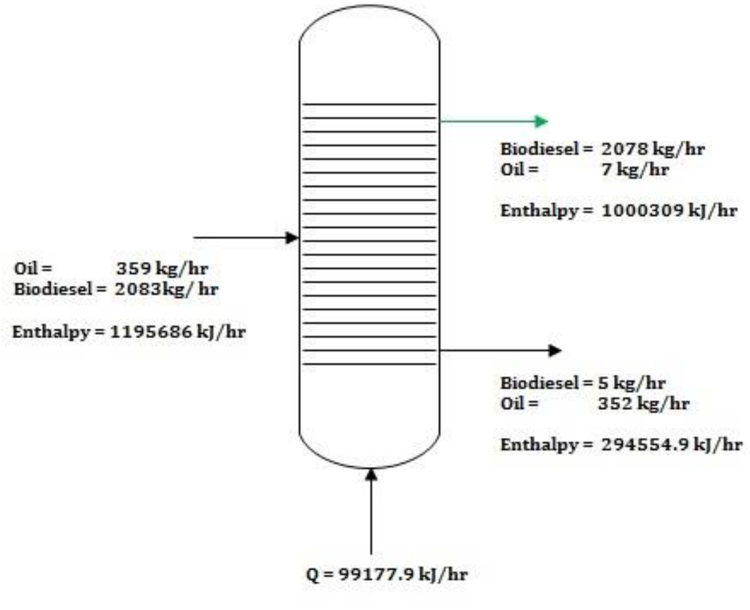


Figure 34: Distillation Column

Table 16: Distillation Column

Components	Specific Heat Capacity kJ/kg	From Pre-heater (250˚C) kJ/hr	To Biodiesel tank (266˚C) kJ/hr	Recycle oil to Mixer (288˚C) kJ/hr
Oil	2.88	258436	994946.4	291962.9
Biodiesel	1.8	937350	5362.6	2592
Total	**-**	**1195686**	**1000309**	**294554.9**

Specific Data:

Temperature of Distillation Column = 277˚C **Formula: $\dot{m}C_p\Delta T$**

Designing

Bioreactor

The inner loop airlift photo bioreactor is used for the algae growth. Carbon dioxide gas is sparged from the bottom of the bioreactor and nutrients are supplied along with water in the system. The bioreactor consists of two portions, dark and light. Algae growth depends of the frequency of dark and light regions. The outer region is exposed to day light or synthetic LED light incase in absence of light whereas the inner portion is kept dark. Stainless steel is used as the inner loop construction material to ensure no light passes through it. Algae are grown by photosynthesis process and sufficient growth is provided when it is circulated between the light and dark regions.

The dilution rate is equal to the growth rate of algae as it is grown continuously. The total lipid content in the specie being grown is 80% dry weight out of which 50% are useful lipids. A growth rate of 5g/liter/day is achieved.

The main advantage of the airlift photo bioreactor was its easy construction from material available on the market at very low cost. For the fluid-dynamic characterization of the reactor, the variation of gas holdup, superficial liquid velocity, and mass transfer coefficients with the superficial gas velocity required.

$$\frac{dx}{dt} = growth - output = 0$$

$$k_l x - \ Dx = 0$$

$$k_l = D$$

$$D = \frac{F}{V}$$

$$V = \frac{F}{D}$$

$$V = \frac{5238800}{60 \ x \ 0.833}$$

$$V = 104.8\ m^3$$

After 20% contingency

$$V = 125.76\ m^3$$

Using standard size reactors

D = 3 ft = 0.9144 m

H = 9 ft = 2.7432 m

$$N = \frac{V}{\frac{\pi D^2 H}{4}}$$

$$N = \frac{125.76}{\frac{\pi\ x\ 0.9144^2\ x\ 2.7432}{4}}$$

$$N = 69.88\ \cong 70$$

Use figure below to get superficial gas velocity (churn turbulent flow)

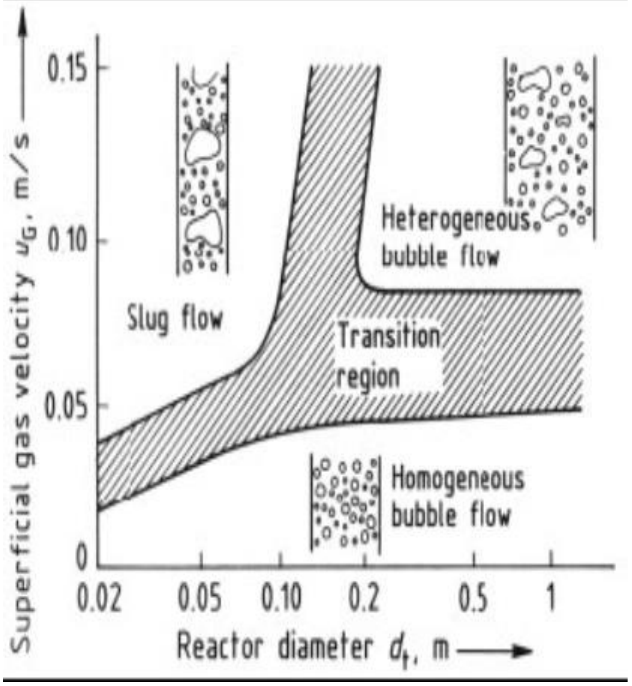


Figure 35: Superficial Gas Velocity

$$u_{gs} = 0.12\ m/s$$

$$volumetric\ flow\ rate = \frac{d^2\pi}{4}\ x\ u_{gs}$$

$$volumetric\ flow\ rate = \frac{0.9144^2\pi}{4}\ x\ 0.12$$

$$volumetric\ flow\ rate = 0.0788\ m^3/s$$

Using perforated plate

d = 0.7 ft = 2.134 m

Perforation = 40 %

Hole diameter = 0.0025 m

Bubble Diameter

Considering surface tension and buoyancy only

$$d_b = \left(\frac{6d_o\sigma}{\rho_l g}\right)^{\frac{1}{3}}$$

$$d_b = \left[\left(\frac{6d_o\sigma}{\rho_l g}\right)^{1.33} + \left(\frac{81\eta_l Q}{\pi\rho_l g}\right) + \left(\frac{135Q}{4\pi^2 g}\right)^{0.8}\right]^{0.25}$$

σ is surface tension of water

d_o is hole diameter

ρ_l is density of liquid

Q = volumetric flow rate

$$d_b = \left[\left(\frac{6\ x\ 2.5\ x\ 10^{-3}\ x\ 7.197\ x\ 10^{-3}}{1000\ x\ 9.81}\right)^{1.33} + \left(\frac{81\ x\ 8.9\ x\ 10^{-4}\ x\ 0.0788}{\pi\ x\ 1000\ x\ 9.81}\right) + \left(\frac{135\ x\ 0.0788}{4\pi^2\ x\ 9.81}\right)^{0.8}\right]^{0.25}$$

$$d_b = 0.0203\ m = 2.03\ cm$$

Gas Hold-up

$$V_T = \sqrt{\left(\frac{2\sigma}{\rho_l d_b}\right) + \left(\frac{\rho_l - \rho_g}{\rho_l}\right)\frac{g d_b}{2}}$$

$$V_T = \sqrt{\left(\frac{2\ x\ 71.97\ x\ 10^{-3}}{1000\ x\ 2\ x\ 10^{-2}}\right) + \left(\frac{1000 - 1.275}{1000}\right)\frac{9.81\ x\ 2\ x\ 10^{-2}}{2}}$$

$$V_T = 0.324\ m/s$$

$$\varepsilon_g = \frac{u_{gs}}{u_{gs} + V_T}$$

$$\varepsilon_g = \frac{0.12}{0.12 + 0.324}$$

$$\varepsilon_g = 0.27$$

Gas Liquid Mass Transfer

$$\frac{k_l a {D_c}^2}{D_l} = 0.6\left(\frac{\eta_l}{D_l \rho_l}\right)^{0.5}\left(\frac{g {D_c}^2 \rho_l}{\sigma}\right)^{0.62}\left(\frac{g {D_c}^3 {\rho_l}^2}{\eta_l}\right){\varepsilon_g}^{1.1}$$

Where

$k_l a$ is 1.48 x 10^{-2} /min (CO_2) from literature

D_c is column diameter

η_l is viscosity of liquid (water)

ε_g is gas hold up

$$D_l = 3.09\ x\ 10^{-4}\ m/s$$

CO_2

Calculated using

$$P_A = \ H{C_A}^*$$

Where

P_A is partial pressure of CO_2

H is Henry's constant

CA^* is concentration of CO_2 in liquid

Consumption

$$Q_A = \frac{\mu C_X}{Y_{X/CO_2}}$$

$Y_{X/CO2}$ is cell yield based on CO_2

μ is the growth rate

C_X is the cell density

Literature is 4% in air optimum

$$volumetric\ flow\ rate = 0.04\ x\ 0.0788$$

$$volumetric\ flow\ rate = 3.152\ x\ 10^{-3}\ m^3/h$$

Light intensity

$$Iz = \ I_o e^{-k_l C_{xz}}$$

Where

k_l is growth rate

C_x is cell density

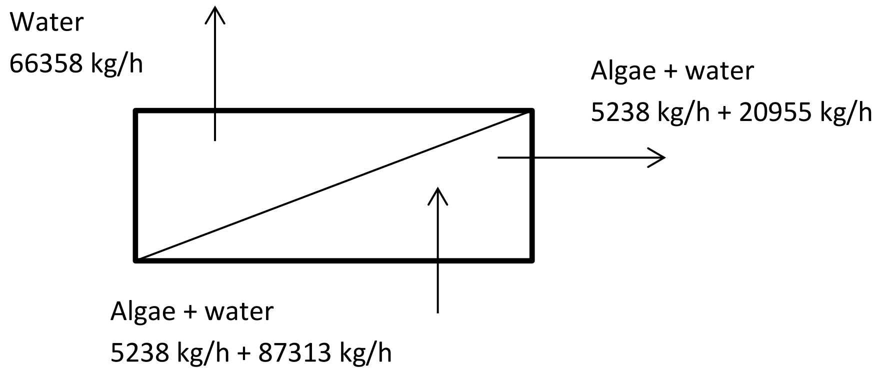
z is thickness of wall

Literature is 150 µmol.m^2/s optimum

Table 17: Bioreactor Specifications

Equipment	Airlift Photo Bioreactor
Height of reactor	2.7432 m
Diameter of reactor	0.9144 m
Operating temperature	25 ˚C
Operating pressure	1 atm
Material of construction	Industrial glass, stainless steel

Membrane



We have selected a ceramic membrane to filter our algae to reduce the water content. Filtration is the most energy efficient method of dewatering algae. We have selected a ceramic membrane based on the pore size requirement and because it is a low pressure system.

$$concentration\ required = \frac{5238}{20.955}$$

$$concentration\ required = 249.96\ kg/m^3h$$

P = 1 x 10^{-11} m/s

l = 500 µm

For industrially used ceramic membranes of pore size less than 20 µm

Thickness 500 µm

$$J = P(C_{io} - C_{il})$$

$$J = \frac{1\ x\ 10^{-5}(249.96 - 0)}{500\ x\ 10^{-6}}\ x\ 3600$$

$$J = 17997.12\ m^2kg/h$$

Area

$$A = \frac{J}{m_i}$$

$$A = \frac{17997.12}{92551}$$

$$A = 0.1945\ m^2$$

Table 18: Membrane Specifications

Equipment	Membrane
Area	0.1945 m^2
Thickness	500 µm
Pore size	20 µm
Operating temperature	25˚C
Operating pressure	1 atm
Material of construction	Ceramic

