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Microalgae cultivation for the production of algae jet fuel in the Netherlands

A feasibility study

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Abstract
Tata Steel produces the highest amount of CO_2 emissions of all private companies in the Netherlands. To mitigate these flue gas emissions, Tata Steel and the Schiphol group are proposing to use 1 Mton of this CO_2 for large scale microalgae cultivation in the Netherlands, with algae jet fuel (AJF) as the main targeted product. Although, literature indicates that algae based fuels are currently not commercially feasible, they show potential for the future. Furthermore, the potential of algae cultivation and processing in the Netherlands is currently unclear. To assess the potential of the microalgae to AJF cycle in the Netherlands, the technical design and costs of the three most used onshore cultivation methods and one pilot offshore cultivation method are investigated and modelled in this research. The considered reactor designs are the raceway open pond, tubular reactor, plate reactor and the offshore membrane enclosures for growing algae (OMEGA). Furthermore, the microalgae processing in the Netherlands to AJF and other valuable co-products is investigated and modelled for cost indications. The corrected total annual costs are: 1.14 to 1.20 billion euros/yr for the open pond system, 1.93 to 2.54 billion euros/yr for the tubular reactor, 2.64 to 3.73 billion euros/yr for the plate reactor and 6.97 to 7.29 billion euros/yr for the OMEGA reactor. These costs are divided into an initial investment and annual operating costs. For both the initial investment and operation, the costs are mainly driven by the algae cultivation. The revenue of the systems is based on the market size and price of the valuable algae biomass fractions (including AJF) obtained from processing, as well as the CO_2 permit revenue. This research indicates that it is currently not commercially feasible to cultivate and process microalgae for AJF in the Netherlands. For the future, the open pond system could become commercially feasible if cost reductions are achieved, since the AJF price and the CO_2 price needed for a commercially feasible system in the future are highly unlikely. Furthermore, a revenue assessment indicates that the AJF fraction has an insignificant value compared to the revenue created from the co-products. Therefore, future research should focus on the other fractions with a higher value and substantial market size to potentially obtain a commercially feasible system.

Executive summary

Introduction

Tata Steel wants to reduce its CO_2 emissions in the Netherlands. Therefore, Tata Steel and the Schiphol group intend to use 1 Mton of this CO_2 for large scale microalgae cultivation in the Netherlands. They propose to use the produced microalgae biomass for algae jet fuel (AJF) and other valuable co-products. However, the exact commercial feasibility of both the current and future algae to jet fuel cycle in the Netherlands is unclear and thus investigated in this research. This results in the following research question, which is answered through four sub-questions:

To what extent is it commercially feasible to use microalgae for the production of AJF on the short and long term in the Netherlands, while making use of 1Mton of CO_2 from Tata Steel annually?

1. What is the scale of the research, including algae strain, productivity and land occupation?
2. What are the costs for the cultivation of wet algae biomass in the Netherlands? What are the main cost drivers?
3. What are the costs for the algae processing? What are the main cost drivers?
4. What is the current and future commercial feasibility of the AJF, considering the total costs, the total revenue (including co-products, AJF and CO_2 permits) and possible future improvements?

Method

To answer the research question and sub-questions first a general considerations and preliminary assessment is made. In this assessment the microalgae strain used, the productivity of the microalgae and the land occupation are investigated. The productivity is assessed by using the photosynthetic efficiency of the three most used onshore cultivation methods (raceway open pond, tubular reactor and plate reactor) and one pilot offshore cultivation method (OMEGA). After that, the technical specifications for each of the cultivation methods, including harvesting/dewatering, are described. This technical design is used for the biomass cultivation costs. The same is done for the algae processing. Afterwards, the revenue per cultivation method is calculated by adding the revenue created by each valuable fraction obtained from the algae processing with the revenue obtained from CO_2 permits. These total costs and revenues for each cultivation method are used to assess the commercial feasibility, where a net present value (NPV) and payback period (PBP) are used as indicators. This is supplemented by an assessment of the future potential, by implementing cost reductions and higher AJF/CO_2 permit prices in the calculation. An overview of the method is given below.

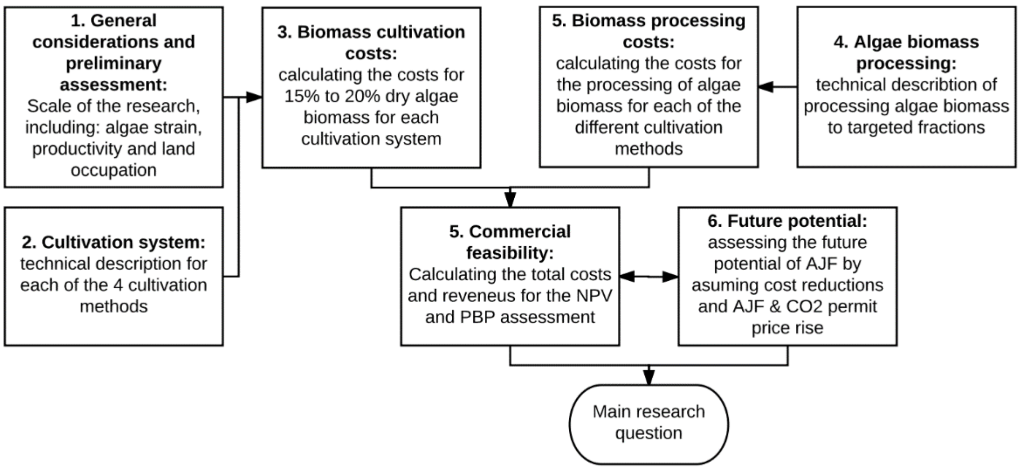


Figure A: Overview of the method

Results

The microalgae strains used in this research are the marine Nannochloropsis and freshwater Chlorella, as they have a high growth rate, a high lipid content and are able to survive in the Dutch environment for both the bioreactors and open pond systems. The productivities of these algae strains per cultivation method are indicated in the table below.

