# Queensland Sustainable Aviation Fuels Initiative

Costs and Assumptions

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# 1 Introduction

## 1.1 Background

The Queensland Sustainable Aviation Fuels Initiative is focused on evaluating biofuels production, with a particular focus on environmental impact, technical feasibility, and economic viability. This study considers the production of sustainable aviation fuel from three different feed stocks (sugarcane, algae, and the oils seeds of the *Pongamia* tree) through both technoeconomic and lifecycle analyses.[1]

## 1.2 Technoeconomic Analysis

These technoeconomic models have developed been based on open and accountable information from journal reports, patents, and industry to answer important questions about the manufacturing process for sustainable aviation fuels from the three feed stocks. A production scale of 15 million gallons per year (15,000,000 Gallons/year) has been adopted to reflect a mature, full-scale biofuels production facility.

[1] Klein-Marcuschamer, Daniel, Turner, Christopher, Allen, Mark, Gray, Peter, Dietzgen, Ralf G., Gresshoff, Peter M., Hankamer, Ben, Heimann, Kirsten, Scott, Paul T., Stephens, Evan, Speight, Robert and Nielsen, Lars K. (2013) Technoeconomic analysis of renewable aviation fuel from microalgae, Pongamia pinnata, and sugarcane. Biofuels, Bioproducts and Biorefining, 7 4: 416-428. doi:10.1002/bbb.1404

# 2 Costs

## 2.1 Materials and Labour

### 2.1.1 Labour Costs

Labour costs were calculated from the mean weekly earning statistics for factory process workers,[2] using the Wessel method (Equation 9-7) for estimating process labour requirements.[3]

| **Labour Costs:** | | |
|---|---|---|
| Operator | 23.42 \$/hr | [2] [3] |
| Supervisor | 46.84 \$/hr | [2] [3] |

Table 1: Labour Costs

### 2.1.2 Material & Utility Costs

Feedstock, nutrient and utilities were costed using a combination of industry quotes and published cost history.

| **Feedstock Costs:** | | |
|---|---|---|
| Pongamia Seeds | 590 \$/MT | [4] |
| A-Molasses | 190 \$/M T | [5] |
| **Utility Costs:** | | |
| Electricity | 10 c/kWh | |
| Cooling Water | 0.05 \$/MT | |
| Process Steam | 12 \$/MT | [6] |
| **Feed Nutrients:** | | |
| Ammonium Hydroxide | 229.41 \$/MT | [7] |
| Diammonium Phosphate | 703 \$/MT | [7] |
| Nitrate | 500 \$/MT | |
| **Processing Chemicals:** | | |
| Hydrogen | 1.014 \$/kg | [8] |
| Industrial Hexane | 2 \$/kg | |

Table 2: Material & Utility Costs

[2]Employee Earnings, Benefits and Trade Union Membership. Canberra (ACT): Australian Bureau of Statistics 6310.0: 2011

[3]Perry RH and Green DW, Perry's Chemical Engineers' Handbook. McGraw-Hill New York, 2008

[4]Gresshoff PM. unpublished results. 2012

[5]Lavarack BP. personal communication. 2012

[6]Patel M, Crank M, Dornburg V, Hermann B, Roes L, Husing B, et al., Medium and Long-term Opportunities and Risks of the Biotechnological Production of Bulk Chemicals from Renewable Resources (BREW). Utrecht (NL): Utrecht University: (2006).

[7] CRU News, Fertilizer Week Article.

[8]Molburg JC and Doctor RD. Hydrogen from Steam-Methane Reforming with CO2 Capture. 20th Annual International Pittsuburgh Coal Conference; Pittsburgh (PA) 2003.

## 2.2 Capital Expenses

Facility-dependent costs were derived from relevant studies and corrected using the Chemical Engineering Plant Cost Index (CEPCI), as well as from Aspen Process Economic Analyzer. Highly-specific unit operations (e.g. Evodos centrifuges, cracking mills, hexane extractors) were priced from vendor quotes.

### 2.2.1 Unit Operation Costing

- Evodos Centrifuge [9]
- Pumps, Hydrogen Stripper, Distillation Column [10]
- Compressors, Fired Heaters, Hydroisomerization/Hydrocracker, HDO Reactors [11]
- Vessels, Heat Exchangers/Coolers/Heaters, Anaerobic Digestor, Methane Boiler, Turbogenerator [12]
- Amine Scrubber [13]
- Sonicator [14]
- Algae Pond [15]
- Fermentors [12]
- Cracking Mill, Flaking Mill, Hexane Extractor [16]

## 2.3 Financial Assumptions

Financial assumptions were made with the expectation of low technological risk (i.e. based on an Nth plant), similar to previous studies.[12] [17]

The plants were financed with a 60-40 debt-equity split, assuming an interest rate of 8% for the debt, and a 10% discount rate for NPV analysis. Minimum selling price (MSP) analysis was performed assuming a project lifetime of 25 years for all cases.

[9] Quotation: Evodos BV

[10] Aspentech. Aspen Process Economic Analyzer. 7.0 ed. Burlington (MA) 2008.