Cell Lysis Heater

$$LMTD = \frac{(T_{h\,in} - T_{c\,out}) - (T_{h\,out} - T_{c\,in})}{\ln \frac{(T_{h\,in} - T_{c\,out})}{(T_{h\,out} - T_{c\,in})}}$$

$$LMTD = \frac{(120 - 90) - (105 - 20)}{\ln \frac{(120 - 90)}{(105 - 20)}}$$

$$LMTD = 52.81\,°C$$

$$NTU_{(hot)} = \frac{T_{h\,in} - T_{h\,out}}{LMTD}$$

$$NTU_{(hot)} = \frac{120 - 105}{52.81}$$

$$NTU_{(hot)} = 0.28$$

$$NTU_{(cold)} = \frac{T_{c\,out} - T_{c\,in}}{LMTD}$$

$$NTU_{(cold)} = \frac{90 - 20}{52.81}$$

$$NTU_{(cold)} = 1.33$$

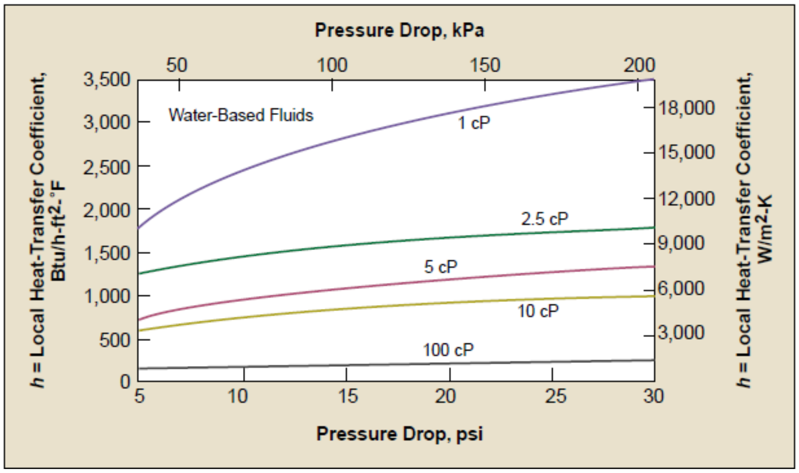


Figure 36: To Find Heat Transfer Coefficients

$$\frac{1}{U} = \frac{1}{h_{hot}} + \frac{\Delta X}{k} + \frac{1}{h_{cold}}$$

h_{hot} is 14000 W/m^{2}K

h_{cold} is 15000 W/m^{2}K

$$\frac{1}{U} = \frac{1}{15000} + \frac{5\ x\ 10^{-4}}{49.19} + \frac{1}{14000}$$

$$U = 6745\ W/m^2Kh$$

$$Q = mc_p\Delta T$$

$$Q = 20955\ x\ 4.18\ x\ 70$$

$$Q = 6131433\ J$$

$$A = \frac{Q}{U.LMTD}$$

$$A = \frac{6131433}{6745\ x\ 52.81}$$

$$A = 17.2\ m^2$$

Hexane Extraction

The invention uses an appropriate solvent to extract oil from relatively fine particles in aqueous slurry Without the need to dry the slurry or reform the material to create larger particles. Another technical advantage of this invention is that it provides a novel method for separating edible oil from biological material that overcomes the problems of conventional methods. When disrupting the biomass in an aqueous phase and extracting without further drying, the fines stay in the aqueous phase and do not contaminate the solvent. Therefore, additional treatment of the solvent to remove the fines may be avoided. Moreover, hexane can be more easily removed from the aqueous liquid. Although hexane is soluble in water up to 3%, this hexane may be easily removed by heating the aqueous liquid. These and other technical advantages are provided through one or more of the following embodiments. In one embodiment, a method for separating oil from biological material includes: providing biological material containing oil in an aqueous suspension; contacting a solvent. With the aqueous suspension of biological material, the solvent being essentially immiscible in Water; collecting the solvent, Which now contains oil extracted from the aqueous suspension of biological material; and separating the oil from the solvent. Typically, the aqueous slurry Will have less than 50% solids (W/W), preferably less than 35% solids. (*United States Patent*)

Table 19: Hexane Extraction

		m (kg/h)	MR (kg/kmol)	moles	density (kg/m^3)	volume (m^3/h)
organic	Hexane	26194	86.1	304.2	654.3	40.0
solute	Lipid	2619	856.42	3.06	920	2.85
aqueous	Water	20955	18	1164.17	999.97	20.96

$$[S_{org}] = \frac{moles_s}{moles_{org} + moles_s}$$

$$[S_{org}] = \frac{3.06}{304.2 + 3.06}$$

$$[S_{org}] = 0.00996$$

$$[S_{aq}] = \frac{moles_s}{moles_{aq} + moles_s}$$

$$[S_{aq}] = \frac{3.06}{1164.17 + 3.06}$$

$$[S_{aq}] = 0.002622$$

$$D = \frac{[S_{org}]}{[S_{aq}]}$$

$$D = \frac{0.00996}{0.002622}$$

$$D = 3.799$$

$$(q_{aq})_1 = \frac{V_{org}}{DV_{aq} + V_{org}}$$

$$(q_{aq})_1 = \frac{20.96}{3.799\ x\ 40 + 20.96}$$

$$(q_{aq})_1 = 0.12$$

$$efficiency = 1 - (q_{aq})_1$$

$$efficiency = 1 - 0.12$$

$$efficiency = 0.88\ or\ 88\ \%$$

Find volume of reactor

$$total\ volume\ flow = V_{org} + V_s + V_{aq}$$

$$total\ volume\ flow = 40 + 2.85 + 20.96$$

$$total\ volume\ flow = 63.81\ m^3/h$$

$$V = total\ volume\ flow\ x\ residence\ time$$

$$V = 63.81\ x\ 1$$

$$V = 63.81\ m^3$$

20% contingency

$$V = 76.57\ m^3$$

Aspect ratio 1:7.5

$$H = 7.5D$$

$$V = \frac{\pi D^2 H}{4}$$

$$V = \frac{7.5\pi D^3}{4}$$

$$D = \sqrt[3]{\frac{4V}{7.5\pi}}$$

$$D = \sqrt[3]{\frac{4\ x\ 76.57}{7.5\pi}}$$

$$D = 2.35\ m$$

$$H = 7.5D$$

$$H = 7.5\ x\ 2.35$$

$$H = 17.63\ m$$

Volume of packing is 75% of column

$$V = 57.429\ m^3$$

Table 20: Hexane Extraction Specification

Equipment	Solvent Extraction Unit
Equipment	Hexane Extraction
Volume	76.57 m^3
Height	17.63 m
Diameter	2.35 m
Operating temperature	50 ˚C
Operating pressure	1 atm
Material of construction	Mild Steel

Hexane Flash

This first flash is placed after the solvent traction column to separated oil from hexane using flashing. The flash tank operates at 75 °C to ensure maximum flashing occurs. A flash tank is used as the mixture is miscible and other separation techniques will not apply.

Hexane

$$\log P_H = a - \frac{b}{T + c}$$

a = 4.003

b = 1171.53

c = -48.784

At 85 °C

$$\log P_H = 4.003 - \frac{1171.53}{(85 + 273) - 48.784}$$

$$P_H = 1.64\ atm$$

Oil

$$\log P_O = a - \frac{b}{T + c}$$

a = 10.752

b = 3424

c = -69.43

At 85 °C

$$\log P_O = 10.752 - \frac{3424}{(85 + 273) - 69.43}$$

$$P_O = 0.077\ atm$$

$$P_{sys} = x_H P_H + x_O P_O$$

$$P_{sys} = 0.97\ x\ 1.64 + 0.03\ x\ 0.077$$

$$P_{sys} = 1.59\ atm$$

$$P_{sys} > P_{atm}$$

Using formula

$$V = \frac{F}{\rho}\ x\ 2\tau$$

$$V = \frac{2619.4 + 26194}{879.4}\ x\ 2\left(\frac{4}{60}\right)$$

$$V = 4.281\ m^3$$

After 20% contingency

$$V = 5.137\ m^3$$

Using aspect ratio 4:1

$$H = 4D$$

$$V = \frac{\pi D^2 H}{4}$$

$$V = \pi D^3$$

$$D = \sqrt[3]{\frac{V}{\pi}}$$

$$D = \sqrt[3]{\frac{5.137}{\pi}}$$

$$D = 1.178\ m$$

$$H = 4D$$

$$H = 4\ x\ 1.178$$

$$H = 4.712\ m$$

Table 21: Hexane Flash Tank Specifications

Equipment	Hexane Flash tank
Volume	5.137 m3
Height	4.712 m
Diameter	1.178 m
Operating temperature	85˚c
Operating pressure	1 bar
Material of construction	Mild Steel

Mcgyan Reactor

The Mcgyan Reactor has been selected due to its high conversion efficiency based on the catalyst selection. The Mcgyan reactor is a fixed bed catalytic reactor with zirconia oxide as a catalyst. The specialty of the catalyst is that it has high conversion efficiency; it is not poisoned by water, giving it a long life and converts both free fatty acids and triglycerides.

$$space\ velocity\ = \frac{volumetric\ feed\ rate}{reactor\ volume}$$

$$\tau = \frac{1}{space\ velocity} = \frac{reactor\ volume}{volumetric\ feed\ rate}$$

$$F = F_{methanol} + F_{oil}$$

$$F = \frac{m_{methanol}}{\rho_{methanol}} + \frac{m_{oil}}{\rho_{oil}}$$

$$F = \frac{528}{791} + \frac{2392}{920}$$

$$F = 3.2675\ m^3/h$$

From literature the value of τ = 1 min = 0.0167 h

$$reactor\ volume\ = \ \tau\ x\ volumetric\ feed\ rate$$

$$reactor\ volume = 0.0167\ x\ 3.2675$$

$$reactor\ volume\ = 0.0545\ m^3$$

Taking aspect ratio 3:1

$$L = 3D$$

$$V = \frac{\pi D^2 L}{4}$$

$$V = \frac{3\pi D^3}{4}$$

$$D = \sqrt[3]{\frac{4V}{3\pi}}$$

$$D = \sqrt[3]{\frac{4\ x\ 0.0545}{3\pi}}$$

$$D = 0.285\ m$$

$$L = 3D$$

$$L = 3\ x\ 0.285$$

$$L = 0.850\ m$$

Weight of catalyst

$$W = \rho_{catalyst}\ x\ volume\ of\ reactor$$

$$W = 5680\ x\ 0.0545$$

$$W = 309.6\ kg$$

Table 22: Mcgyan Reactor Specifications

Equipment	Fixed Bed Mcgyan Reactor
Volume	0.0545 m3
Height	0.850 m
Diameter	0.285 m
Operating temperature	300˚c
Operating pressure	17 atm
Catalyst	Zirconia Oxide
Weight of Catalyst	309.6 kg
Material of construction	Carbon Steel, Refractory Material

Flash Tank 1

After cooling down the exit stream of Mcgyan reactor, it is sent to a flash column. Before sending this stream to the flash column the Antoine constant of each of its constituent was selected and using the antinode equation we were able to determine the temperature at which this stream would flash at 1 atm (atmospheric pressure). We are using a flash column over here in order to separate our solvents which are methanol and water from biodiesel, unreacted oil and glycerol. This is done via flashing because both these solvents are extremely volatile and the difference in the boiling point of these components and the rest of the components present is too large to employ a normal distillation column. Once the stream at the temperature of 175°C will enter the flash column flashing will take place which will allow water and methanol to be removed from the top as a gaseous phase and the rest of the components which included bio diesel, glycerin and unconverted oil are removed as liquid from the bottom.