Table A: Daily dry weight production data of the different algae cultivation methods in the Netherlands

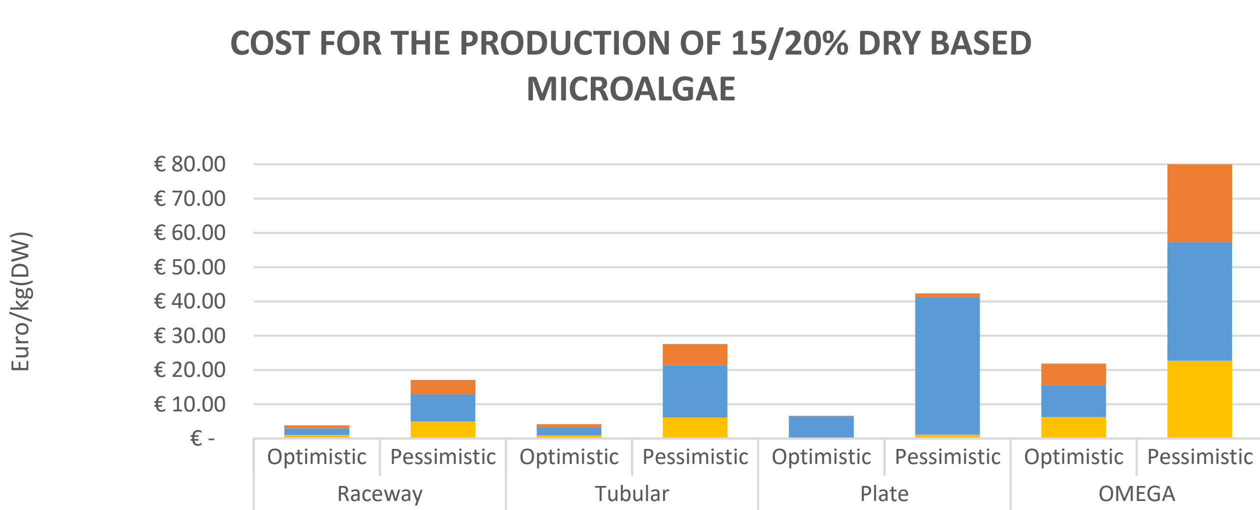
Method	Optimistic	Pessimistic	Unit
Productivity raceway pond	23.5	18.8	Ton/ha/yr
Productivity tubular reactor	47.1	23.5	Ton/ha/yr
Productivity plate reactor	78.5	42.4	Ton/ha/yr
Productivity OMEGA	23.3	23.3	Ton/ha/yr

Each of these cultivation methods uses a different amount of land. Since the available land in the Netherlands limited, these values were also calculated, as given below.

Table B: Total amount of land occupation for each cultivation method

Total amount of land needed per cultivation method					
Method	Optimistic		Pessimistic		
	Irradiated	Total land	Irradiated	Total land	Unit
Raceway open pond	90.9	113.6	113.6	142.0	Km^2
Tubular reactor	68.2	85.2	136.3	170.4	Km^2
Plate reactor	40.9	51.1	75.7	94.7	Km^2
OMEGA	140.8	201.2	140.8	201.2	Km^2

After these general considerations, the microalgae to AJF cycle consisting of algae cultivation and processing is investigated. The algae cultivation costs for the production of 15% to 20% dry algae biomass are given below.



	Raceway Optimistic	Raceway Pessimistic	Tubular Optimistic	Tubular Pessimistic	Plate Optimistic	Plate Pessimistic	OMEGA Optimistic	OMEGA Pessimistic
Total	€ 3.84	€ 17.12	€ 4.21	€ 27.56	€ 6.61	€ 42.36	€ 21.84	€ 79.99
Other fixed costs	€ 0.90	€ 4.22	€ 0.91	€ 6.15	€ 0.19	€ 1.13	€ 6.23	€ 22.67
Operating costs	€ 1.88	€ 7.94	€ 2.39	€ 15.26	€ 6.24	€ 40.11	€ 9.37	€ 34.65
MEC	€ 1.06	€ 4.96	€ 0.91	€ 6.15	€ 0.19	€ 1.13	€ 6.23	€ 22.67

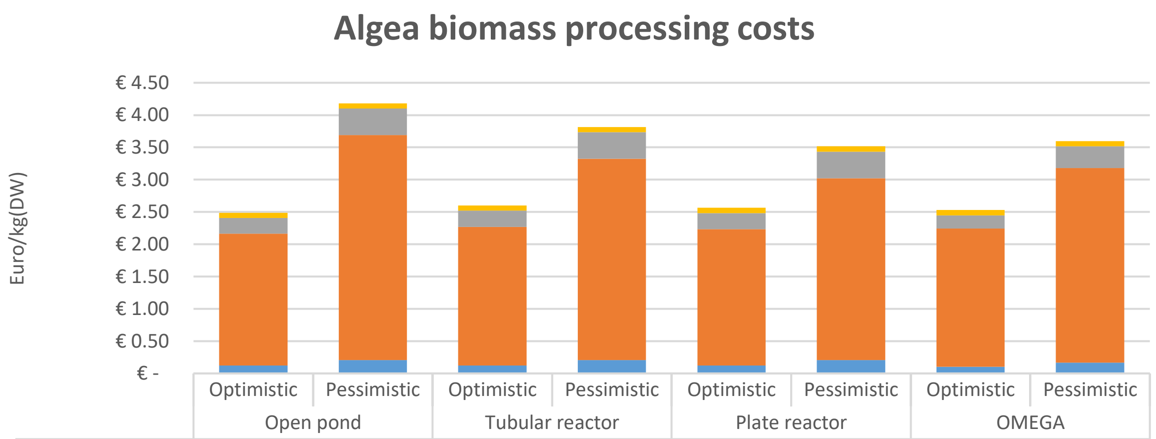
Scenario

Figure B: Costs for the production of dry algae biomass for each of the 4 cultivation methods

From this figure, it becomes clear that the OMEGA method has the highest cultivation costs and the raceway open pond the lowest. Furthermore, the figure indicates a large difference between the optimistic and pessimistic scenario, which is mainly caused by the difference in photosynthetic efficiencies. The photosynthetic efficiency influences productivity as well as the land occupation of the different methods. Besides, the cost division indicates that the operating costs account for the highest share of the total costs.

The detailed cost division, given in appendix C, indicates that other significant cost drivers for cultivation are: land occupation, medium filters, culture medium premix, piping, buildings and maintenance. Furthermore, all cultivation methods show method-specific cost drivers: the open pond reactor has very high costs for a PVC liner inside the ponds; the tubular reactor has very high investment and operating costs for the circulation pump; the plate reactor has very high operating costs for the air blowers (which power the circulation); and the OMEGA method has very high investment costs for the offshore bioreactor.

Furthermore, the 15% to 20% dry weight algae biomass has to be processed. The processing consists of fractionation and lipid refining into jet fuel. The total costs of the algae biomass processing are given below.