[11] Jones SB, Holladay JE, Valkenburg C, Stevens DJ, Walton CW, Kinchin C, et al., Production of Gasoline and Diesel from Biomass via Fast Pyrolysis, Hydrotreating and Hydrocracking: A Design Case. Richland (WA): Pacific Northwest National Laboratory PNNL-18284: (2009).

[12] Aden A, Ruth M, Ibsen K, Jechura J, Neeves K, Sheehan J, et al., Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis for Corn Stover. Golden (CO): National Renewable Energy Laboratory NREL/TP-510-32438: (2002)

[13] White, C.W. ASPEN Plus Simulation of CO2 Recovery Process. (DOE/NETL-2002/1182: 2002)

[14] Quotation: Hielscher Ultrasonics GmbH

[15] Benemann J and Oswald WJ, Systems and economic analysis of microalgae ponds for conversion of CO2 to biomass: United States Department of Energy. Pittburgh Energy Technology Center (1996)

[16] Quotation: Crown Iron Works

[17] Klein-Marcuschamer D, et al. (2010) Technoeconomic analysis of biofuels: A wiki-based platform for lignocellulosic biorefineries. Biomass Bioenerg 34:1914-21

# 3 Sugarcane

Production of bio-jet fuel from sugarcane has been divided into two models, the first modelling a industry standard sugarcane mill which extracts and purifies the sucrose from the harvested sugarcane. Only a single crystallization pass is modelled with the molasses directed to the fermentation model. The fermentation model is based upon Amyris technology for farnesene production in yeast. The product farnesene is recovered as an organic phase prior to refining. UOP$^{TM}$ refining processes have been modelled, which hydrogenates and isomerizes farnesene to produce a hydrocarbon product with a rich aviation fraction. A atmospheric distillation column is used to recover the aviation fuel, along with side-products including naptha and diesel.

**Design Basis:** 16,000,000 gallons per year of Jet A-1 blend or drop in substitute.

## 3.1 Sugarcane Mill Unit Operations

### 3.1.1 Cane Preparation

**3.1.1.1 Shredding Mill** The sugarcane stalks are processed to allow efficient recovery of the sugars. A shredding mill is modelled, reducing the size of the pieces of cane to a size suitable for extraction.

Need for review: low
Need for innovation: low

**3.1.1.2 Cane Mill** The shredded cane is further processed in a 5 mill tandem, mass balance based upon a Bundaberg high extraction mill.[18] A counter-current flow of imbibition water is added to carry the extracted sugars.

Need for review: low
Need for innovation: low

### 3.1.2 Juice Clarification

**3.1.2.1 Raw Juice Heater** The raw juice is heated in a plate heater (higher heat transfer coefficient) to 75°C.[18]

Need for review: low
Need for innovation: low

**3.1.2.2 Lime Mixing Tank** Intermediate Liming, using a lime-saccharate mixture of evaporator syrup combined with A 15° Baum (13.26 wt% CaO) solution. This is added at a rate of 0.9 kg CaO / tonne of sugarcane.[18]

Need for review: medium
Need for innovation: low

**3.1.2.3 Juice Flash Vessel** The limed juice is heated in a plate heater and flashed to 100°C to removed entrained or dissolved air.

Need for review: low
Need for innovation: low

**3.1.2.4 Juice Clarifier** The raw juice is purified, decolorized and clarified in a SRI Rapid clarifier. Phosphorous Pentoxide (300 mg/kg cane juice)[18] is added as a flocculant.

Need for review: low
Need for innovation: low

### 3.1.3 Juice Concentration

**3.1.3.1 Juice Evaporator** The sucrose is concentrated in a 5 effect evaporator, operating at 10 kPa.[18] The resulting syrup has a concentration of 67 Brix[18] (67 wt%) and is fed to the crystallization pans.

Need for review: low
Need for innovation: low

[18] Rein, P. Cane Sugar Engineering. (Verlag Dr. Albert Bartens KG: Berlin, 2007).

#### 3.1.3.2 Vapour Condenser
The water vapour from the juice evaporator is condensed using a contacting condenser with 20 water/vapour flowrate ratio of 20.[18]
Need for review: low
Need for innovation: low

#### 3.1.3.3 Rotary Vacuum Filter
The rotary vacuum filter recovers water from the clarifier muds. The muds are filtered and washed to recover fine sugarcane material called bagacillo which is recycled to the x while the water is recycled to the x. A typical wash water rate of 200g/100g of cake is modelled with a cake LOD of 70% and solids retention of 85%.[18]
Need for review: low
Need for innovation: low

### 3.1.4 Pan House

#### 3.1.4.1 Syrup Heater
The evaporated syrup is heated to 65°C[18] using low pressure steam (144 kPa[18]) before being feed to the crystallization pans.
Need for review: low
Need for innovation: low

#### 3.1.4.2 A Pans
The sucrose is crystallized from the concentrated syrup using a continuous, unstirred, crystallization pan. The mass balance is based upon a single crystallization stage with the crystallized sugar being sold as product while the molasses is fed onto the fermentation process. The crystallization mass balance is based upon the A Pans from a conventional three-boiling scheme.[18]
Need for review: medium
Need for innovation: low