Biodiesel

$$\log P_B = a - \frac{b}{T+c}$$

a = 5.2252

b = 2723.181

c = -91.822

At 175 °C

$$\log P_B = 5.2252 - \frac{2723.181}{(175+273) - 91.822}$$

$$P_B = 3.798\ x\ 10^{-3}\ atm$$

Glycerol

$$\log P_B = a - \frac{b}{T+c}$$

a = 3.93737

b = 1411.531

c = -200.566

At 175 °C

$$\log P_G = 3.93737 - \frac{1411.531}{(175 + 273) - 200.566}$$

$$P_G = 0.0171\, atm$$

Oil

$$\log P_O = a - \frac{b}{T + c}$$

a = 5.04842

b = 2555.604

c = -127.258

At 175 °C

$$\log P_O = 5.04842 - \frac{2555.604}{(175 + 273) - 127.258}$$

$$P_O = 1.2040\; x\; 10^{-3}\, atm$$

Methanol

$$\log P_M = a - \frac{b}{T + c}$$

a = 5.15853

b = 1569.613

c = -34.846

At 175 °C

$$\log P_M = 5.15853 - \frac{1569.613}{(175 + 273) - 34.846}$$

$$P_M = 22.8786\ atm$$

Water

$$\log P_W = a - \frac{b}{T + c}$$

a = 3.55959

b = 643.748

c = -198.043

At 175 °C

$$\log P_W = 3.55959 - \frac{643.748}{(175 + 273) - 198.043}$$

$$P_O = 9.6417\ atm$$

$$P_{sys} = x_B P_B + x_G P_G + x_O P_O + x_M P_M + x_W P_W$$

$$P_{sys} = 0.48\ x\ 3.79\ x\ 10^{-3} + 0.06\ x\ 0.017 + 0.06\ x\ 1.204\ x\ 10^{-3} + 0.11\ x\ 22.878 + 0.29\ x\ 9.641$$

$$P_{sys} = 5.315\ atm$$

$$P_{sys} > P_{atm}$$

Using formula

$$V = \frac{F}{\rho}\ x\ 2\tau$$

$$V = \frac{2916}{815.5}\ x\ 2\left(\frac{4}{60}\right)$$

$$V = 0.4767\ m^3$$

After 20% contingency

$$V = 0.5721\ m^3$$

Using aspect ratio 4:1

$$H = 4D$$

$$V = \frac{\pi D^2 H}{4}$$

$$V = \pi D^3$$

$$D = \sqrt[3]{\frac{V}{\pi}}$$

$$D = \sqrt[3]{\frac{0.5721}{\pi}}$$

$$D = 0.5668\ m$$

$$H = 4D$$

$$H = 4\ x\ 0.5668$$

$$H = 2.2672\ m$$

Table 23: Flash Tank 1 Specifications

Equipment	Flash tank 1
Volume	0.5721 m^3
Height	2.2672 m
Diameter	0.5668 m
Operating temperature	175 ˚C
Operating pressure	1 atm
terial of construction	Mild Steel

Flash tank 2

Methanol is further separated from water using another flash tank which is operated at 90°C to extract pure methanol. The removed solvent is then recovered and sent back to the reactor. As our process requires us to use solvents well in excess hence most of the solvent added is recovered which further optimizes our process and reduces our cost. Methanol is used as a reagent in our process. Flash (or partial) evaporation is the partial vapor that occurs when a saturated liquid stream undergoes a reduction in pressure by passing through a throttling valve or other throttling device. This process is one of the simplest unit operations. If the throttling valve or device is located at the entry into a pressure vessel so that the flash evaporation occurs within the vessel, then the vessel is often referred to as a flash drum. If the saturated liquid is a multi-component liquid, the flashed vapor is richer in the more volatile components than is the remaining liquid

Methanol

$$\log P_M = a - \frac{b}{T + c}$$

a = 5.15853

b = 1569.613

c = -34.846

At 90 °C

$$\log P_M = 5.15853 - \frac{1569.613}{(90 + 273) - 34.846}$$

$$P_M = 2.3734\ atm$$

Water

$$\log P_W = a - \frac{b}{T + c}$$

a = 3.55959

b = 643.748

c = -198.043

At 90 °C

$$\log P_W = 3.55959 - \frac{643.748}{(90+273) - 198.043}$$

$$P_O = 0.4540\ atm$$

$$P_{sys} = x_M P_M + x_W P_W$$

$$P_{sys} = 0.3\ x\ 2.3734 + 0.7\ x\ 0.4540$$

$$P_{sys} = 1.0298\ atm$$

$$P_{sys} > P_{atm}$$

Using formula

$$V = \frac{F}{\rho}\ x\ 2\tau$$

$$V = \frac{302.8 + 64.7}{827.6}\ x\ 2\left(\frac{4}{60}\right)$$

$$V = 0.059\ m^3$$

After 20% contingency

$$V = 0.071\ m^3$$

Using aspect ratio 4:1

$$H = 4D$$

$$V = \frac{\pi D^2 H}{4}$$

$$V = \pi D^3$$

$$D = \sqrt[3]{\frac{V}{\pi}}$$

$$D = \sqrt[3]{\frac{0.071}{\pi}}$$

$$D = 0.2828\ m$$

$$H = 4D$$

$$H = 4\ x\ 0.2828$$

$$H = 1.1312\ m$$

Table 24: Flash Tank 2 Specifications

Equipment	Flash tank 2
Volume	0.071 m^3
Height	0.2828 m
Diameter	1.1312 m
Operating temperature	90 ˚C
Operating pressure	1 atm
Material of construction	Mild Steel

Decanter

The operating principle of a decanter is based on gravitational separation. Naturally, a component with a higher density would fall to the bottom of a mixture, while the less dense component would be suspended above it. A decanter centrifuge increases the rate of settling through the use of continuous rotation, producing a gravitational force between 1000 to 4000 times that of normal gravitational force. This reduces the settling time of the components by a large magnitude, whereby mixtures previously having to take hours to settle can be settled in a matter of seconds using a decanter centrifuge. This form of separation enables more rapid and controllable results.

Decanter centrifuges have a clean appearance and have little to no odor problems. Not only is the device easy to install and fast at starting up and shutting down, but also only requires a small area for operation compared to other competitive processes. The decanter centrifuge is versatile as different lengths of the cylindrical bowl section and the cone angle can be selected for different applications. Also the system can be pre-programmed with various design curves to predict the sludge type, while some competitive processes, such as a belt filter press, cannot change the belt type to operate for different sludge types. Its versatility allows the machine to have various functions such as operating for thickening or dewatering. The machine can operate with a higher throughput capacity than smaller machines. This also reduces the number of units required. The device is simple to optimize and operate as it has few major variables and reliable feedback information. The decanter centrifuge has reduced labor costs compared to other processes, as it requires low continuous maintenance and operator attention. Compared to some competitive process such as the belt filter process, the decanter centrifuge has more process flexibility and higher levels of performance.

Dispersed phase is glycerol

Continuous phase is biodiesel and oil

$$u_d = \frac{d^2 g(\rho_d - \rho_c)}{18\mu_c}$$

$\mu c = 1.025$ cp $= 1.025 \times 10^{-3}$ Ns/m^2

d = 150 µm = 1.5 x 10^{-4} m

ρc = 766.4 kg/m^3

ρd = 1022 kg/m^3

$$u_d = \frac{(1.5\ x\ 10^{-4})^2\ x\ 9.81\ x\ (1022 - 766.4)}{18\ x\ 1.025\ x\ 10^{-3}}$$

$$u_d = 3.063\ x\ 10^{-3}\ m/s$$

$$u_c = 2083 + 358.8$$

$$u_c = 2441.8\ kg/h$$

$$L_c = \frac{m_c}{\rho_c}$$

$$L_c = \frac{2441.8}{766.4}\ x\ \frac{1}{3600}$$

$$L_c = 8.85\ x\ 10^{-4}\ m^3/s$$

$$A_c = \frac{L_c}{u_d}$$

$$A_c = \frac{8.85\ x\ 10^{-4}}{3.06\ x\ 10^{-3}}$$

$$A_c = 0.289\ m^2$$

After 20% contingency

$$A = 0.3468\ m^2$$

$$A = \pi r^2$$

$$r = \sqrt{\frac{A}{\pi}}$$

$$r = \sqrt{\frac{0.3468}{\pi}}$$

$$r = 0.33\ m$$

Aspect ratio 1: 2

$$H = 2r$$

$$H = 2\ x\ 0.33$$

$$H = 0.66\ m$$

Residence time

$$residence\ time = \frac{h}{u_d}$$

$$residence\ time = \frac{0.66}{3.06\ x\ 10^{-3}}$$

$$residence\ time = 217\ s = 3.6\ min$$

Table 25: Decanter Specifications

Equipment	Decanter
Volume	0.226 m^3
Height	0.66 m
Diameter	0.66 m
Operating temperature	175˚C
Operating pressure	1 atm
Material of construction	Mild Steel

Distillation

Biodiesel and unconverted oil is then sent to a distillation column so as to attain diesel having a purity of 99.7% so as to comply with the ASTM standards. Here we are using vacuum distillation where biodiesel is extracted from the top and unconverted oil is extracted from the bottom. By using vacuum distillation product is obtained at a lower temperature reducing thermal decomposition. This allows us to obtain biodiesel of the required purity and ASTM standards and also obtain unconverted vegetable oil which can be recycled and further sent to the reactor for direct trans-esterification in order to optimize our process and reduce our cost. The diesel obtained from the top of the distillation column is marketable and according to the standards ready to be sold to our buyers at the cost later calculated.

Distillation is simply defined as a process in which a liquid or vapor mixture of two or more substances is separated into its component fractions of desired purity, by the application and removal of heat. The process is based on the fact that the vapor of a boiling mixture will be richer in the components that have lower boiling points. Hence, when this vapor is cooled and condensed, the condensate will contain more volatile components. At the same time, the original mixture will contain more of the less volatile material

Moles of feed

Biodiesel

$$F_B = \frac{m_B}{MR_B}$$

$$F_B = \frac{2083}{296}$$

$$F_B = 7.037\ moles$$

Oil

$$F_O = \frac{m_O}{MR_O}$$

$$F_O = \frac{358.8}{856}$$

$$F_O = 0.4192\,moles$$

Mole balance of both components gives

Biodiesel

$$F_B = D_B + B_B$$

$$7.037 = 0.997D + 0.01B$$

Oil

$$F_O = D_O + B_O$$

$$0.4192 = 0.003D + 0.99B$$

Solving simultaneously gives

$$D = 7.054\,moles$$

$$B = 0.40206\,moles$$

Distillation occurs under reduced pressure of 100mmHg

New boiling points are calculated by the Clausius-Clapeyron equation

$$\ln\left(\frac{P_1}{P_2}\right) = \frac{\Delta H}{R}\left(\frac{1}{T_1} - \frac{1}{T_2}\right)$$

Where

ΔH is the enthalpy change of formation

R is the gas constant

By the Clausius-Clapeyron equation

Biodiesel

Boiling point at 760 mmHg is 340 °C

Boiling point at 100 mmHg is 266 °C

Oil

Boiling point at 760 mmHg is 360 °C

Boiling point at 100 mmHg is 288 °C

The column operating temperature is 277 °C

The relative volatility is found by means of the partial pressures.