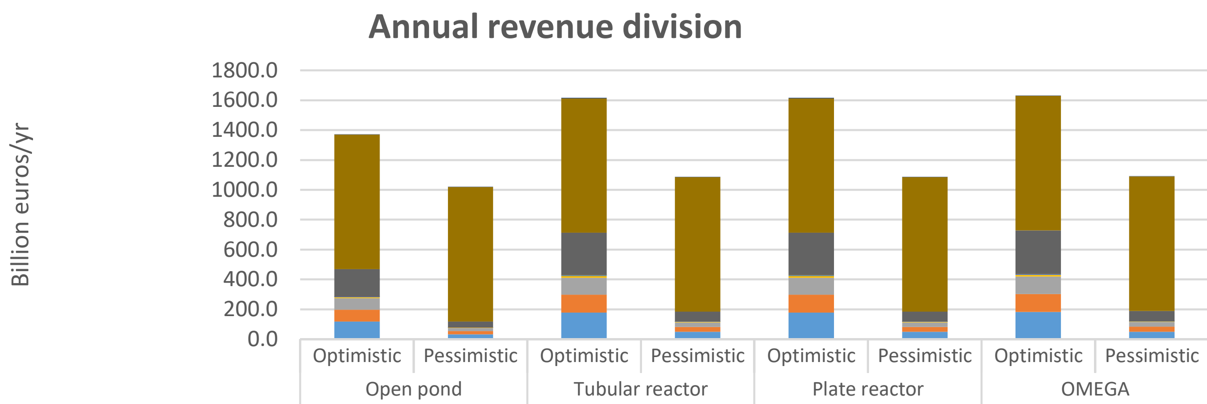
	Open pond		Tubular reactor		Plate reactor		OMEGA	
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
Total	€ 2.49	€ 4.18	€ 2.60	€ 3.81	€ 2.56	€ 3.52	€ 2.53	€ 3.60
Jet refining	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08
Other fixed costs	€ 0.24	€ 0.41	€ 0.25	€ 0.41	€ 0.25	€ 0.41	€ 0.20	€ 0.34
Operating costs	€ 2.04	€ 3.48	€ 2.14	€ 3.11	€ 2.11	€ 2.82	€ 2.14	€ 3.01
MEC	€ 0.12	€ 0.21	€ 0.13	€ 0.21	€ 0.13	€ 0.21	€ 0.10	€ 0.17

Figure C: Algae biomass fractionation costs per cultivation method for both the optimistic and pessimistic scenario

This figure indicates that similar to the cultivation, the operating costs account for the highest cost share in the processing. Again, there is a strong difference between the optimistic and pessimistic scenario which is mainly caused by the fact that in the optimistic scenario the biomass concentration is assumed to be 20%, whereas a value of 15% was adopted in the pessimistic scenario. Therefore, more water is processed in the pessimistic scenario, which makes for a less efficient process. The total processing costs are calculated by modelling the exact fractionation process, as this process is not yet used on a large scale and costs are thus uncertain. For the jet refinery, a premium is assumed, as the factories that can produce jet fuel from lipid fractions are already widely used. This premium is added to the total fractionation costs.

The exact cost division in appendix E indicates that the main cost drivers for algae processing are: bead mill energy, spray dry energy, utilities (such as PEG400, Isopropanol, hexane), distillation column for methanol recovery, cooling energy (especially for the bead mill), the labour costs and the jet refinery. These drivers are the same for each method, as the used processing assumptions are also the same for each method. The difference between the cultivation methods is caused by the labour costs, which are linked to the land occupation (which is different for each method).

The revenues of the valuable fractions obtained from the algae processing are given below.



	Open pond		Tubular reactor		Plate reactor		OMEGA	
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
Total	1372.3	1020.4	1615.9	1088.2	1615.9	1088.2	1630.9	1091.9
CO2 permits	2.0	2.0	3.0	3.0	3.0	3.0	2.5	2.5
Pigments high value	900.0	900.0	900.0	900.0	900.0	900.0	900.0	900.0
Pigments low value	183.6	39.6	283.0	66.9	283.0	66.9	289.3	68.7
Propane	0.6	0.2	0.8	0.2	0.8	0.2	0.9	0.2
LPG	0.4	0.1	0.6	0.2	0.6	0.2	0.6	0.2
Naphtha	0.8	0.2	1.2	0.3	1.2	0.3	1.3	0.3
Diesel	3.4	0.9	5.1	1.4	5.1	1.4	5.2	1.4
Jet	7.7	2.1	11.6	3.2	11.6	3.2	11.8	3.3
Carbohybrates	75.7	20.8	113.5	31.2	113.5	31.2	115.9	31.9
Non-water soluble proteins	79.5	21.8	119.2	32.8	119.2	32.8	121.7	33.5
Water soluble proteins	118.6	32.6	177.9	48.9	177.9	48.9	181.7	50.0

Scenario

Figure D: Annual revenue division

This revenue division indicates that the highest revenue is by far obtained from the pigment fractions. The same revenue for the high value pigments is indicated for each scenario, as this market (1 kton/yr) is saturated for all the scenarios. Furthermore, this figure indicates that the revenue created by the jet fuel fraction is insignificant compared to the total revenue created. This low revenue is caused by the low jet fuel price, which is currently 480 €/ton.

The cultivation, processing costs and the revenues are incorporated in the calculation of an NPV and PBP for all the cultivation methods and scenarios. The values are given below.

Table C: *Current NPV and PBP for each cultivation method for the whole project*

	Open pond		Tubular		Plate		Omega		
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Unit
NPV	1.60	-1.05	-2.86	-13.14	-9.30	-23.98	-51.34	-53.38	Billion Euros
PBP	6.54	11.61	17.91	-24.35	-2.34	-0.86	-19.16	-17.08	Years

The market of the lipid and other fractions is furthermore investigated to assess future improvement of revenues, as shown in figure E. In this figure, the relation between the costs per cultivation method and the revenue of each fraction is depicted. All the fractions below the method specific cost lines have higher costs then revenues and all the fractions above the cost lines have higher revenues than costs.