#### 3.1.4.3 A Centrifuge
High grade continuous centrifugal. Wash with condensate.
Need for review:
Need for innovation:

#### 3.1.4.4 Molasses Storage Tank
Molasses needs to be stored to account between year-round fermentation and fuel production, and seasonal processing of cane.
Need for review: low
Need for innovation: low

### 3.1.5 Cogeneration

#### 3.1.5.1 Bagasse Boiler
Waste bagasse from the 5 mill train is combusted in a fluidized bed boiler[18] to generate high pressure (80 bar) steam for use in covering process electricity and steam requirements.
Need for review: low
Need for innovation: low

#### 3.1.5.2 Turbogenerator
A multi-stage turbogenerator is used to generate electricity and generate the processes requirement for process steam.
Need for review: low
Need for innovation: low

## 3.2 Fermentation Mill Unit Operations

### 3.2.1 Fermentation

**3.2.1.1 Growth Fermentor Chain** The growth chain uses a series of fermentors to produce the yeast concentration in the main fermentor. Growth performance was based on reported yields by Amyris [19]. The nutrient A farnesene productivity of 16.9 g/L/d, biomass yield of 40g/100g of sugar, and 25% molar excess of nutrients is modelled.

Need for review: medium
Need for innovation: medium

**3.2.1.2 Main Fermentor** Assumed 25% molar excess of nutrients Farnesene Titer 104.3 g/L, Farnesene Productivity 16.9 g/L/d, Operating Temperature 35 °C, assumed negligible biomass production.[19]

Need for review: medium
Need for innovation: medium

### 3.2.2 Fermentation Separation

**3.2.2.1 Yeast Centrifuge** The whole broth product from the fermentation will have a combination of aqueous and organic phases with an emulsion formed due to the presence of extracellular materials. The cellular biomass (and some aqueous phase) is removed first in a continuous disk stack centrifuge. Farnesene losses in this centrifuge are reported at 10 wt%.[20]

Need for review: medium
Need for innovation: medium

**3.2.2.2 Blending Tank** The clarified broth (without yeast cells) is treated to breakdown the emulsion. Effective conditions have been reported as solution pH of 9.5, salt concentration 1.2M, and 0.5 % by volume Tergitol (a non-ionic surfactant with emulsion stabilization applications).[20] After 1 hour emulsion breakdown was reported at 90-95%.

Need for review: medium
Need for innovation: medium

**3.2.2.3 Organic Centrifuge** The remaining clarified broth is treated in a continuous disk stack centrifuge to recover the organic phase, separating the aqueous phases which contains the trace solids and debris. Reported farnesene recovery can be as high as 97 wt%.[20]

Need for review: medium
Need for innovation: medium

### 3.2.3 Hydrocracking

**3.2.3.1 Cracking Fired Heater** The cracking reactor feed is heated to the reaction temperature of 332°C by combusting the propane rich waste gases in a fired heater.[21]

Need for review: low
Need for innovation: low

**3.2.3.2 Hydroisomerization/Hydrocracking Reactor** The straight chain alkanes produced in the hydrodeoxygenation reactor are unsuitable for use as aviation fuel and require further processes. These alkanes need to be cracked and isomerized to generate a suitable product for use as an aviation fuel substitute.[21] A cracking reactor operating at 332 °C, 5171 kPag, 0.5 hour$^{-1}$ LHSV, with a specialized UOP catalyst is able to convert around 50 wt% of the feed material into a aviation fuel fraction.[21] The remainder going to either diesel, naptha, or light gas fractions.

Need for review: medium
Need for innovation: medium

[19] Pray, T. (2010) Amyris Biomass R&D, Technical Advisory Committee: Drop-in Fuels Panel.

[20] P. Tabur and G. Dorin, "Purification Methods for Bio-Organic Compounds," WO Patent 115074, October 7, 2010.

[21] McCall, M.J., Kocal, J.A., Bhattacharyya, A. Kalnes, T.N. & Brandvold, T.A. 'Production of Aviation Fuel from Renewable Feedstocks' US Patent 2009/0283442 A1. November 19, 2009

### 3.2.4 Product Recovery

**3.2.4.1 Hot Flash Vessel** High pressure flash vessel to remove light gases prior to product distillation. The temperature and pressure of the flash vessel have been modelled to minimise losses of the key aviation fraction.

Need for review: medium

Need for innovation: low

**3.2.4.2 Product Distillation** The separation and recovery of the naptha, aviation, and diesel fractions has been modelled in an atmospheric crude distillation unit.[22]

Need for review: medium

Need for innovation: low

[22] Parkash, S. Refining Processes Handbook: Gulf Professional Publishing (2003)

# 4 Algae

Algae is grown using flue gas from a coal fired power station as a carbon enriched feed. *Nannochloropsis*, a microalgae with reported high growth and oil content has been used as the basis of the composition and growth data. The harvested algae is recovered using centrifuges. The intracellular oil is extracted through a combination of sonication to disrupt the cells and a solvent extraction process to recover an oil enriched stream. The algae muds are digested to produce biogas which is combusted for electricity and steam recycle. The triacylglyceride oil is degummed (treated to remove phospholipid impurities) prior to being refined into a combination of fuel products. UOP$^{TM}$ refining processes have been modelled, producing hydrocarbon product with a rich aviation fraction. A atmospheric distillation column is used to recover the aviation fuel, along with side-products including naptha and diesel.