Biodiesel

$$\log P_B = a - \frac{b}{T + c}$$

a = 9.9155

b = 2583.52

c = -96.15

$$\log P_B = 9.9155 - \frac{2583.52}{(277 + 273) - 96.15}$$

$$P_B = 1.6712$$

Oil

$$\log P_O = a - \frac{b}{T + c}$$

a = 10.7518

b = 34.24

c = -69.43

$$\log P_O = 10.7518 - \frac{34.24}{(277 + 273) - 69.43}$$

$$P_O = 4235.7$$

Relative volatility

$$\propto = \frac{P_B}{P_O}$$

$$\propto = \frac{16712.7}{4235.7}$$

$$\propto = 3.95$$

Fenske equation for number of plates

$$N = \frac{\log\left[\left(\frac{D_B}{1 - D_B}\right) + \left(\frac{1 - B_B}{B_B}\right)\right]}{\log \propto}$$

$$N = \frac{\log\left[\left(\frac{0.997}{1 - 0.997}\right) + \left(\frac{1 - 0.01}{0.01}\right)\right]}{\log 3.95}$$

$$N = 7.57 \cong 8$$

By the Gilliland equation

$$\frac{N - N_{min}}{N + 1} = 0.75\left[1 - \left(\frac{R - R_{min}}{R + 1}\right)^{0.566}\right]$$

From literature, R = 2

$$R = R_{min} x\ 1.5$$

$$R_{min} = \frac{R}{1.5}$$

$$R_{min} = \frac{2}{1.5}$$

$$R_{min} = 1.33$$

$$\frac{N-8}{N+1} = 0.75\left[1-\left(\frac{2-1.33}{2+1}\right)^{0.566}\right]$$

$$N = 14.6 \cong 15$$

Height of column

$$H_c = (N-1)H_s + \Delta H + total\ plate\ thickness$$

Where

H_s is tray spacing

ΔH is 1 for bubble cap plates

Thickness is 5mm (0.005 m)

$$H_c = (15-1)0.5 + 1 + (0.005\ x\ 15)$$

$$H_c = 8.075\ m$$

Plate efficiency

$$E_o = 51 - 32.5\left[\log\left(\mu_{avg} \propto_{avg}\right)\right]$$

$$E_o = 51 - 32.5[\log(0.325\ x\ 3.95)]$$

$$E_o = 47.47\ \%$$

Actual number of plates

$$N_{actual} = \frac{N}{E_o}$$

$$N_{actual} = \frac{15}{0.4747}$$

$$N_{actual} = 31.5 \cong 32$$

Location of feed plate

$$\log\left(\frac{N_D}{N_B}\right) = 0.206 \log\left[\left(\frac{B}{D}\right)\left(\frac{X_{HK}}{X_{LK}}\right)\left(\frac{(X_{LK})_B}{(X_{HK})_D}\right)^2\right]$$

$$X_{LK} = \frac{F_B}{F_B + F_O}$$

$$X_{LK} = \frac{7.037}{7.037 + 0.4192}$$

$$X_{LK} = 0.94378$$

$$X_{HK} = \frac{F_O}{F_B + F_O}$$

$$X_{HK} = \frac{0.4192}{7.037 + 0.4192}$$

$$X_{HK} = 0.05622$$

$$\log\left(\frac{N_D}{N_B}\right) = 0.206 \log\left[\left(\frac{0.40206}{7.054}\right)\left(\frac{0.05622}{0.94378}\right)\left(\frac{0.01}{0.003}\right)^2\right]$$

$$\frac{N_D}{N_B} = 0.50908$$

$$N_D + N_B = 7.45606$$

Solving simultaneously gives

$$N_B = 4.94 \cong 5$$

Table 26: Distillation Column Specifications

Equipment	Distillation Column
Volume	76.57 m^3
Height	8.075 m
Diameter	0.766 m
Number of trays	32
Operating temperature	270 ˚C
Operating pressure	0.8 atm
Material of construction	Mild Steel

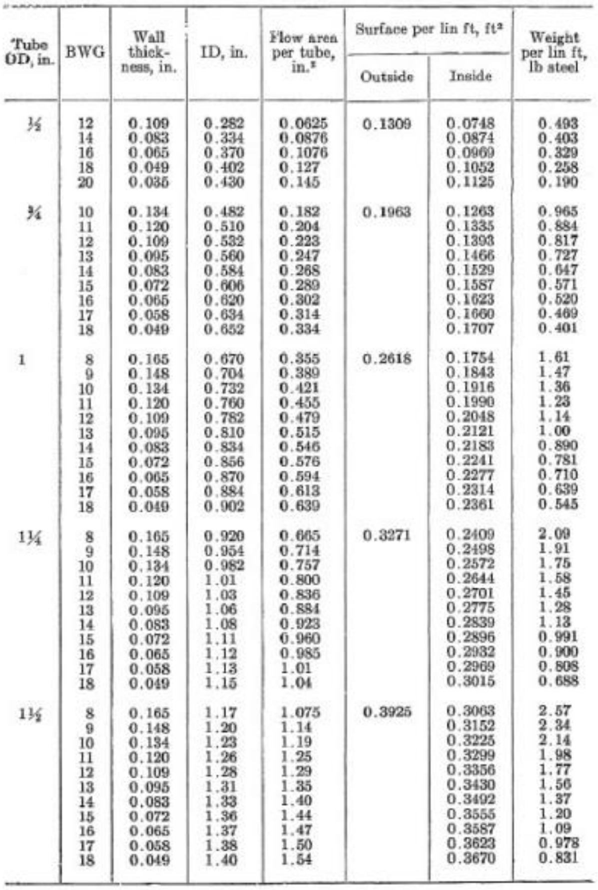
Hexane Column Pre Cooler

Table 27: Hexane Column Pre cooler

Hot fluid: Algal slurry	Cold fluid: Water
Shell side	Tube side
$T_{h\ in}$ = 90 °C	$T_{c\ in}$ = 25 °C
$T_{h\ out}$ = 50 °C	$T_{c\ out}$ = 40 °C
q = 4255729 kJ/h	c_{pc} = 4.18 kJ/kg°C
h_{do} = 0.0005 (sea water)	h_{di} = 0.0005 (water)
m_h = 26913 kg/h	

Assume outer tube diameter, BWG and length, find inner diameter from table 28 below (Kern)

Table 28: Tube Specifications



Tube OD, in.	BWG	Wall thickness, in.	ID, in.	Flow area per tube, in.²	Surface per lin ft, ft²		Weight per lin ft, lb steel
					Outside	Inside	
½	12	0.109	0.282	0.0625	0.1309	0.0748	0.493
	14	0.083	0.334	0.0876		0.0874	0.403
	16	0.065	0.370	0.1076		0.0969	0.329
	18	0.049	0.402	0.127		0.1052	0.258
	20	0.035	0.430	0.145		0.1125	0.190
¾	10	0.134	0.482	0.182	0.1963	0.1263	0.965
	11	0.120	0.510	0.204		0.1335	0.884
	12	0.109	0.532	0.223		0.1393	0.817
	13	0.095	0.560	0.247		0.1466	0.727
	14	0.083	0.584	0.268		0.1529	0.647
	15	0.072	0.606	0.289		0.1587	0.571
	16	0.065	0.620	0.302		0.1623	0.520
	17	0.058	0.634	0.314		0.1660	0.469
	18	0.049	0.652	0.334		0.1707	0.401
1	8	0.165	0.670	0.355	0.2618	0.1754	1.61
	9	0.148	0.704	0.389		0.1843	1.47
	10	0.134	0.732	0.421		0.1916	1.36
	11	0.120	0.760	0.455		0.1990	1.23
	12	0.109	0.782	0.479		0.2048	1.14
	13	0.095	0.810	0.515		0.2121	1.00
	14	0.083	0.834	0.546		0.2183	0.890
	15	0.072	0.856	0.576		0.2241	0.781
	16	0.065	0.870	0.594		0.2277	0.710
	17	0.058	0.884	0.613		0.2314	0.639
	18	0.049	0.902	0.639		0.2361	0.545
1¼	8	0.165	0.920	0.665	0.3271	0.2409	2.09
	9	0.148	0.954	0.714		0.2498	1.91
	10	0.134	0.982	0.757		0.2572	1.75
	11	0.120	1.01	0.800		0.2644	1.58
	12	0.109	1.03	0.836		0.2701	1.45
	13	0.095	1.06	0.884		0.2775	1.28
	14	0.083	1.08	0.923		0.2839	1.13
	15	0.072	1.11	0.960		0.2896	0.991
	16	0.065	1.12	0.985		0.2932	0.900
	17	0.058	1.13	1.01		0.2969	0.808
	18	0.049	1.15	1.04		0.3015	0.688
1½	8	0.165	1.17	1.075	0.3925	0.3063	2.57
	9	0.148	1.20	1.14		0.3152	2.34
	10	0.134	1.23	1.19		0.3225	2.14
	11	0.120	1.26	1.25		0.3299	1.98
	12	0.109	1.28	1.29		0.3356	1.77
	13	0.095	1.31	1.35		0.3430	1.56
	14	0.083	1.33	1.40		0.3492	1.37
	15	0.072	1.36	1.44		0.3555	1.20
	16	0.065	1.37	1.47		0.3587	1.09
	17	0.058	1.38	1.50		0.3623	0.978
	18	0.049	1.40	1.54		0.3670	0.831

d_o = ¾ inch = 0.01905 m

BWG = 16

d_i = 0.620 inch = 0.015748

L = 15 m

Calculate flow rate of steam m_h required

$$q = m_c c_{pc}(T_{c\ out} - T_{c\ in}) = m_h c_{ph}(T_{h\ out} - T_{h\ in})$$

$$4255729 = m_c\ x\ 4.18(40 - 25)$$

$$m_h = 67874\ kg/h$$

Calculate LMTD

$$LMTD = \frac{(T_{h\ in} - T_{c\ out}) - (T_{h\ out} - T_{c\ in})}{\ln \frac{(T_{h\ in} - T_{c\ out})}{(T_{h\ out} - T_{c\ in})}}$$

$$LMTD = \frac{(90 - 40) - (50 - 25)}{ln \frac{(90 - 40)}{(50 - 25)}}$$

$$LMTD = 36.07\ °C$$

Calculate R and S to find correction factor from figure (Kern)

$$R = \frac{T_1 - T_2}{t_1 - t_2}$$

$$R = \frac{90 - 50}{40 - 25}$$

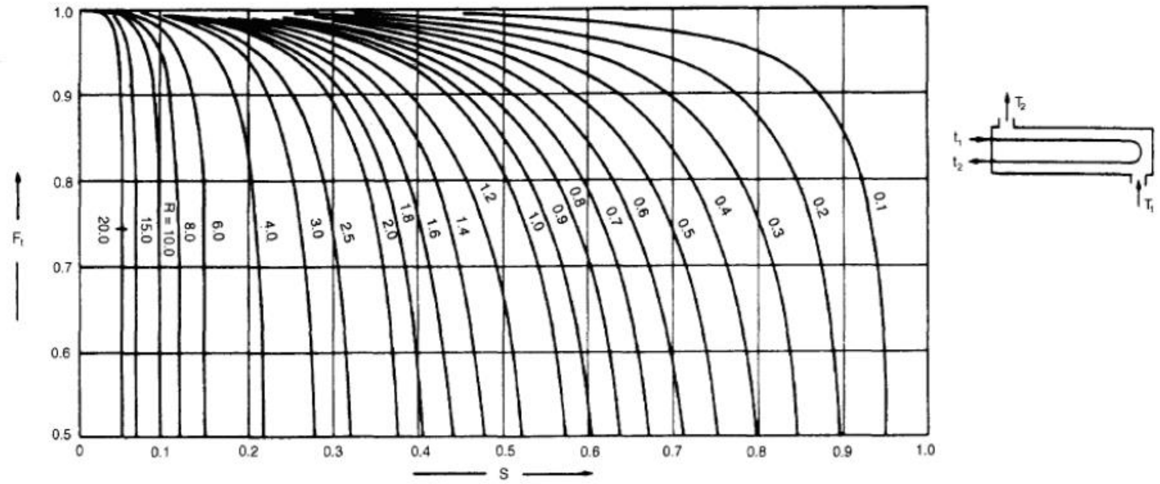
$$R = 2.67$$

$$S = \frac{t_2 - t_1}{T_1 - t_1}$$

$$S = \frac{40 - 25}{90 - 25}$$

$$S = 0.23$$

F_T from graph in figure 37 (Kern)