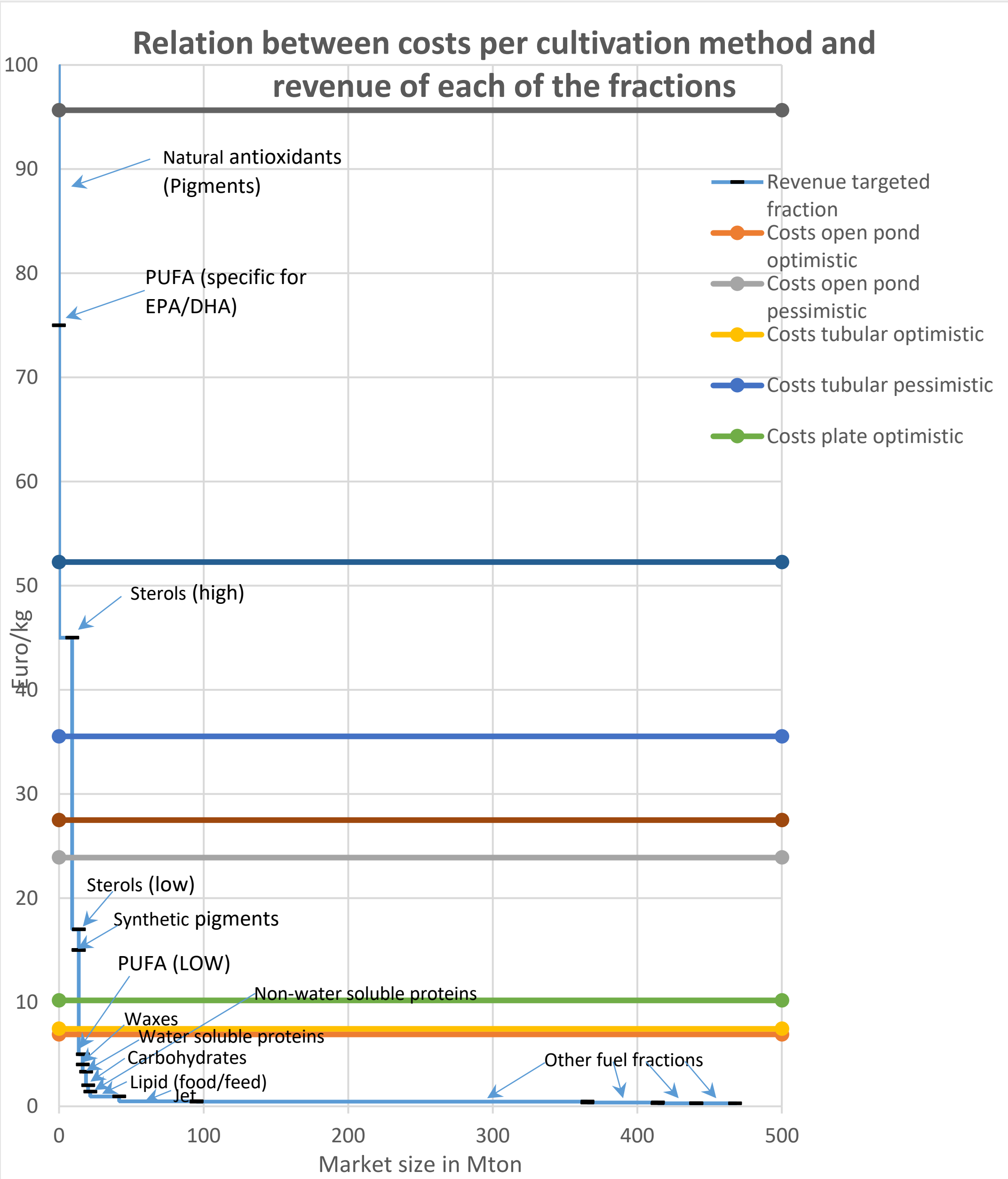


Figure E: Overview of revenue of each fraction compared to the costs per kg of each cultivation method

From this figure, it can be deduced that the jet fuel and other fuel fractions have the lowest values of all fractions. Other markets for the lipid fraction are indicated in this figure, such as: sterols, waxes, PUFA and lipid for food and feed. All these markets show a higher price than the jet fuel price and will therefore be explored first. However, when looking at the market size on the x-axis, it is clear that the fuel market is much bigger than the other markets. Still, due to the low price, the fuel market will only be entered in the case that the higher lipid markets are saturated (35 Mton/yr), which will probably not happen in the near future, as the current lipid production rates are 8.0 to 44.8 kton/yr.

Furthermore, the future price developments of the AJF and CO_2 permits needed to make all cultivation methods commercially feasible (NPV=0) are investigated. The prices of both the AJF and CO_2 permits are given below.

Table D: Prices HEFA for future commercial feasibility

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	-11	27	Thousand Euro/ton
Tubular	14	219	Thousand Euro/ton
Plate	43	399	Thousand Euro/ton
OMEGA	230	869	Thousand Euro/ton

Table E: Prices CO_2 permits for future commercial feasibility

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	€ -438	€ 295	Euro/ton
Tubular	€ 530	€ 2,418	Euro/ton
Plate	€ 1,713	€ 5,291	Euro/ton
OMEGA	€ 11,316	€ 11,766	Euro/ton

Note that a negative value is given for the optimistic scenario of the open pond system, as this system is currently already commercially feasible. All the other prices are extremely high compared to the current jet fuel price (480 €/ton) and the CO_2 permit price (5 €/ton).

Finally, an optimal scenario based on literature is assessed, in which an AJF price of 1200 €/ton, a CO_2 permit price of 120 €/ton and a 30% cost reduction for both the initial investment and the operating costs are assumed. The final NPV and PBP are given below.

Table F: Future NPV and PBP of the optimal scenario

	Open pond		Tubular		Plate		Omega		
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Unit
NPV	5.39	2.49	3.18	-5.57	-1.33	-13.16	-30.81	-33.82	Billion Euros
PBP	3.16	5.23	5.09	53.58	2215.43	-1.13	-34.43	-23.50	Years

Conclusion

The current NPV and PBP show only a positive value for the optimistic scenario of the open pond system, whereas the pessimistic scenario of the open pond system and all the scenarios of the other cultivation methods show a negative value. As it is unclear whether the optimistic or pessimistic scenario of the open pond system is more accurate, all cultivation methods are currently considered non-commercially feasible. The optimal scenario, indicates that both the scenarios of the open pond system and the optimistic scenario of the tubular reactor show a positive NPV and sufficient PBP. Therefore, even in the unlikely optimal scenario, only the open pond reactor could be commercially feasible as it is unclear whether the optimistic or pessimistic scenario of the tubular reactor is more accurate. Furthermore, a more in-depth investigation of the lipid fraction shows that other markets with higher value will probably be explored before the fuel market. Finally, the AJF and CO_2 permit price rise needed to make each system commercially feasible (NPV=0) in the future will probably never be met. Therefore, it is very unlikely that the use of microalgae for the production of AJF will be commercially feasible in the future.