**Design Basis:** 16,000,000 gallons per year of Jet A-1 blend or drop in substitute.

## 4.1 Algae Ponds

Algae growth in open raceway ponds.

### 4.1.1 Seawater Feed Pump

The seawater feed pump transfers seawater to the algae ponds. The seawater flow is controlled to maintain the algae ponds salinity. Seawater is assumed to be readily available to the algae ponds site.

Need for review: high
Need for innovation: low

### 4.1.2 Groundwater Feed Pump

The groundwater feed pump provides water to the algae ponds to offset the moisture losses through evaporation. The water could be sourced through the water grid, a water bore, or a combination of both.

Need for review: medium
Need for innovation: low

### 4.1.3 Algae Open Ponds

Open raceway ponds are growing Nannochloropsis microalgae. Growth performance is modelled from reported year-averaged data. An average production rate of 20.0 g/m$^2$/day[23] has been modelled for an algae culture concentration of 500 mg/L.[24] This growth performance was achieved in 0.12 m deep ponds at a salinity of 35 g/L.[24] These growth conditions give a residence time of 3 days. Algae pond mixing is achieved through paddle wheels.

| |
|---|
| Phosphorus |
| Nitrogen |
| Diammonium Phosphate |
| Nitrogen |
| Nitrate Salt |
| Sulphur |
| Sulphate (groundwater) |
| Carbon |
| Flue Gas |

Table 3: Nutrients

It was assumed 10% excess of nutrients (apart from Carbon Dioxide) is needed.

[23] Ben-Amotz, A. Biofuel and $CO_2$ Capture by Algae. Second Algae Biomass Summit (2008) https://newbusiness.grc.nasa.gov/wp-content/uploads/2008/12/ben-amotz-nasa-nov-2008.pdf

[24] Boussiba, S., Vonshak, A., Cohen, Z., Avissar, Y. & Richmond, A. Lipid and biomass production by the halotolerant microalga Nannochloropsis salina. Biomass 12, 37-47 (1987).

| | | |
|---|---|---|
| Ash | 7 wt% | [25] |
| Biomass | 73 wt% | [24] |
| Lipid | 20 wt% | [24] |

Table 4: Algae Composition

A detailed breakdown of lipid composition[26] was simplified to three lipid fractions to model the different lipid and phospholipids.

| | | |
|---|---|---|
| Neutral Lipid | (TAG) | Triacylglyceride C16:O |
| Phospholipid | (PC) | Phosphatidylcholine C16:1(n-7) |
| Non-hydratable Phospholipid | (PG) | Phosphatidylglycerol C16:0 tripalmitin |

Table 5: Algae Lipid Composition

Need for review: medium
Need for innovation: medium

#### 4.1.4 Flue Gas Cooler

The flue gas needs to be cooled before bubbling through the algae ponds. The flue gas needs to be cooled to 32°C, the maximum temperature for Nannochloropsis growth.[24]
Need for review: medium
Need for innovation: low

### 4.2 Algae Separation

#### 4.2.1 Algae Concentration

The concentration of the harvested algae stream is achieved through a Phytolutions preconcentration unit specifically engineered to operate with the Evodos centrifuge.
Need for review: High
Need for innovation: High

#### 4.2.2 Algae Centrifugation

The concentrated algae stream is centrifuged to further separate algae and water. The modelled centrifuge is based upon the Evodos Centrifuge 2/250. Final algae concentration is a 30 wt% algae paste.[27]
Need for review: High
Need for innovation: High

### 4.3 Algae Hexane Extraction

The algae cells are disrupted using sonication and the oil components extracted using hexane as solvent.

[25] Negoro, M. et al. Carbon dioxide fixation by microalgae photosynthesis using actual flue gas discharged from a boiler. Appl Biochem Biotechnol 39-40, 643-653 (1993).

[26] Schneider, J. & Roessler, P. Radiolabeling studies of lipids and fatty acids in Nannochloropsis (Eustigmatophyceae), An oleaginous marine alga. Journal of Phycology 30, 594-598 (1994).

[27] Collet P et al, Life Cycle Assessment of Microalgae Culture Coupled to Biogas Production. Bioresource Technology 102, 207-214, 2011

### 4.3.1 Hexane Addition

Algal oil extraction has been modelled using a wet biomass extraction technique.[28] Suitable moisture content and solvent loading data [29] has been adapted for extraction with hexane as an industrial solvent. Sonication moisture content has been modelled at 82 wt%[29], while the extraction moisture content is increased to 92.4 wt%.[29]

Need for review: medium

Need for review: medium

### 4.3.2 Sonicator

Sonication has been modelled based upon specific performance data for the Hielsher sonicator. Optimal cell lysis and oil recovery was reported after 60 min sonication at 45°C. [30]