Temperature correction factor: one shell pass; two or more even tube 'passes

Figure 37: Temperature Correction Factor

$$F_T = 0.96$$

Corrected LMTD

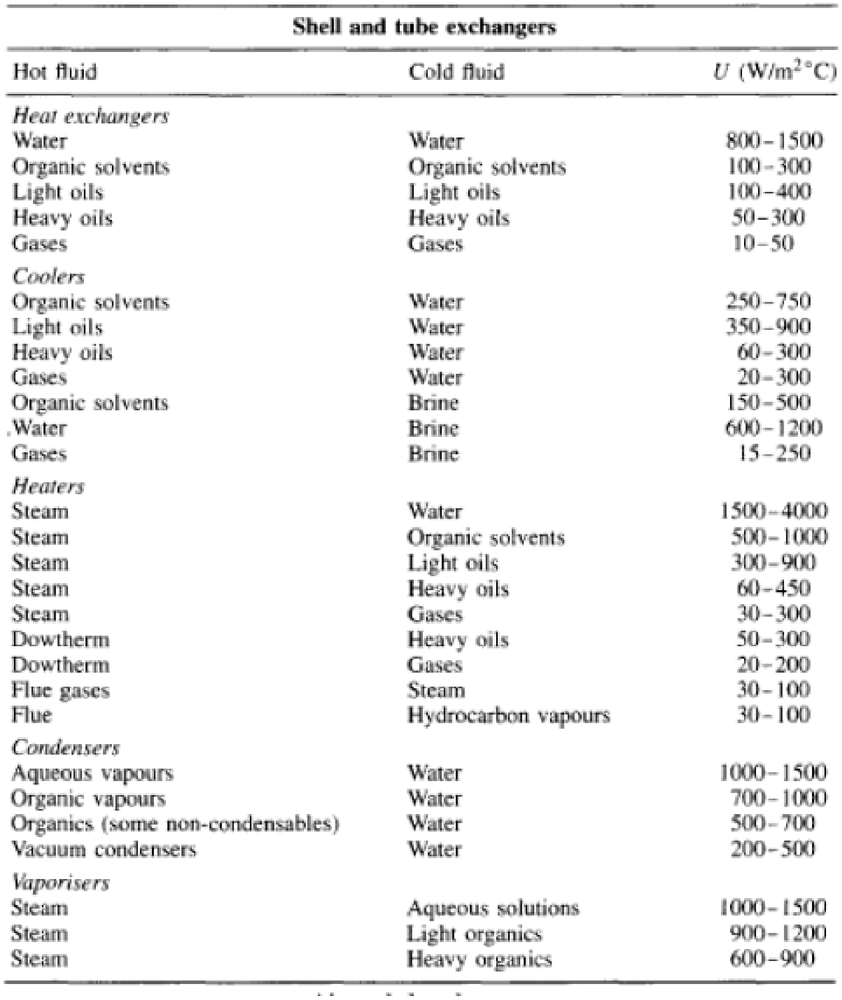
$$DT_m = F_T \ x \ LMTD$$

$$DT_m = 0.96 \ x \ 36.07$$

$$DT_m = 34.63 \ °C$$

Assume value of overall heat transfer coefficient from table 29 below

Table 29: Heat Transfer Coefficients



Shell and tube exchangers		
Hot fluid	Cold fluid	U (W/m^2°C)
Heat exchangers		
Water	Water	800–1500
Organic solvents	Organic solvents	100–300
Light oils	Light oils	100–400
Heavy oils	Heavy oils	50–300
Gases	Gases	10–50
Coolers		
Organic solvents	Water	250–750
Light oils	Water	350–900
Heavy oils	Water	60–300
Gases	Water	20–300
Organic solvents	Brine	150–500
Water	Brine	600–1200
Gases	Brine	15–250
Heaters		
Steam	Water	1500–4000
Steam	Organic solvents	500–1000
Steam	Light oils	300–900
Steam	Heavy oils	60–450
Steam	Gases	30–300
Dowtherm	Heavy oils	50–300
Dowtherm	Gases	20–200
Flue gases	Steam	30–100
Flue	Hydrocarbon vapours	30–100
Condensers		
Aqueous vapours	Water	1000–1500
Organic vapours	Water	700–1000
Organics (some non-condensables)	Water	500–700
Vacuum condensers	Water	200–500
Vaporisers		
Steam	Aqueous solutions	1000–1500
Steam	Light organics	900–1200
Steam	Heavy organics	600–900

Water (hot fluid) and water (cold fluid)

U = 800 W/m^2°C to 1500 W/m^2°C

U = 1150 W/m^2°C

Calculate area

$$A = \frac{q}{U\ x\ DT_m}$$

$$A = \frac{4255729}{1150\ x\ 34.63}$$

$$A = 106.87\ m^2$$

Calculate number of tubes

$$N_t = \frac{A}{\pi d_o L}$$

$$N_t = \frac{106.87}{\pi\ x\ 0.01905\ x\ 15}$$

$$N_t = 119\ \cong 120$$

Calculate tube pitch

$$p_t = 1.25 d_o$$

$$p_t = 1.25\ x\ 0.01905$$

$$p_t = 0.0238\ m$$

Select K1 and n1 values based on pitch arrangement and number of passes

Triangular arrangement, 2 passes of tubes

K_1 = 0.249

n_1 = 2.207

Calculate baffle spacing

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

$$D_b = 0.01905 \left(\frac{120}{0.249}\right)^{\frac{1}{2.207}}$$

$$D_b = 0.313\ m$$

Using figure 38 (Kern), find bundle diameter clearance

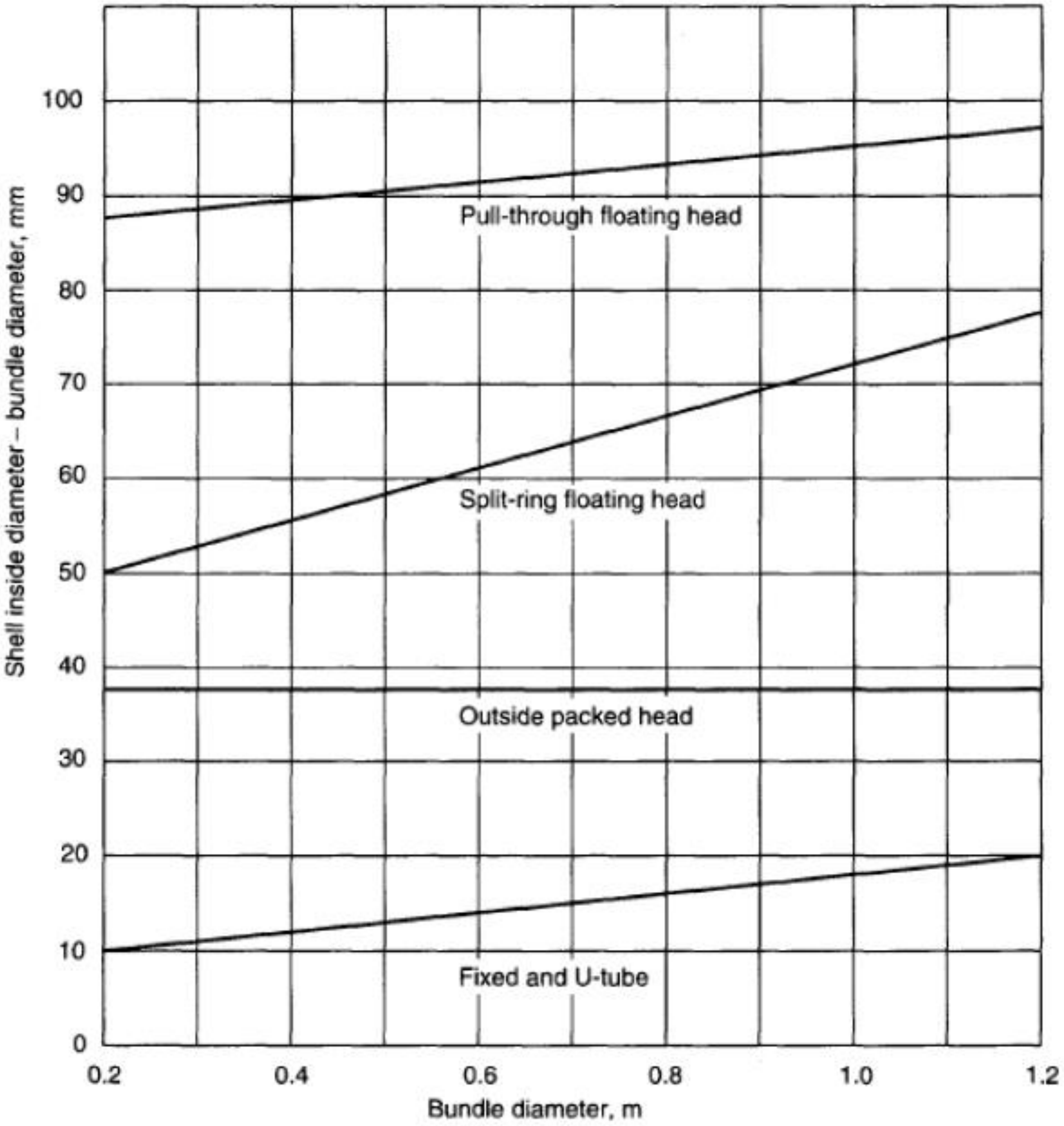


Figure 38: Bundle Diameter Clearance

BDC = 11 mm = 0.011 m

Calculate shell diameter

$$D_s = D_b + BDC$$

$$D_s = 0.313 + 0.011$$

$$D_s = 0.324\ m$$

Calculate baffle spacing

$$B_s = 0.4D_s$$

$$B_s = 0.4\ x\ 0.324$$

$$B_s = 0.1296\ m$$

Calculate area for cross flow

$$A_s = \frac{(p_t - \ d_o)D_s B_s}{p_t}$$

$$A_s = \frac{(0.0238\ - 0.01905)x\ 0.324\ x\ 0.1296}{0.0238}$$

$$A_s = \ 8.38\ x\ 10^{-3}\ m^2$$

Shell side mass velocity

$$G_s = \frac{shell\ side\ flow\ rate\ (m_h)}{A_s}$$

$$G_s = \frac{67874}{8.38\ x\ 10^{-3}}$$

$$G_s = 8099522\ kg/m^3h$$

Calculate number of tubes per pass

$$N_{tpp} = \frac{N_t}{number\ of\ passes}$$

$$N_{tpp} = \frac{120}{2}$$

$$N_{tpp} = 60$$

Calculate tube side mass velocity

$$G_m = \frac{tube\ side\ flow\ rate\ (m_c)}{N_{tpp}\pi \frac{{d_i}^2}{4}}$$

$$G_m = \frac{26193}{60\ x\ \pi\ x\ \frac{0.015478^2}{4}}$$

$$G_m = 2241265 \frac{kg}{m^2}/h$$

Table 30: Hexane Column Specifications

Total surface area	106.87 m^2
Number of tubes	120
Exchanger length	15 m
Exchanger diameter	0.324 m
Number of passes (s-t)	1-2

Hexane Recovery Preheater

Table 31: Hexane Recovery Preheater

Hot fluid: Steam	Cold fluid: Lipid and hexane
Shell side	Tube side
$T_{h\ in}$ = 145 °C	$T_{c\ in}$ = 50 °C
$T_{h\ out}$ = 100 °C	$T_{c\ out}$ = 85 °C
c_{ph} = 1.97 kJ/kg°C	q = 5076454 kJ/h
h_{do} = 0.0005 (water)	h_{di} = 0.0005 (oil)
	m_c = 26913 kg/h

Assume outer tube diameter, BWG and length, find inner diameter from table 28 (Kern)

d_o = ¾ inch = 0.01905 m

BWG = 16

d_i = 0.620 inch = 0.015748

L = 15 m

Assume thermal conductivity based on material of tube

k_W = 45 W/m°C (Steel at 100 °C)