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1. Introduction

1.1 Problem definition

The current fossil hydrocarbon fuel economy is considered unsustainable (Collotta et al., 2016). Fossil based fuels emit atmospheric CO_2, which causes climate change and the accompanied negative influences such as temperature rise and freshwater depletion (IPCC, 2008). Recently, 195 countries adopted the first universal climate agreement in Paris to mitigate this climate change (European Commission, 2016a). The agreement states that governments need to mitigate their emissions, be transparent and strengthen societies so that they are resilient to adaptation (European Commission, 2016b). One of the 195 countries is the Netherlands (United Nations, 2015).

With around 6% of the total Dutch emissions, Tata Steel, a steel company situated in the Netherlands, has the highest CO_2 emissions of all private companies in the Netherlands (Bouma, 2012; PBL, 2015). This translates to roughly 12 million tons of CO_2 in 2014 (Tata Steel, 2016). However, Tata Steel is trying to reduce its emissions by improving energy efficiency, perform circularity of operations and improve yields (Tata Steel, 2016). Furthermore, Tata Steel is currently investing in ULCOS (Ultra-Low CO_2 Steelmaking) by a partnership of 48 companies from 15 European countries, with the aim to reduce the CO_2 emissions by around 50% in 2050 (Tata Steel, 2017). Although these improvements will lower the CO_2 emissions, the energy intensity for steel production is so high that it is impossible to change the reliance on fossil fuels without undermining process efficiency and costs in the near future (Carpenter, 2012). Therefore, Tata steel is currently investigating the opportunities to capture the CO_2 of their steel blast furnace flue gas and use it for value creation (Louwerse & Koelemeijer, 2016).

The transport sector, which is responsible for 25% of the total EU emissions, is also considered a substantial emitter of CO_2 (European Commission, 2016d). Within this sector, the aviation industry, with roughly 3% of the total emissions in the EU, is one of the largest and fastest-growing emitters (European Commission, 2016c). To make sure that these emissions are lowered, an ambitious target of 50% CO_2 reduction compared to 2005 is set for 2050 (ATAG, 2016). Renewable jet fuels (RJF) are proposed as a solution to reach this target (de Jong et al., 2015; Smith et al., 2017). RJF are hydrocarbon based fuels that do not originate from fossil fuels (de Jong et al., 2015; Smith et al., 2017), also known as “drop-in” biofuels (Karatzos, McMillan, & Saddler, 2014). These biofuels can lower the greenhouse gas emission significantly, without implementing profound changes in current technologies (Carere et al., 2008; Collotta et al., 2016). Aviation tests already indicate that flying with bio jet fuels meet the performance required (Macfarlane, Mazza, & Allan, 2011; W.-C. Wang et al., 2016). However, the current alternatives are still too expensive (de Jong et al., 2015).

Biofuels, can be divided into three general categories: first- (food-crop feedstock), second- (discarded mass) and third-generation biofuels (Saladini et al., 2016). Third generation biofuels, also known as algae, are considered the most viable option due to their high lipid content, high growth rate, ability to improve strains, production of valuable co-products and the fact that they do not compete for fertile land in the meantime (Hannon et al., 2010; Stepan et al., 2016; Trent et al., 2012). Especially in the Netherlands, with low land availability, third generation biofuels could have potential for the fuel market (RLI, 2016). Furthermore, as most algae are photoautotrophic these algae could be used to take up CO_2 (Eriksen, 2008).

1.2 Background Tata Steel and Schiphol Group

Working with these concepts, Tata Steel and the Schiphol Group successfully investigated the possibility to grow algae on blast furnace flue gasses produced by steel manufacturing (Louwerse & Koelemeijer, 2016). However, now they intend to apply this process on a large scale. Therefore, they are proposing to build a pilot plant that could uptake 1 Mton of CO_2 from Tata Steel annually. To tackle both the carbon capture for Tata Steel and to gain insights in algae biomass alternatives for jet fuel, Tata Steel and the Schiphol Group are proposing to use the potential algae biomass produced from this pilot plant for the production of jet fuel in the Netherlands (Louwerse & Koelemeijer, 2016). Furthermore, Tata Steel and Schiphol want to find out what the potential of algae cultivation is in the Netherlands (especially offshore North Sea) and how competitive jet fuel produced from algae is compared to conventional jet fuel (Louwerse & Koelemeijer, 2016). However, there is still a lot of uncertainty concerning large scale algae cultivation and biomass processing in the Netherlands. Therefore, prior to formulating a research question, an in-depth literature review concerning algae cultivation and processing is performed.

1.3 Literature review

Literature indicates that there are three types of algae: microalgae, macro algae (seaweed) and cyanobacteria (blue algae) (Singh et al., 2011). In this research cyanobacteria are included in the microalgae category, as they are both micro-organisms that have high lipid mass potential, which is the base product for oil and are cultivated in similar setups (Hannon et al., 2010; A. Singh et al., 2011). Nevertheless, it is noted that these organisms are not algae but photosynthetic bacteria (Hannon et al., 2010; A. Singh et al., 2011). From the algae types, the microalgae are considered to have the highest potential for the production of biofuels (Chisti, 2007; Stepan et al., 2016; Trent et al., 2012; Zhu, Hiltunen, & Takala, 2012). They produce 15-300 times more oil than traditional oil crops per area, can double their biomass in twenty-four hours and can fixate large amounts of CO_2 (Chisti, 2007; Collotta et al., 2016; Vyas, Verma, & Subrahmanyam, 2009; Zhang, 2015). Macroalgae show less potential due to their high energy need in the oil extraction phase of the fuel production process (Bastianoni et al., 2008). Furthermore, they show small potential for CO_2 uptake (Titlyanov & Titlyanova, 2010) and are not considered as viable for fuel production as microalgae (Chen et al., 2015). Therefore, macroalgae are not taken into account in this research.