Need for review: medium

Need for innovation: medium

### 4.3.3 Hexane Extraction

Improved oil recovery requires additional mixing of the solvent with the algae. 93.8% recovery of algal oil has been reported using a 6 hour residence time.[28]

Need for review: medium

Need for innovation: low

### 4.3.4 Biomass Centrifuge

A decanting centrifuge is used to separate the aqueous phase and cellular debris from the solvent. The formation of an emulsion is likely and so the centrifuge performance has been modelled similarly to a degumming centrifuge.[31]

Need for review: medium

Need for innovation: low

### 4.3.5 Hexane Evaporator

A three effect evaporator has been modelled for the evaporation and recovery of the hexane solvent for recycle. The evaporation is modelled using performance data for an evaporator in a seed oil extraction plant. [31]

Need for review: low

Need for innovation: low

### 4.3.6 Hexane Storage Tank

Make-up hexane flow to cover 1.0 wt% losses to sludge and 120 ppm in the raw oil.[32]

Need for review: low

Need for innovation: low

## 4.4 Algae Degumming

### 4.4.1 Feed Heater

Prior to the addition of citric acid addition the raw oil is heated to 70°C.[32]

Need for review: low

Need for innovation: low

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[28] Converti, A., Casazza, A.A., Ortiz, E.Y., Perego, P. & Del Borghi, M. Effect of temperature and nitrogen concentration on the growth and lipid content of Nannochloropsis oculata and Chlorella vulgaris for biodiesel production. Chemical Engineering and Processing: Process Intensification 48, 1146-1151 (2009).

[29] Belarbi, E.H., Molina, E. & Chisti, Y. A process for high yield and scaleable recovery of high purity eicosapentaenoic acid esters from bicroalgae and fish oil. Enzyme and Microbial Technology 26, 516-529 (1999).

[30] Cravotto, G. et. al. Improved extraction of vegetable oils under high-intensity ultrasound and/or microwaves. Ultrasonics Sonochemistry 15, 898-902 (2008).

[31] Sheehan, J., Camobreco, V., Duffield, J., Graboski, M. & Shapouri, H. Life Cycle Inventory of Biodiesel and Petroleum Diesel for Use in an Urban Bus: Final Report. 314 (National Renewable Energy Laboratory: 1998).

[32] H. Ringers and J. Segers, "Degumming Process for Triglyceride Oils," U.S. Patent 4049686, September 20, 1977.

### 4.4.2 Citric Acid Addition

A 50 wt% citric acid solution is added to the oil which reacts with the non-hydratable phospholipids (NHP), transforming them into a hydratable form which can be removed in the centrifuge. A citric acid feed rate of 0.3wt% on raw oil is sufficient to reduce oil phosphorus content to 22 ppm.[32]

Need for review: medium

Need for innovation: low

### 4.4.3 Oil Cooler

The oil-acid mixture needs to be cooled to 32°C before the water wash.[32]

Need for review: low

Need for innovation: low

### 4.4.4 Water Wash

The cooled oil is mixed with distilled water to hydrate all the phospholipids into a semi-crystalline phase ready for centrifugation. Wash water is added at a rate of 1.0 wt% of the feed oil rate.[32]

Need for review: low

Need for innovation: low

### 4.4.5 Separation Heater

Prior to centrifugation the oil/aqueous mixture is heated to 85°C to improve the viscosity for centrifuging.[32]

Need for review: low

Need for innovation: low

### 4.4.6 Oil Centrifuge

The oil and aqueous phases are separated in a disk-stack centrifuge. Oil losses in the centrifuge are reported at 3.11 wt%.[31]

Need for review: medium

Need for innovation: low

## 4.5 Algae UOP Refining

### 4.5.1 Hydrodeoxygenation Heater

The hydrodeoxygenation reactor oil feed is heated to the reaction temperature of 350°C by combusting the propane rich waste gases in a fired heater.[33]

Need for review: low

Need for innovation: low

### 4.5.2 Hydrodeoxygenation Reactor

The hydrodeoxygenation reactor is down flow packed reactor divided into three stages, with an equicurent feed of oil and hydrogen. In each stage the triacylglyceride oil is reduced to straight alkanes, corresponding to the lipids, and propane, from the glycerol backbone. The UOP catalyst reduces the oil through both the deoxygenation reaction (60 wt%) and the hydrogenation reaction (40 wt%).[34] Stage 1 & 2 oil conversion is 90 wt%, while the final finishing stage operates at the higher 98 wt% conversion.[33]

**Deoxygenation Reaction:** 1 Glycerol Trioleate ($C_{57}H_{104}O_6$) + 6 Hydrogen ($H_2$) -> 3 Heptadecane ($C_{17}H_{36}$) + 1 Propane ($C_3H_8$) + 3 Carbon Dioxide ($CO_2$)

**Hydrogenation Reaction:** 1 Glycerol Trioleate ($C_{57}H_{104}O_6$) + 15 Hydrogen ($H_2$) -> 3 Octadecane ($C_{18}H_{38}$) + 1 Propane ($C_3H_8$) + 6 Water ($H_2O$)

Need for review: low

[33] Kokayeff, P., Marker, T.L. & Petri, J.A. "Production of Fuel from Renewable Feedstocks using a Finishing Reactor" US Patent 2010/0133144 A1. June 3, 2010

[34] Perego, C., Sabatino, L., Baldiraghi, F. & Faraci, G. "Process for Producing Hydrocarbon Fractions from Mixtures of a Biological Origin" US Patent 2009/0300970 A1. December 10, 2009

Need for innovation: low

### 4.5.3 Intermediate Gas Flash

Before the final finishing stage of the hydrodeoxygenation reactor all the waste gases ($CO_2$, $H_2$, $H_2O$, and short hydrocarbons - particularly propane) are flashed off. Model conditions of 100 °C and 35 bar were optimised for hydrocarbon recovery.