Calculate flow rate of steam m_h required

$$q = m_c c_{pc}(T_{c\,out} - T_{c\,in}) = m_h c_{ph}(T_{h\,out} - T_{h\,in})$$

$$5076454 = m_h\ x\ 41.97(100 - 145)$$

$$m_h = 57264\ kg/h$$

Calculate LMTD

$$LMTD = \frac{(T_{h\,in} - T_{c\,out}) - (T_{h\,out} - T_{c\,in})}{\ln \frac{(T_{h\,in} - T_{c\,out})}{(T_{h\,out} - T_{c\,in})}}$$

$$LMTD = \frac{(145-85)-(100-50)}{ln\frac{(145-85)}{(100-50)}}$$

$$LMTD = 54.85\ °C$$

Calculate R and S to find correction factor

$$R = \frac{T_1 - T_2}{t_1 - t_2}$$

$$R = \frac{145-100}{85-50}$$

$$R = 1.29$$

$$S = \frac{t_2 - t_1}{T_1 - t_1}$$

$$S = \frac{85-50}{145-50}$$

$$S = 0.37$$

F_T from graph in figure 37 (Kern)

$$F_T = 0.93$$

Corrected LMTD

$$DT_m = F_T\ x\ LMTD$$

$$DT_m = 0.93\ x\ 54.85$$

$$DT_m = 51.01\ °C$$

Assume value of overall heat transfer coefficient from table 29

Water (hot fluid) and organic solvent (cold fluid)

U = 500 W/m^2°C to 1000 W/m^2°C

U = 750 W/m^2°C

Calculate area

$$A = \frac{q}{U\ x\ DT_m}$$

$$A = \frac{5076454}{750\ x\ 51.01}$$

$$A = 132\ m^2$$

Calculate number of tubes

$$N_t = \frac{A}{\pi d_o L}$$

$$N_t = \frac{132}{\pi\ x\ 0.01905\ x\ 15}$$

$$N_t = 148\ \cong 150$$

Calculate tube pitch

$$p_t = 1.25 d_o$$

$$p_t = 1.25\ x\ 0.01905$$

$$p_t = 0.0238\ m$$

Select K1 and n1 values based on pitch arrangement and number of passes

Triangular arrangement, 2 passes of tubes

K_1 = 0.249

n_1 = 2.207

Calculate baffle spacing

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

$$D_b = 0.01905 \left(\frac{150}{0.249}\right)^{\frac{1}{2.207}}$$

$$D_b = 0.346\, m$$

Using figure 38 (Kern), find bundle diameter clearance

BDC = 12 mm = 0.012 m

Calculate shell diameter

$$D_s = D_b + BDC$$

$$D_s = 0.346 + 0.012$$

$$D_s = 0.358\, m$$

Calculate baffle spacing

$$B_s = 0.4 D_s$$

$$B_s = 0.4\, x\, 0.358$$

$$B_s = 0.1432\, m$$

Calculate area for cross flow

$$A_s = \frac{(p_t - d_o) D_s B_s}{p_t}$$

$$A_s = \frac{(0.0238 - 0.01905) x\, 0.358\, x\, 0.1432}{0.0238}$$

$$A_s = 10.2\, x\, 10^{-3}\, m^2$$

Shell side mass velocity

$$G_s = \frac{shell\ side\ flow\ rate\ (m_h)}{A_s}$$

$$G_s = \frac{57264}{10.2\ x\ 10^{-3}}$$

$$G_s = 5614117.6\ kg/m^3h$$

Calculate number of tubes per pass

$$N_{tpp} = \frac{N_t}{number\ of\ passes}$$

$$N_{tpp} = \frac{150}{2}$$

$$N_{tpp} = 75$$

Calculate tube side mass velocity

$$G_m = \frac{tube\ side\ flow\ rate\ (m_c)}{N_{tpp}\pi\frac{{d_i}^2}{4}}$$

$$G_m = \frac{2750.}{75\ x\ \pi\ x\ \frac{0.015478^2}{4}}$$

$$G_m = 1882686.95\ \frac{kg}{m^2}/h$$

Table 32: Hexane Recovery Preheater Specifications

Total surface area	132 m^2
Number of tubes	150
Exchanger length	15 m
Exchanger diameter	0.358 m
Number of passes (t-s)	2-1

Reactor Preheater

Table 33: Reactor Preheater

Hot fluid: Steam	Cold fluid: Lipid and hexane
Tube side	Shell side
$T_{h\ in}$ = 370 °C	$T_{c\ in}$ = 110.2 °C
$T_{h\ out}$ = 310 °C	$T_{c\ out}$ = 300 °C
c_{ph} = 2.26 kJ/kg°C	q = 1781923 kJ/h
h_{do} = 0.0005 (water)	h_{di} = 0.0005 (oil)
	m_c = 2921 kg/h

Assume outer tube diameter, BWG and length, find inner diameter from table 28(Kern)

d_o = ¾ inch = 0.01905 m

BWG = 16

d_i = 0.620 inch = 0.015748

L = 15 m

Assume thermal conductivity based on material of tube

k_W = 45 W/m°C (Steel at 100 °C)

Calculate flow rate of steam m_h required

$$q = m_c c_{pc}(T_{c\,out} - T_{c\,in}) = m_h c_{ph}(T_{h\,out} - T_{h\,in})$$

$$1781923 = m_h\,x\,2.26(310 - 370)$$

$$m_h = 13141\ kg/h$$

Calculate LMTD

$$LMTD = \frac{(T_{h\,in} - T_{c\,out}) - (T_{h\,out} - T_{c\,in})}{\ln\frac{(T_{h\,in} - T_{c\,out})}{(T_{h\,out} - T_{c\,in})}}$$

$$LMTD = \frac{(370 - 300) - (310 - 110.2)}{ln\frac{(370 - 300)}{(310 - 110.2)}}$$

$$LMTD = 123.75\,°C$$

Calculate R and S to find correction factor

$$R = \frac{T_1 - T_2}{t_1 - t_2}$$

$$R = \frac{370 - 310}{300 - 110.2}$$

$$R = 0.316$$

$$S = \frac{t_2 - t_1}{T_1 - t_1}$$

$$S = \frac{300 - 110.2}{370 - 110.2}$$

$$S = 0.731$$

F_T from graph in figure 37 (Kern)

$$F_T = 0.88$$

Corrected LMTD

$$DT_m = F_T \ x \ LMTD$$

$$DT_m = 0.88 \ x \ 123.75$$

$$DT_m = 108.9 \ °C$$

Assume value of overall heat transfer coefficient from table 29

Water (hot fluid) and light oil (cold fluid)

U = 300 W/m^2°C to 900 W/m^2°C

U = 600 W/m^2°C

Calculate area

$$A = \frac{q}{U \ x \ DT_m}$$

$$A = \frac{1781923}{600 \ x \ 108.9}$$

$$A = 27.27 \ m^2$$

Calculate number of tubes

$$N_t = \frac{A}{\pi d_o L}$$

$$N_t = \frac{27.27}{\pi \; x \; 0.01905 \; x \; 15}$$

$$N_t = 30.37 \; \cong 40$$

Calculate tube pitch

$$p_t = 1.25 d_o$$

$$p_t = 1.25 \; x \; 0.01905$$

$$p_t = 0.0238 \; m$$

Select K1 and n1 values based on pitch arrangement and number of passes

Triangular arrangement, 2 passes of tubes

K_1 = 0.249

n_1 = 2.207

Calculate baffle spacing

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

$$D_b = 0.01905 \left(\frac{40}{0.249}\right)^{\frac{1}{2.207}}$$

$$D_b = 0.190 \; m$$

Using the chart in figure 38, find bundle diameter clearance

BDC = 10 mm = 0.010 m

Calculate shell diameter

$$D_s = D_b + BDC$$

$$D_s = 0.190 + 0.010$$

$$D_s = 0.200\ m$$

Calculate baffle spacing

$$B_s = 0.4D_s$$

$$B_s = 0.4\ x\ 0.200$$

$$B_s = 0.08\ m$$

Calculate area for cross flow

$$A_s = \frac{(p_t - d_o)D_s B_s}{p_t}$$

$$A_s = \frac{(0.0238\ - 0.01905)x\ 0.200\ x\ 0.08}{0.0238}$$

$$A_s = 3.19\ x\ 10^{-3}\ m^2$$

Shell side mass velocity

$$G_s = \frac{shell\ side\ flow\ rate\ (m_h)}{A_s}$$

$$G_s = \frac{2921}{3.19\ x\ 10^{-3}}$$

$$G_s = 914734\ kg/m^3h$$

Calculate number of tubes per pass

$$N_{tpp} = \frac{N_t}{number\ of\ passes}$$

$$N_{tpp} = \frac{40}{2}$$

$$N_{tpp} = 20$$

Calculate tube side mass velocity

$$G_m = \frac{tube\ side\ flow\ rate\ (m_c)}{N_{tpp}\pi \frac{{d_i}^2}{4}}$$

$$G_m = \frac{13141}{20\ x\ \pi\ x\ \frac{0.015478^2}{4}}$$

$$G_m = 3492037\ \frac{kg}{m^2}/h$$

Table 34: Reactor Preheater Specifications

Total surface area	27.27 m^2
Number of tubes	40
Exchanger length	15 m
Exchanger diameter	0.200 m
Number of passes (t-s)	2-1

Flash Tank 1

Table 35: Flash Tank 1

Hot fluid: Biodiesel	Cold fluid: Water
Tube side	Shell side
$T_{h\ in}$ = 228 °C	$T_{c\ in}$ = 25 °C
$T_{h\ out}$ = 175 °C	$T_{c\ out}$ = 40 °C
q = 311615 kJ/h	c_{pc} = 4.18 kJ/kg°C
h_{do} = 0.0005 (oil)	h_{di} = 0.0005 (water)
m_h = 2916 kg/h	

Assume outer tube diameter, BWG and length, find inner diameter from table 28 (Kern)

d_o = ¾ inch = 0.01905 m

BWG = 16

d_i = 0.620 inch = 0.015748

L = 15 m

Assume thermal conductivity based on material of tube

k_W = 45 W/m°C (Steel at 100 °C)

Calculate flow rate of steam m_h required

$$q = m_c c_{pc}(T_{c\ out} - T_{c\ in}) = m_h c_{ph}(T_{h\ out} - T_{h\ in})$$

$$311615 = m_c\ x\ 4.18(40 - 25)$$

$$m_h = 14969\ kg/h$$

Calculate LMTD

$$LMTD = \frac{(T_{h\ in} - T_{c\ out}) - (T_{h\ out} - T_{c\ in})}{\ln\frac{(T_{h\ in} - T_{c\ out})}{(T_{h\ out} - T_{c\ in})}}$$

$$LMTD = \frac{(228\ - 40) - (175 - 25)}{ln\frac{(228 - 40)}{(175 -\ 25)}}$$

$$LMTD = 168.29\ °C$$

Calculate R and S to find correction factor

$$R = \frac{T_1 -\ T_2}{t_1 -\ t_2}$$

$$R = \frac{25 - 40}{175 - 228}$$

$$R = 0.283$$

$$S = \frac{t_2 -\ t_1}{T_1 -\ t_1}$$

$$S = \frac{175 - 228}{25 - 228}$$

$$S = 0.261$$

F_T from graph in figure 37 (Kern)

$$F_T =\ 0.85$$

Corrected LMTD

$$DT_m =\ F_T\ x\ LMTD$$

$$DT_m = 0.85\ x\ 168.29$$

$$DT_m = 143.05\ °C$$

Assume value of overall heat transfer coefficient from table 29

Water (hot fluid) and light oil (cold fluid)