To grow the microalgae, cultivation systems are used (Chisti, 2007). Literature indicates that there are generally two types of onshore microalgae cultivation methods that allow for artificial CO_2 aeration: open pond systems and a variety of photobioreactors (i.e. tubular, column and plate) (Chisti, 2007; Ruiz et al., 2016). Furthermore, literature research suggest one potential offshore photobioreactor method that could uptake flue gas CO_2, developed by NASA, known as offshore membrane enclosures for growing algae (OMEGA) (Novoveská et al., 2016; Trent, 2013; Trent et al., 2012; Wiley et al., 2013). Microalgae are generally cultivated in contained closed basins/membranes or reactors (Park, Craggs, & Shilton, 2011; Ruiz et al., 2016). These systems need nutrients such as CO_2, nitrate and phosphor to survive, as they are not supplied by the environment (Park et al., 2011; Ruiz et al., 2016). Therefore, these systems show potential for flue gas CO_2 uptake (Zhang, 2015).

To obtain the biomass from these cultivation systems, a variety of harvest and dewatering techniques is used (Brennan & Owende, 2010). The processing of algae biomass, obtained from these cultivation systems, into valuable products consists of fractionation and refining, with a wide variety of options, depending on the desired products (Brennan & Owende, 2010; Ruiz et al., 2016; Xu, (Wim) Brilman, Withag, Brem, & Kersten, 2011). For the production of jet fuel specifically, literature indicates that hydro processed esters and fatty acids (HEFA), also known as Hydro processed Renewable Jet (HRJ), is the most promising option (Carter, 2012; de Jong et al., 2015) of the certified jet fuels that could be used in the short-term (Wang et al., 2016). Therefore, this research focusses on the lipid extraction and refinery options to produce HEFA.

Literature further indicates that the cycle from algae to jet fuel consists of the following main steps: algae strain selection, cultivation, harvesting & dewatering, fractionation and refining (Cox, Renouf, Dargan, Turner, & Klein-Marcuschamer, 2014; Mata, Martins, & Caetano, 2010; Pechsiri et al., 2016; Ruiz et al., 2016). However, the optimal design differs per region (Ruiz et al., 2016), as the factors which cause microalgae to grow such as sunlight, nutrients (including CO_2) and a certain temperature (Chisti, 2007; Park et al., 2011) also differ per region (Ruiz et al., 2016). The extent to which these factors influence the growth of the algae depends on the algae species; some algae grow better in high light intensity, while others grow better in low light intensity (Park et al., 2011). Furthermore, the cultivation method influences the growth, as for example the temperature of a closed system is higher than an open system (Norsker et al., 2011).

Finally, the production of biofuels from microalgae is not yet commercially viable at current fuel prices (Ruiz et al., 2016; Saladini et al., 2016). However, the possibility of implementing certain improvements makes it a promising option nonetheless (Ruiz et al., 2016; Saladini et al., 2016). The biomass of the algae consists of lipids, proteins and nucleic acid and other valuable elements (e.g. fatty acids, vitamins, pigments, enzymes and antioxidants) depending on the species (Becker, 2007; Chisti, 2007; Vassilev & Vassileva, 2016). To become commercially feasible it is important that the full potential of the algae and their products is utilised by advanced fractionation and refinery, because most co-products have high economic value and thus could increase the commercial feasibility of the algae cultivation and processing (Ruiz et al., 2016). Furthermore, integration of processes such as wastewater treatment in algae basins/reactors (Park et al., 2011; Ruiz et al., 2016), recycling options of materials (Collotta et al., 2016; Cox et al., 2014) and new conversion technologies (such as hydro thermal liquefaction (HTL)) could improve energy efficiency and costs (Chiaramonti et al., 2017). However, this technology is not yet fully developed (Chiaramonti et al., 2017) and therefore the influence on advanced refinery for example is unclear. Therefore, microalgae could possibly be a commercially feasible option for fuel production in the future (Chiaramonti et al., 2017; Ruiz et al., 2016).

1.4 Gap in Literature and research question

As mentioned above, Tata Steel wishes to use their flue gasses for algae cultivation in the Netherlands, whereas the Schiphol Group want to investigate the potential of algae jet fuel (AJF). However, they lack evidence which cultivation method is commercially most feasible in the Netherlands for the production of algae biofuels (in this case AJF). Pilot projects and institutions already showed that onshore cultivation of microalgae is possible in practice in the Netherlands (Brinker, 2014; Wolkers et al., 2011). However, these projects do not give a clear cost indication (Brinker, 2014; Wolkers et al., 2011) or do not take all commercial scaled cultivation methods (that allow for CO_2 aeriation) and or the processing of biomass to jet fuel (Norsker et al., 2011) into account. Therefore, despite indications that the technology is currently not commercially feasible (de Jong et al., 2015), the expected time towards commercial deployment remains unclear. This research will aim to find out which aspects are responsible for the highest costs for the algae cultivation and processing, as well as what the main revenue streams are. By identifying these parameters for the whole algae to jet fuel cycle in the Netherlands, the time to commercial deployment can be estimated more clearly. Furthermore, since the potential of offshore microalgae cultivation is unclear, the OMEGA method, currently being the only offshore cultivation method that allows for CO_2 uptake, is also investigated in this research.

Hence, due to the many steps in and alternatives of the algae-to-AJF cycle, this research focusses on the current most used cultivation methods and the pilot offshore method that allow for CO_2 aeration and large-scale implementation. The OMEGA method is taken into account as both Tata Steel and the Schiphol Group have indicated an interest in the potential of offshore algae cultivation. The methods considered in this research are: the raceway open pond, plate reactor, tubular reactor and the OMEGA (Chisti, 2007; Jorquera, Kiperstok, Sales, Embiruçu, & Ghirardi, 2010; Mata et al., 2010; Molina Grima et al., 2003; Oncel & Sukan, 2008; Trent, 2013). This research aims to sketch the current commercial feasibility of algae cultivation for AJF. Furthermore, the future potential of the different cultivation methods and processing is discussed. These uncertainties and consideration result in the following research question:

To what extent is it commercially feasible to use microalgae for the production of AJF on the short and long term in the Netherlands, while making use of 1Mton of CO_2 from Tata Steel annually?

To answer this research question the following sub-questions are formed:

1. What is the scale of the research, including algae strain, productivity and land occupation?
2. What are the costs for the cultivation of wet algae biomass in the Netherlands? What are the main cost drivers?
3. What are the costs for the algae processing? What are the main cost drivers?
4. What is the current and future commercial feasibility of the AJF, considering the total costs, the total revenue (including co-products, AJF and CO_2 permits) and possible future improvements?