Need for review: medium
Need for innovation: low

### 4.5.4 Hot Hydrogen Stripper

A countercurrent stripper removes the waste light gases by using the dry hydrogen feed. This is particularly to reduce the water concentration (catalyst poison for the hydrocracking catalyst) to 100 pm.[35]

Need for review: low
Need for innovation: low

## 4.6 Algae Hydrocracking

### 4.6.1 Cracking Fired Heater

The cracking reactor feed is heated to the reaction temperature of 332°C by combusting the propane rich waste gases in a fired heater.[21]

Need for review: low
Need for innovation: low

### 4.6.2 Hydroisomerization/Hydrocracking Reactor

The straight chain alkanes produced in the hydrodeoxygenation reactor are unsuitable for use as aviation fuel and require further processes. These alkanes need to be cracked and isomerized to generate a suitable product for use as an aviation fuel substitute.[21] A cracking reactor operating at 332 °C, 5171 kPag, 0.5 hour$^{-1}$ LHSV, with a specialized UOP catalyst is able to convert around 50 wt% of the feed material into a aviation fuel fraction.[21] The remainder going to either diesel, naptha, or light gas fractions.

Need for review: medium
Need for innovation: medium

## 4.7 Algae Product Recovery

### 4.7.1 Hot Flash Vessel

High pressure flash vessel to remove light gases prior to product distillation. The temperature and pressure of the flash vessel have been modelled to minimise losses of the key aviation fraction.

Need for review: medium
Need for innovation: low

### 4.7.2 Product Distillation

The separation and recovery of the naptha, aviation, and diesel fractions has been modelled in an atmospheric crude distillation unit.[22]

Need for review: medium
Need for innovation: low

[35] Perego, C., Sabatino, L., Baldiraghi, F. & Faraci, G. "Process for Producing Hydrocarbon Fractions from Mixtures of a Biological Origin" US Patent 2009/0300970 A1. December 10, 2009

## 4.8 Algae Amine Scrubber

### 4.8.1 Amine Scrubber

The waste gas from the hydrodeoxygenation is treated to remove impurities, particularly carbon dioxide and water. An amine scrubber, a counter current gas-liquid contactor, with a 10 wt% MDEA (Methyl diethanol amine) absorbent. At the high pressure conditions used in this process a carbon dioxide removal efficiency of 71 wt%[13] has been reported. The hydrogen rich scrubbed gas is recycled as hydrogen feed to the hydrodexygenation reactor.
Need for review: low
Need for innovation: low

### 4.8.2 Degasifier

The rich MDEA solution from the amine scrubber is treated in a degasifier to remove the absorbed carbon dioxide.
Need for review: medium
Need for innovation: low

### 4.8.3 Carbon Dioxide Flash

The degasified carbon dioxide is flashed to recover trace water.
Need for review: low
Need for innovation: low

## 4.9 Algae Anaerobic Digestion

### 4.9.1 Anaerobic Digester

The biomass sludge from the biomass centrifuge is digested anaerobically. A 31 hour residence time digestor has been reported to give a 24% biomass reduction of algae sludge.[36] 90% of the digester muds is recycled to algae ponds to allow undigested nutrients to be recycled.
Need for review: medium
Need for innovation: medium

## 4.10 Algae Cogeneration

### 4.10.1 Methane Boiler

The digester biogas is combusted in a methane boiler to generate high pressure (80 bar) steam for use in covering process electricity and steam requirements.
Need for review: low
Need for innovation: low

### 4.10.2 Turbogenerator

A multi-stage turbogenerator is used to generate electricity and generate the processes requirement for process steam.
Need for review: low
Need for innovation: low

[36] Samson, R. & Leduy, A. Biogas production from anaerobic digestion of Spirulina maxima algal biomass. Biotechnology and Bioengineering 24, 1919-1924 (1982)

# 5 Pongamia

*Pongamia pinnata* is a legume tree which produces a seed rich in oil. These seeds are harvested, dried, cracked, dehulled, and flaked in preparation for extraction of the oil. Industrial hexane is used to extract the seed oil in a continuous belt extractor. The remainder of the seed biomass is digested to produce biogas which is combusted for electricity and steam recycle. The triacylglyceride oil is degummed (treated to remove phospholipid impurities) prior to being refined into a combination of fuel products. UOP$^{TM}$ refining processes have been modelled, producing hydrocarbon product with a rich aviation fraction. A atmospheric distillation column is used to recover the aviation fuel, along with side-products including naptha and diesel.