$U = 350\ W/m^2°C$ to $900\ W/m^2°C$

$U = 625\ W/m^2°C$

Calculate area

$$A = \frac{q}{U\ x\ DT_m}$$

$$A = \frac{311615}{625\ x\ 143.05}$$

$$A = 3.48\, m^2$$

Calculate number of tubes

$$N_t = \frac{A}{\pi d_o L}$$

$$N_t = \frac{3.48}{\pi\ x\ 0.01905\ x\ 15}$$

$$N_t = 3.87\ \cong 4$$

Calculate tube pitch

$$p_t = 1.25 d_o$$

$$p_t = 1.25\ x\ 0.01905$$

$$p_t = 0.0238\, m$$

Select K1 and n1 values based on pitch arrangement and number of passes

Triangular arrangement, 2 passes of tubes

$K_1 = 0.249$

$n_1 = 2.207$

Calculate baffle spacing

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

$$D_b = 0.01905 \left(\frac{4}{0.249}\right)^{\frac{1}{2.207}}$$

$$D_b = 0.067\ m$$

Using the chart in figure 38, find bundle diameter clearance

BDC = 10 mm = 0.010 m

Calculate shell diameter

$$D_s = D_b + BDC$$

$$D_s = 0.067 + 0.010$$

$$D_s = 0.077\ m$$

Calculate baffle spacing

$$B_s = 0.4D_s$$

$$B_s = 0.4\ x\ 0.077$$

$$B_s = 0.0308\ m$$

Calculate area for cross flow

$$A_s = \frac{(p_t - d_o)D_sB_s}{p_t}$$

$$A_s = \frac{(0.0238\ - 0.01905)x\ 0.077\ x\ 0.0308}{0.0238}$$

$$A_s = 0.474\ x\ 10^{-3}\ m^2$$

Shell side mass velocity

$$G_s = \frac{shell\ side\ flow\ rate\ (m_h)}{A_s}$$

$$G_s = \frac{14969}{0.474\ x\ 10^{-3}}$$

$$G_s = 31560760\ kg/m^3h$$

Calculate number of tubes per pass

$$N_{tpp} = \frac{N_t}{number\ of\ passes}$$

$$N_{tpp} = \frac{4}{2}$$

$$N_{tpp} = 2$$

Calculate tube side mass velocity

$$G_m = \frac{tube\ side\ flow\ rate\ (m_c)}{N_{tpp}\pi\frac{{d_i}^2}{4}}$$

$$G_m = \frac{2916}{4\ x\ \pi\ x\ \frac{0.015478^2}{4}}$$

$$G_m = 7485431\ \frac{kg}{m^2}/h$$

Table 36: Flash Tank 1 Specifications

Total surface area	3.48 m^2
Number of tubes	4
Exchanger length	15 m
Exchanger diameter	0.077 m
Number of passes (t-s)	2-1

Flash Tank 2

Table 37: Flash Tank 2

Hot fluid: Methanol and water	Cold fluid: Water
Shell side	Tube side
$T_{h\ in}$ = 175 °C	$T_{c\ in}$ = 25 °C
$T_{h\ out}$ = 90 °C	$T_{c\ out}$ = 40 °C
q = 71834 kJ/h	c_{pc} = 4.18 kJ/kg°C
h_{do} = 0.0005 (sea water)	h_{di} = 0.0005 (water)
m_h = 337 kg/h	

Assume outer tube diameter, BWG and length, find inner diameter from table 28 (Kern)

d_o = ¾ inch = 0.01905 m

BWG = 16

d_i = 0.620 inch = 0.015748

L = 15 m

Assume thermal conductivity based on material of tube

k_W = 45 W/m°C (Steel at 100 °C)

Calculate flow rate of steam m_h required

$$q = m_c c_{pc}(T_{c\ out} - T_{c\ in}) = m_h c_{ph}(T_{h\ out} - T_{h\ in})$$

$$71834 = m_c\ x\ 4.18(40 - 25)$$

$$m_h = 1145\ kg/h$$

Calculate LMTD

$$LMTD = \frac{(T_{h\ in} - T_{c\ out}) - (T_{h\ out} - T_{c\ in})}{\ln\frac{(T_{h\ in} - T_{c\ out})}{(T_{h\ out} - T_{c\ in})}}$$

$$LMTD = \frac{(175-40)-(90-25)}{ln\frac{(175-40)}{(90-\ 25)}}$$

$$LMTD = 95.77\ °C$$

Calculate R and S to find correction factor

$$R = \frac{T_1 - T_2}{t_1 - t_2}$$

$$R = \frac{25\ -\ 40}{90-175}$$

$$R = 0.176$$

$$S = \frac{t_2 - t_1}{T_1 - t_1}$$

$$S = \frac{90-175}{25-175}$$

$$S = 0.567$$

F_T from graph in figure 37 (Kern)

$$F_T = \ 0.99$$

Corrected LMTD

$$DT_m = \ F_T\ x\ LMTD$$

$$DT_m = 0.99\ x\ 95.77$$

$$DT_m = 94.81\ °C$$

Assume value of overall heat transfer coefficient from table 29

Water (hot fluid) and organic solvent (cold fluid)

U = 250 $W/m^2°C$ to 750 $W/m^2°C$

U = 500 $W/m^2°C$

Calculate area

$$A = \frac{q}{U \ x \ DT_m}$$

$$A = \frac{71834}{500 \ x \ 94.81}$$

$$A = 1.515 \ m^2$$

Calculate number of tubes

$$N_t = \frac{A}{\pi d_o L}$$

$$N_t = \frac{1.515}{\pi \ x \ 0.01905 \ x \ 15}$$

$$N_t = 1.68 \ \cong 2$$

Calculate tube pitch

$$p_t = 1.25 d_o$$

$$p_t = 1.25 \ x \ 0.01905$$

$$p_t = 0.0238 \ m$$

Select K1 and n1 values based on pitch arrangement and number of passes

Triangular arrangement, 2 passes of tubes

K_1 = 0.249

n_1 = 2.207

Calculate baffle spacing

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

$$D_b = 0.01905 \left(\frac{2}{0.249}\right)^{\frac{1}{2.207}}$$

$$D_b = 0.0489 \, m$$

Using the chart in figure 38, find bundle diameter clearance

BDC = 10 mm = 0.010 m

Calculate shell diameter

$$D_s = D_b + BDC$$

$$D_s = 0.0489 + 0.010$$

$$D_s = 0.0589 \, m$$

Calculate baffle spacing

$$B_s = 0.4 D_s$$

$$B_s = 0.4 \, x \, 0.0589$$

$$B_s = 0.02356 \, m$$

Calculate area for cross flow

$$A_s = \frac{(p_t - d_o) D_s B_s}{p_t}$$

$$A_s = \frac{(0.0238 - 0.01905) x \, 0.0589 \, x \, 0.02356}{0.0238}$$

$$A_s = 0.277 \, x \, 10^{-3} \, m^2$$

Shell side mass velocity

$$G_s = \frac{shell\ side\ flow\ rate\ (m_h)}{A_s}$$

$$G_s = \frac{1145}{0.277\ x\ 10^{-3}}$$

$$G_s = 4125579\ kg/m^3 h$$

Calculate number of tubes per pass

$$N_{tpp} = \frac{N_t}{number\ of\ passes}$$

$$N_{tpp} = \frac{2}{2}$$

$$N_{tpp} = 1$$

Calculate tube side mass velocity

$$G_m = \frac{tube\ side\ flow\ rate\ (m_c)}{N_{tpp} \pi \frac{{d_i}^2}{4}}$$

$$G_m = \frac{337}{60\ x\ \pi\ x\ \frac{0.015478^2}{4}}$$

$$G_m = 1730171 \frac{kg}{m^2}/h$$

Table 38: Flash Tank 2 Specifications

Total surface area	1.515 m^2
Number of tubes	2
Exchanger length	15 m
Exchanger diameter	0.0589 m
Number of passes (t-s)	2-1

Distillation Column Preheater

Table 39: Distillation Column Preheater

Hot fluid: Steam	Cold fluid: Biodiesel
Tube side	Shell side
$T_{h\ in}$ = 370 °C	$T_{c\ in}$ = 175 °C
$T_{h\ out}$ = 310 °C	$T_{c\ out}$ = 250 °C
c_{ph} = 2.26 kJ/kg°C	q = 765239 kJ/h
h_{do} = 0.0005 (water)	h_{di} = 0.0005 (oil)
	m_c = 2442 kg/h

Assume outer tube diameter, BWG and length, find inner diameter from table 28 (Kern)

d_o = ¾ inch = 0.01905 m

BWG = 16

d_i = 0.620 inch = 0.015748

L = 15 m

Assume thermal conductivity based on material of tube

k_W = 45 W/m°C (Steel at 100 °C)

Calculate flow rate of steam m_h required

$$q = m_c c_{pc}(T_{c\ out} - T_{c\ in}) = m_h c_{ph}(T_{h\ out} - T_{h\ in})$$

$$765239 = m_h\ x\ 2.26(370 - 310)$$

$$m_h = 5643\ kg/h$$

Calculate LMTD

$$LMTD = \frac{(T_{h\ in} - T_{c\ out}) - (T_{h\ out} - T_{c\ in})}{\ln\frac{(T_{h\ in} - T_{c\ out})}{(T_{h\ out} - T_{c\ in})}}$$

$$LMTD = \frac{(370 - 250) - (310 - 175)}{ln\frac{(370 - 250)}{(310 - 175)}}$$

$$LMTD = 127.35\,°C$$

Calculate R and S to find correction factor

$$R = \frac{T_1 - T_2}{t_1 - t_2}$$

$$R = \frac{175 - 250}{310 - 370}$$

$$R = 1.25$$

$$S = \frac{t_2 - t_1}{T_1 - t_1}$$

$$S = \frac{310 - 370}{175 - 370}$$

$$S = 0.308$$

F_T from graph in figure 37 (Kern)

$$F_T = 0.95$$

Corrected LMTD

$$DT_m = F_T\ x\ LMTD$$

$$DT_m = 0.95\ x\ 127.35$$

$$DT_m = 120.98\,°C$$

Assume value of overall heat transfer coefficient from table 29

Water (hot fluid) and light oil (cold fluid)

U = 300 W/m^2°C to 900 W/m^2°C

U = 600 W/m^2°C

Calculate area

$$A = \frac{q}{U\ x\ DT_m}$$

$$A = \frac{765239}{600\ x\ 120.98}$$

$$A = 10.54\ m^2$$

Calculate number of tubes

$$N_t = \frac{A}{\pi d_o L}$$

$$N_t = \frac{10.54}{\pi\ x\ 0.01905\ x\ 15}$$

$$N_t = 11.75\ \cong 12$$

Calculate tube pitch

$$p_t = 1.25 d_o$$

$$p_t = 1.25\ x\ 0.01905$$

$$p_t = 0.0238\ m$$

Select K1 and n1 values based on pitch arrangement and number of passes

Triangular arrangement, 2 passes of tubes

K_1 = 0.249

n_1 = 2.207

Calculate baffle spacing

$$D_b = d_o \left(\frac{N_t}{K_1}\right)^{\frac{1}{n_1}}$$

$$D_b = 0.01905 \left(\frac{12}{0.249}\right)^{\frac{1}{2.207}}$$

$$D_b = 0.110\ m$$

Using the chart in figure 38, find bundle diameter clearance

BDC = 10 mm = 0.010 m

Calculate shell diameter

$$D_s = D_b + BDC$$

$$D_s = 0.110 + 0.010$$

$$D_s = 0.120\ m$$

Calculate baffle spacing

$$B_s = 0.4D_s$$

$$B_s = 0.4\ x\ 0.120$$

$$B_s = 0.048\ m$$

Calculate area for cross flow

$$A_s = \frac{(p_t - d_o)D_s B_s}{p_t}$$

$$A_s = \frac{(0.0238\ - 0.01905) x\ 0.120\ x\ 0.048}{0.0238}$$

$$A_s = 1.152\ x\ 10^{-3}\ m^2$$

Shell side mass velocity

$$G_s = \frac{shell\ side\ flow\ rate\ (m_h)}{A_s}$$

$$G_s = \frac{2442}{1.152\ x\ 10^{-3}}$$

$$G_s = 2119791.6\ kg/m^3h$$

Calculate number of tubes per pass

$$N_{tpp} = \frac{N_t}{number\ of\ passes}$$

$$N_{tpp} = \frac{12}{2}$$

$$N_{tpp} = 6$$

Calculate tube side mass velocity

$$G_m = \frac{tube\ side\ flow\ rate\ (m_c)}{N_{tpp}\pi \frac{{d_i}^2}{4}}$$

$$G_m = \frac{5643}{6\ x\ \pi\ x\ \frac{0.015478^2}{4}}$$

$$G_m = 4828565\ \frac{kg}{m^2}/h$$

Table 40: Distillation Column Preheater Specifications

Total surface area	10.54 m^2
Number of tubes	12
Exchanger length	15 m
Exchanger diameter	0.120 m
Number of passes (t-s)	2-1