1.5 Research relevance

This study is a scoping study of the production of algal jet fuel from microalgae, aerated with CO_2, in the Netherlands. This is the first study in its kind with these specific components. Therefore, this study presents a basic outline of an algae to jet fuel cycle, while focussing on the different cultivation methods and their potential. As this study is considered a starting point, a lot of estimates are made, which causes a high level of uncertainty. However, this study sets a baseline which gives insights in possible relevant further in-depth research and algae cultivation research in other regions.

Possible issues with double counting are taken into account in this research. CO_2 mitigation is only obtained by Tata Steel, as the microalgae grow on CO_2 from flue gas and not directly from the air. For Schiphol, there is no CO_2 mitigation, as airplanes emit roughly the same amount of CO_2 independent of fuel source. Therefore, the term AJF is used in this research instead of renewable jet fuels (RJF).

1.6 Outline

In the next section of this proposal the methodology for answering the research questions is proposed, including the theory and method. Thereafter, the scale of the research is assessed by a preliminary assessment. Next, the cultivation system design is described and the costs for the algae biomass obtained from the cultivation system are calculated. Subsequently, the algae biomass processing is treated, including algae biomass composition, algae biomass processing and algae jet refinery and the accompanying costs. Finally, the current and future commercial feasibility is assessed and discussed.

2. Theory

This section explains the fundamental background of the full algal to jet fuel cycle, including cultivation and processing. The fuel cycle steps and the considerations for each step are discussed separately. Furthermore, commercial feasibility is explained including the way in which the commercial feasibility of an algae fuel cycle can be assessed.

2.1 Algae to jet fuel cycle

The algal fuel cycle generally consists of two steps, the cultivation, where the algae are grown and the biomass is obtained, and the processing (Baicha et al., 2016). The cultivation system consist of: algae strain selection, cultivation, harvest and dewatering (Cox et al., 2014; Mata et al., 2010). The processing consists of fractionation (including lipid extraction) and refinery (Collotta et al., 2016; Cox et al., 2014; Klein-Marcuschamer et al., 2013; Mata et al., 2010). However, how to fill in each step specifically differs amongst studies (Collotta et al., 2016). Therefore, the considerations for each of the steps are given below.

Cultivation

2.1.1 Algae strain selection

Selection of an algae strain, is the first step of the cultivation (Chisti, 2007). The algae species for biofuels ideally have a high lipid content, high growth rate and allow for easy harvesting and lipid extraction (Chisti, 2007). Furthermore, the algae culture should be able to survive in robust environments and be stronger than wild cultures in the case of an open pond system (Brennan & Owende, 2010). Besides, the algae should have a high CO_2 sinking capacity (autotrophic) and a low demand for nutrients needed for growth (Brennan & Owende, 2010). Furthermore, the algae strain should withstand temperature differences caused by seasonal variations and is preferably able to produce viable co-products (Brennan & Owende, 2010). Finally, the algae strain needs to have a high photosynthetic efficiency and is preferably able to self-flocculate[1], to ensure lower harvesting costs (Brennan & Owende, 2010). Currently, the *chlorella* algae species is mostly used, although a lot of research is done to improve algae strains in general (Hulsman, Reinders, & van Aalst, 2011). The optimal algae strain depends on the cultivation system, as the medium differs between open and closed systems (Jorquera et al., 2010).

2.1.2 Cultivation method

The strain selection is followed by cultivation of the microalgae. The customary way to cultivate microalgae is in an open pond system or in a photobioreactor (Chisti, 2007; Richardson, Johnson, & Outlaw, 2012). The open pond system is the oldest and simplest way to produce microalgae, of which the raceway pond is the dominant design (Chisti, 2007; Richardson et al., 2012). The photobioreactors can be subdivided into tubular reactors, plate reactors and vertical column reactors (Brennan & Owende, 2010). As mentioned before, recently NASA developed the OMEGA method, which is an offshore photobioreactor (Trent, 2013). Both the open pond and photobioreactor systems have the possibility to use CO_2 aeration (Chisti, 2007); especially closed photobioreactors need artificial aeration to maintain optimal growth (Brennan & Owende, 2010; Zhang, 2015). A more detailed description of these cultivation systems can be found in section 5.2.

Algae cultivation systems have different aspects that influence growth: CO_2 capture, light, nutrients removal and temperature (FAO, 2009; Florentinus, Hamelinck, de Lint, & van Iersel, 2014; Mata et al., 2010; Skjånes, Rebours, & Lindblad, 2013). Generally, around 1.8 tonnes of CO_2 is needed to grow 1

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[1] Self-flocculation is a process in which an algae species is able to form a cluster of algae cells, as the algae cling together by themselves, without addition of chemicals or other substances (Molina Grima et al., 2003)

tonne of biomass (FAO, 2009). However, the efficiency of the uptake differs a lot among cultivation methods (Ketheesan & Nirmalakhandan, 2012). Light is needed for algae to perform photosynthesis and thus to grow and capture CO_2 (FAO, 2009). However, only 45% of visible light spectrum (400-700nm) can be used for the photosynthesis (FAO, 2009). Therefore, the cultivation system must aim to maximize solar capture (Chisti, 2007). Nutrients that generally need to be supplied to these systems besides CO_2 are nitrogen (N) and phosphorus (P) (FAO, 2009). These sources can be supplied through agricultural fertilizer (Chisti, 2008) or by cheaper options such as wastewater, flue gas or nutrient recycling (Elmoraghy & Farag, 2012; Lardon et al., 2009; Minowa & Sawayama, 1999; Zhang, 2015). Finally, the temperature influences algae growth (FAO, 2009). In the Netherlands, the general growth season for algae runs from April till November, due to very low temperatures and the accompanied low growth rates in the remaining months (FAO, 2009). The growth season can be expanded by artificial cooling or heating by for example a heat exchanger (Chisti, 2007), a sprinkler system (Richardson et al., 2012) or residual heat from the industry (FAO, 2009). The exact growth season is assessed in a later stage of this research.

2.1.3. Harvest & dewatering

After the cultivation of the microalgae, the biomass is harvested and partly dewatered (Mata et al., 2010). The harvesting method depends on the characteristics of the microalgae, the desired products and the cultivation method (Brennan & Owende, 2010). Harvesting in general consists of two stages: bulk harvesting and thickening (Brennan & Owende, 2010). Bulk harvesting aims to separate the algae biomass from the medium, with generally a 2% to 7% solid matter concentration (Brennan & Owende, 2010). The methods used for bulk harvesting are flotation, flocculation, gravity sedimentation and electrophoresis techniques (Brennan & Owende, 2010; Chen et al., 2011; Uduman et al., 2010). Thickening aims to improve the solid matter concentration by processes such as filtration, centrifugation or ultrasonic aggregation (Brennan & Owende, 2010). This step is generally more energy intensive (Brennan & Owende, 2010). The final solid content obtained differs per method (Brennan & Owende, 2010). Extra information for each of the harvest and dewatering methods is given in appendix A. Depending on the lipid extraction method, the biomass is often dried to obtain roughly 95% solid biomass content (Brennan & Owende, 2010; Xu et al., 2011). Which is done by methods such as sun drying, fluidised bed drying, drum drying, spray drying, low-pressure shelf drying and reactance window technology drying (Brennan & Owende, 2010).