**Design Basis:** 16,000,000 gallons per year of Jet A-1 blend or drop in substitute.

## 5.1 Seed Processing

### 5.1.1 Prestorage Dryer

Prior to storage the raw kernels are dried to a moisture content of 14.0 wt%[37] in a fluidized bed dryer. Drying conditions and energy consumption were modelled based upon soybean drying performance.[38]

Need for review: medium

Need for innovation: low

### 5.1.2 Kernel Silo

The pongamia kernels are stored to compensate for the seasonal production of seeds. 240 days worth of material are stored, allowing for a 90 day seed harvesting season.[38]

Need for review: low

Need for innovation: low

### 5.1.3 Kernel Dryer

The moisture content of the kernels needs to be reduced to 10.0 wt% to allow efficient cracking and separation of the hulls[37]. Drying conditions and energy consumption were modelled based upon soybean drying performance.[38]

Need for review: medium

Need for innovation: low

### 5.1.4 Cracking Mill

The dried kernels are broken into particles one-fourth to one-sixth the size of the kernel.[37]

Need for review: medium

Need for innovation: low

### 5.1.5 Aspirator

An aspirator removes the hulls from the cracked kernels, removing the hulls for anaerobic digestion with the extracted meal.

Need for review: low

Need for innovation: low

### 5.1.6 Flaking Mill

The kernels are rolled into flakes 0.25-0.37mm prior to hexane extraction.[37]

Need for review: medium

Need for innovation: low

[37] Moore, N.H. Oilseed handling and preparation prior to solvent extraction. J Am Oil Chem Soc 60, 189-192 (1983).

[38] Soponronnarit, S., Swasdisevi, T., Wetchacama, S. & Wutiwiwatchai, W. Fluidised bed drying of soybeans. Journal of Stored Products Research 37, 133-151 (2001).

## 5.2 Pongamia Hexane Extraction

### 5.2.1 Hexane Extraction

The oil is extracted from the flaked pongamia seeds using industrial hexane in a counter-current belt extractor, assumed to operate with similar conditions and performance as when processing soybeans. Oil extraction performance, extractor mass balance, and energy requirements where based on reported soybean data[31]. An extraction residence time of 0.57 hours[39] and operating temperature of 60 °C[40] are modelled.

Need for review: medium
Need for innovation: low

### 5.2.2 Desolventiser

The desolventiser (meal desolventiser-toaster) using hexane vapour to recover hexane solvent retained in the meal for recycling.[40]

Need for review: medium
Need for innovation: low

### 5.2.3 Meal Cyclone

The desolventised pongamia flakes are cycloned to separate the hexane vapour from the flakes.

Need for review: medium
Need for innovation: low

### 5.2.4 Hexane Evaporator

A three effect evaporator has been modelled for the evaporation and recovery of the hexane solvent for recycle. The evaporation is modelled using performance available for soybean oil. [31]

Need for review: low
Need for innovation: low

### 5.2.5 Hexane Storage Tank

Make-up hexane flow of 0.0024 kg/kg flaked seed to cover losses in the hexane extraction to the pongamia meal.[31]

Need for review: low
Need for innovation: low

## 5.3 Pongamia Degumming

### 5.3.1 Feed Heater

Prior to the addition of citric acid addition the raw oil is heated to 70°C.[32]

Need for review: low
Need for innovation: low

### 5.3.2 Citric Acid Addition

A 50 wt% citric acid solution is added to the oil which reacts with the non-hydratable phospholipids (NHP), transforming them into a hydratable form which can be removed in the centrifuge. A citric acid feed rate of 0.3wt% on raw oil is sufficient to reduce oil phosphorus content to 22 ppm.[32]

Need for review: medium
Need for innovation: low

[39] Faner, S.A., Durand, G.A., Crapiste, G.H. & Bandoni, J.A. A Model for Optimal Operation in a Soy Oil and Flour Plant. 2nd Mercosur Congress on Chemical Engineering 1-10 (2005).

[40] Shahidi, F. Bailey's Industrial Oil and Fat Products. 1, (Wiley-Interscience: Hoboken, New Jersey, 2005).

### 5.3.3 Oil Cooler

The oil-acid mixture needs to be cooled to 32°C before the water wash.[32]
Need for review: low
Need for innovation: low

### 5.3.4 Water Wash

The cooled oil is mixed with distilled water to hydrate all the phospholipids into a semi-crystalline phase ready for centrifugation. Wash water is added at a rate of 1.0 wt% of the feed oil rate.[32]
Need for review: low
Need for innovation: low

### 5.3.5 Separation Heater

Prior to centrifugation the oil/aqueous mixture is heated to 85°C to improve the viscosity for centrifuging.[32]
Need for review: low
Need for innovation: low

### 5.3.6 Oil Centrifuge

The oil and aqueous phases are separated in a disk-stack centrifuge. Oil losses in the centrifuge are reported at 3.11 wt%.[31]
Need for review: medium
Need for innovation: low