Costing

Purchase Cost of Equipment

Nutrient Tank

Material: mild steel

Size = S = 150 m^3

n = 0.55

$$Ce = CS^n$$

$$Ce = 435000 \ x \ 150^{0.55}$$

$$Ce = 6{,}844{,}400 \ PKR$$

Bioreactor

Glass

Material: ½ inch industrial grade glass

Surface area = SA = 551.6 m^2 = 5937.4 ft^2

Rate = R = 170 PKR/ft^2

$$Ce_g = SA \ x \ R$$

$$Ce_g = 5937.4 \ x \ 170$$

$$Ce_g = \ 1{,}009{,}358 \ PKR$$

Stainless Steel

Material: ½ inch stainless steel

Surface area = SA = 91.95 m^2

Volume = V =1.16775 m^3

Density of steel = 7480 kg/m^3

Mass = M = 8734.8 kg

Rate = R = 500 PKR/kg

$$Ce_{ss} = M\ x\ R$$

$$Ce_{ss} = 8734.8\ x\ 500$$

$$Ce_{ss} = 4{,}367{,}393\ PKR$$

Sparger

Number = N =140

Rate = R = 1500 PKR/piece

$$Ce_{s} = N\ x\ R$$

$$Ce_{s} = 140\ x\ 15000$$

$$Ce_{s} = 2{,}100{,}000\ PKR$$

Total cost of bioreactors

$$Ce = Ce_{g} + Ce_{ss} + Ce_{s}$$

$$Ce = 1009358 + 4367393 + 210000$$

$$Ce = 5{,}586{,}751\ PKR$$

Lipid Extraction Column

Column

Material: mild steel

Size = S = 76.57 m^3

n = 0.6

$$Ce_c = CS^n$$

$$Ce_c = 240000\ x\ 76.57^{0.6}$$

$$Ce_c = 3{,}240{,}700\ PKR$$

Packing

Type: Saddlox

Volume = V = 57.43 m^3

Rate = R = 96000 PKR/m^3

$$Ce_p = V\ x\ R$$

$$Ce_p = 57.43\ x\ 96000$$

$$Ce_p = 5{,}513{,}040\ PKR$$

Total cost of column

$$Ce = \ Ce_c + \ Ce_p$$

$$Ce = 3240700 + 5513040$$

$$Ce = 8{,}753{,}740\ PKR$$

Hexane Recovery Flash Tank

Material: mild steel

Size = S = 5.137 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000\ x\ 5.137^{0.6}$$

$$Ce = 640{,}600\ PKR$$

Hexane Storage Tank

Material: mild steel

Size = S = 48 m^3

n = 0.55

$$Ce = CS^n$$

$$Ce = 435000\ x\ 48^{0.55}$$

$$Ce = 3{,}657{,}300\ PKR$$

Methanol Storage Tank

Material: mild steel

Size = S = 1 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000\ x\ 1^{0.6}$$

$$Ce = 240{,}000\ PKR$$

Oil Tank

Material: mild steel

Size = S = 3.12 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000\ x\ 3.12^{0.6}$$

$$Ce = 475{,}100\ PKR$$

Mcgyan Reactor

Reactor

Material: mild steel

Size = S = 0.0545 m^3

n = 0.6

$$Ce_r = CS^n$$

$$Ce_r = 240000\ x\ 0.0545^{0.6}$$

$$Ce_r = 41{,}800\ PKR$$

Catalyst

Type: zirconium oxide

Mass = M = 9.80 kg

Rate = R = 228 PKR/kg

$$Ce_c = M\ x\ R$$

$$Ce_c = 9.80\ x\ 228$$

$$Ce_c = 223{,}400\ PKR$$

Total cost of reactor

$$Ce = \ Ce_r + \ Ce_c$$

$$Ce = 41800 + 223400$$

$$Ce = 26{,}520{,}000$$

Flash Tank 1

Material: mild steel

Size = S = 0.572 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000\ x\ 0.572^{0.6}$$

$$Ce = 199{,}200\ PKR$$

Flash Tank 2

Material: mild steel

Size = S = 0.066 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000\ x\ 0.066^{0.6}$$

$$Ce = 47{,}000\ PKR$$

Decanter

Material: mild steel

Size = S = 0.225 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000 \ x \ 0.225^{0.6}$$

$$Ce = 98{,}100 \ PKR$$

Distillation Column

Column

Material: mild steel

Size = S = 3.722 m^3

n = 0.6

$$Ce_c = CS^n$$

$$Ce_c = 240000 \ x \ 3.722^{0.6}$$

$$Ce_c = 528{,}000 \ PKR$$

Plates

Type: bubble cap plates

Number = N = 32

Rate = R = 27000 PKR/ plate

$$Ce_p = N \ x \ R$$

$$Ce_p = 32 \ x \ 27000$$

$$Ce_p = 864{,}000 \ PKR$$

Total cost of Distillation

$$Ce = \ Ce_c + \ Ce_p$$

$$Ce = 528000 + 864000$$

$$Ce = 1{,}392{,}000$$

Methane Storage Tank

Material: mild steel

Size = S = 1937 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000\ x\ 1937^{0.6}$$

$$Ce = 22{,}500{,}000\ PKR$$

Glycerol Storage Tank

Material: mild steel

Size = S = 107 m^3

n = 0.6

$$Ce = CS^n$$

$$Ce = 240000\ x\ 107^{0.6}$$

$$Ce = 150{,}000\ PKR$$

Heat Exchangers

from table

$$Ce_1 = 2{,}200{,}000\ PKR$$

$$Ce_2 = 4{,}000{,}000\ PKR$$

$$Ce_3 = 17{,}000{,}000\ PKR$$

$$Ce_4 = 1{,}700{,}000\ PKR$$

$$Ce_5 = 500{,}000\ PKR$$

$$Ce_6 = 200{,}000\ PKR$$

$$Ce_7 = 700{,}000\ PKR$$

Total Cost of Heat Exchangers

$$Ce = Ce_1 + Ce_2 + Ce_3 + Ce_4 + Ce_5 + Ce_6 + Ce_7$$

$$Ce = 2200000 + 4000000 + 17000000 + 1700000 + 500000 + 200000 + 700000$$

$$Ce = 26{,}300{,}000\ PKR$$

Compressor

Size = S = 50 kW

n = 0.8

$$Ce = CS^n$$

$$Ce = 1920\ x\ 50^{0.8}$$

$$Ce = 4{,}390{,}000\ PKR$$

Total Purchase Cost of Equipment (PCE)

Table 41: Purchase Cost of Equipment

Equipment	PCE (PKR)
Nutrient Tank	6,844,400
Bioreactor	5,586,700
Lipid Extraction Column	8,753,740
Hexane Recovery Flash Tank	640,600
Hexane Storage Tank	3,657,300
Methanol Storage Tank	2,400,000
Oil Storage Tank	475,100
Mcgyan Reactor	26,520,000
Flash Tank 1	199,200
Flash Tank 2	47,000
Decanter	98,100
Distillation Column	1,392,000
Methane Tank	40,810,000
Glycerol Tank	150,000
Heat exchangers	26,300,000
Compressor	4,390,000
Total	128,264,140

Physical Plant Cost (PPC)

Table 42: Physical Plant Cost Lang Factor

Factor	Value
f_1 – Equipment erection	0.45
f_2 – Piping	0.45
f_3 – Instrumentation	0.15
f_4 – Electrical	0.10
f_5 – Buildings	0.10
f_6 – Utilities	0.45
f_7 – Storages	0.20
f_8 – Site development	0.05
f_9 – Ancillary buildings	0.20
Total	3.15

$$PPC = PCE\ x\ f_{1-9}$$

$$PPC = 128264140\ x\ 3.15$$

$$PPC = 404{,}032{,}041\ PKR$$

Fixed Capital Cost (FCC)

Table 43: Fixed Capital Cost Lang Factor

Factor	Value
f_{10} – Design and engineering	0.25
f_{11} – Contractor’s fee	0.05
f_{12} – Contingency	0.10
Total	1.40

$$FC = PPC\ x\ f_{10-12}$$

$$FC = 404032041\ x\ 1.40$$

$$FC = 565{,}644{,}857\ PKR$$

Variable Cost (VC) (per annum)

Table 44: Variable Cost

Item	Cost (PKR)
Methanol (100 PKR/kg)	219,000,000
Hexane (100 PKR/kg)	68,837,800
Nutrients (40 PKR/kg)	183,539,520
Carbon dioxide ()	15,457,000
Steam (1.20 PKR/kg)	594,147,007
Electricity (13.50 PKR/(kW/h)	138,302,304
Total (VC)	1,219,283,631

Revenue (per annum)

Table 45: Revenue

Item	Cost (PKR)
Biodiesel (60 PKR/kg)	1,094,824,800
Methane (Gasification) (50 PKR/kg)	559,983,000
Glycerol (5 PKR/kg)	4,730,400
Total (R)	1,659,538,200

Profit (per annum)

$$profit = revenue - variable\ cost$$

$$profit = 1659538200 - 1219283631$$

$$profit = 440{,}254{,}569\ PKR$$

Fixed Cost (per annum)

Table 46: Fixed Cost

Item	Cost (PKR)
Maintenances	22,421,000
Labour	15,000,000
Laboratory costs	-
Supervision	4,484,200
Plant overheads	11,210,500
Insurance and loyalties	13,452,600
Total (FC)	66,568,300

Annual Production Cost

Includes sales expense, general overheads and research and development

$$annual\ production\ cost = 1.2(variable\ cost + fixed\ cost)$$

$$annual\ production\ cost = 1.2(1219283631 + 66568300)$$

$$annual\ production\ cost = \ 1{,}543{,}022{,}317\ PKR$$

Production Cost

$$production\ cost = \frac{annual\ production\ cost}{annual\ production\ rate}$$

$$production\ cost = \frac{1{,}543{,}022{,}317.2}{18247080}$$

$$production\ cost = 84.56\ PKR/kg$$

Rate of Return

Investment = 565,644,857 PKR

Production Cost = 1,543,022,317 PKR

Profit = 440,254,569 PKR

Revenue = 1,659,538,200 PKR

$$rate\ of\ return = \frac{profit}{investment + production\ cost}\ x\ 100$$

$$rate\ of\ return = \frac{440254569}{565644857 + 1543022317}\ x\ 100$$

$$rate\ of\ return = 20.88\%$$

Payback period

$$payback\ period = \frac{100}{rate\ of\ return}$$

$$payback\ period = \frac{100}{20.88}$$

$$payback\ period = 4.8\ years$$

Conclusion

Our project has successfully addressed the two major problems hindering the algal biodiesel industry. The process we have designed provides an energy and cost efficient way of continuous harvesting of algae in a closed environment and continuous conversion of oil into biodiesel at a high conversion rate. The reactor we have designed processes the oil into biodiesel in less than a minute compared to other batch processes that takes about 2 hours. Hence the process that we have proposed can produce 100 times more biodiesel. Dewatering of algae is a very energy consuming process which accounts for the major running cost. The technique we have used to dewater algae is using a membrane that does not require much energy; moreover our biodiesel reactor is not affected by the presence of water.

All in we can conclude our proposed project can be used to model a sustainable process for biodiesel production from algae in future. Pakistan is a fuel deficient country; production of biodiesel will reduce the foreign trade bill and put less strain on the economy while providing employment to many professionals.

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