## 5.4 Pongamia UOP Refining

### 5.4.1 Hydrodeoxygenation Heater

The hydrodeoxygenation reactor oil feed is heated to the reaction temperature of 350°C by combusting the propane rich waste gases in a fired heater.[33]
Need for review: low
Need for innovation: low

### 5.4.2 Hydrodeoxygenation Reactor

The hydrodeoxygenation reactor is down flow packed reactor divided into three stages, with an equicurent feed of oil and hydrogen. In each stage the triacylglyceride oil is reduced to straight alkanes, corresponding to the lipids, and propane, from the glycerol backbone. The UOP catalyst reduces the oil through both the deoxygenation reaction (60 wt%) and the hydrogenation reaction (40 wt%).[34] Stage 1 & 2 oil conversion is 90 wt%, while the final finishing stage operates at the higher 98 wt% conversion.[33]

**Deoxygenation Reaction:** 1 Glycerol Trioleate ($C_{57}H_{104}O_6$) + 6 Hydrogen ($H_2$) -> 3 Heptadecane ($C_{17}H_{36}$) + 1 Propane ($C_3H_8$) + 3 Carbon Dioxide ($CO_2$)

**Hydrogenation Reaction:** 1 Glycerol Trioleate ($C_{57}H_{104}O_6$) + 15 Hydrogen ($H_2$) -> 3 Octadecane ($C_{18}H_{38}$) + 1 Propane ($C_3H_8$) + 6 Water ($H_2O$)

Need for review: low
Need for innovation: low

### 5.4.3 Intermediate Gas Flash

Before the final finishing stage of the hydrodeoxygenation reactor all the waste gases ($CO_2$, $H_2$, $H_2O$, and short hydrocarbons - particularly propane) are flashed off. Model conditions of 100 °C and 35 bar were optimised for hydrocarbon recovery.
Need for review: medium
Need for innovation: low

### 5.4.4 Hot Hydrogen Stripper

A countercurrent stripper removes the waste light gases by using the dry hydrogen feed. This is particularly to reduce the water concentration (catalyst poison for the hydrocracking catalyst) to 100 pm.[34]

Need for review: low
Need for innovation: low

## 5.5 Pongamia Hydrocracking

### 5.5.1 Cracking Fired Heater

The cracking reactor feed is heated to the reaction temperature of 332°C by combusting the propane rich waste gases in a fired heater.[21]

Need for review: low
Need for innovation: low

### 5.5.2 Hydroisomerization/Hydrocracking Reactor

The straight chain alkanes produced in the hydrodeoxygenation reactor are unsuitable for use as aviation fuel and require further processes. These alkanes need to be cracked and isomerized to generate a suitable product for use as an aviation fuel substitute.[21] A cracking reactor operating at 332 °C, 5171 kPag, 0.5 hour$^{-1}$ LHSV, with a specialized UOP catalyst is able to convert around 50 wt% of the feed material into a aviation fuel fraction.[21] The remainder going to either diesel, naptha, or light gas fractions.

Need for review: medium
Need for innovation: medium

## 5.6 Pongamia Product Recovery

### 5.6.1 Hot Flash Vessel

High pressure flash vessel to remove light gases prior to product distillation. The temperature and pressure of the flash vessel have been modelled to minimise losses of the key aviation fraction.

Need for review: medium
Need for innovation: low

### 5.6.2 Product Distillation

The separation and recovery of the naptha, aviation, and diesel fractions has been modelled in an atmospheric crude distillation unit.[22]

Need for review: medium
Need for innovation: low

## 5.7 Pongamia Amine Scrubber

### 5.7.1 Amine Scrubber

The waste gas from the hydrodeoxygenation is treated to remove impurities, particularly carbon dioxide and water. An amine scrubber, a counter current gas-liquid contactor, with a 10 wt% MDEA (Methyl diethanol amine) absorbent. At the high pressure conditions used in this process a carbon dioxide removal efficiency of 71 wt%[13] has been reported. The hydrogen rich scrubbed gas is recycled as hydrogen feed to the hydrodexygenation reactor.

Need for review: low
Need for innovation: low

### 5.7.2 Degasifier

The rich MDEA solution from the amine scrubber is treated in a degasifier to remove the absorbed carbon dioxide.

Need for review: medium
Need for innovation: low

### 5.7.3 Carbon Dioxide Flash

The degasified carbon dioxide is flashed to recover trace water.
Need for review: low
Need for innovation: low

## 5.8 Pongamia Anaerobic Digestion

### 5.8.1 Anaerobic Digester

The extracted meal from the meal cyclone and hulls from the aspirator are digested anaerobically. A 31 hour residence time digestor has been reported to give a 24% biomass reduction of algae sludge.[36] 90% of the digester muds is recycled to algae ponds to allow undigested nutrients to be recycled.
Need for review: medium
Need for innovation: medium

## 5.9 Pongamia Cogeneration

### 5.9.1 Methane Boiler

The digester biogas is combusted in a methane boiler to generate high pressure (80 bar) steam for use in covering process electricity and steam requirements.
Need for review: low
Need for innovation: low

### 5.9.2 Turbogenerator

A multi-stage turbogenerator is used to generate electricity and generate the processes requirement for process steam.
Need for review: low
Need for innovation: low