Processing

2.1.4 Fractionation

After the biomass is harvested, the lipids and other valuable fractions from the biomass are obtained (Mata et al., 2010). There are a lot of alternatives for the fractionation and extraction of biofuels and other valuable co-products from algae biomass (Brennan & Owende, 2010; Ruiz et al., 2016). Specifically for biofuels, there are two main routes to extract lipids from biomass, which entail a dry route and a wet route (Xu et al., 2011). For the dry route, the biomass is dried as mentioned before. For the wet route, drying is no longer needed (Xu et al., 2011). However, usually cell disruption or other types of pre-treatment are used to make the lipids more accessible in the wet algae sludge extraction (Dong, Knoshaug, Pienkos, & Laurens, 2016).

For the extraction of lipid from the algae, different methods such as pyrolysis (Chiaramonti et al., 2017), hexane extraction (Halim et al., 2011), direct transesterification, supercritical CO_2 lipid extraction or a mixture of other solvents are used (Molina Grima et al., 2003; Soares et al., 2014). However, most studies use hexane extraction (Cox et al., 2014; Klein-Marcuschamer et al., 2013; Xu et al., 2011) or pyrolysis (Chiaramonti et al., 2017). For the extraction process of lipid from algae biomass a balance between drying efficiency and cost-effectiveness has to be established, to maximize energy output (Brennan & Owende, 2010). Furthermore, properties of the cell membrane are important for the fuel extraction, as a cell membrane may impede the extraction (Brennan & Owende, 2010).

As this research aims to use the full potential of the microalgae, the algae are separated in the different biomass fractions, such as: lipids, hydrocarbons and proteins (Ruiz et al., 2016). The challenging part of this process is to separate each of the different algae fractions without damaging the other fractions (Vanthoor-Koopmans, Wijffels, Barbosa, & Eppink, 2013), especially because each fraction supplies its own market (see figure 1).

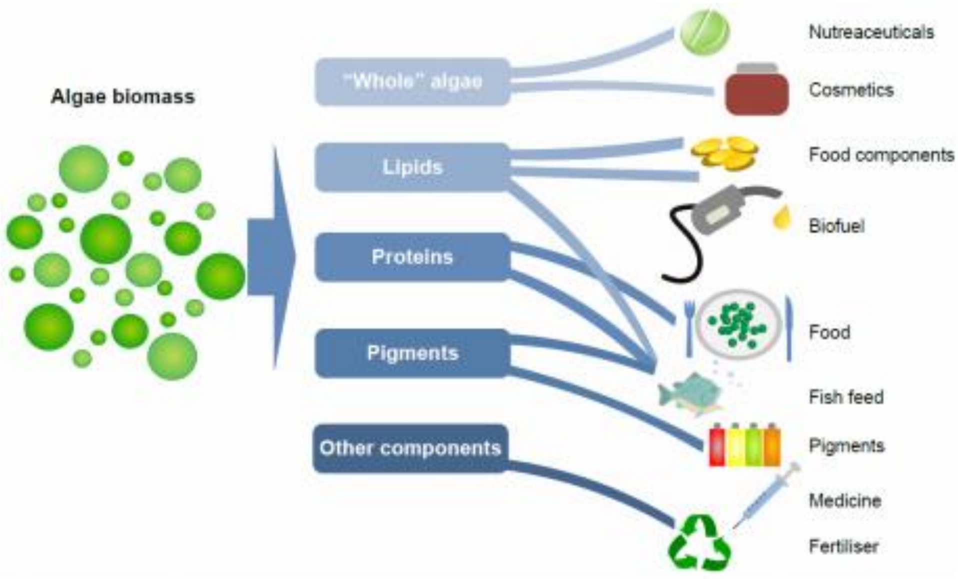


Figure 1:Overview products of refinery (Slegers, 2014)

The exact fraction extraction process of the lipids and other valuable products are explained later in this research.

2.1.5 Refinery

The final step of the fuel cycle is to convert the algae lipid fraction into valuable products, including AJF. As can be seen in figure 1, biofuels are produced from the lipids in the algae. However, these lipids need to be put through a refinery process towards the hydro processed renewable jet (HEFA). After the lipid extraction, the lipid is often degummed, which is a process that removes small phospholipids (Halim et al., 2011). Thereafter, the lipid is upgraded by hydrotreatment, cracked/isomerised and fractionised by a separation process (for example distillation) into jet fuel and other biofuel fractions (Klein-Marcuschamer et al., 2013; Wang et al., 2016). Later in this research the exact refining process will be discussed in more detail.

2.2 Commercial feasibility

In this research, the commercial feasibility is measured by the net present value (NPV). The NPV is used as it includes the time preference of earning money now compared to earning it later by taking into account the discount rate (Blok, 2006). Moreover, it contains all the aspects on which commercial feasibility depends: initial investment (CAPEX), annual or operational costs (OPEX) and annual benefit (Richardson et al., 2012). The NPV formula is given as follows (Blok, 2006):

$$NPV = \sum_{i=0}^{n} \frac{B_i - C_i}{(1+r)^i}$$

NPV= net present value in year 0
B_i= the benefits in year i
C_i= the costs in year i (at the beginning of the project this could include the initial investment)
r= discount rate
n= the lifetime of the project

This formula is used for energy systems with annually fluctuating costs and production rates (Blok, 2006; Blok & Nieuwlaar, 2016). However, a simplified formula can be used when costs and production rates are the same each year, which is the case in this research as the same assumptions are used each year:

$$NPV = -I + \left(\frac{B - C}{\alpha}\right)$$

$$\alpha = \frac{r}{1 - (1+r)^{-L}}$$

I= Initial investment (CAPEX)
B= Annual benefits (annual revenue)
C= Annual costs (OPEX)
α= The capital recovery factor (or annuity factor)
r= Discount rate (normally around 10%)
L= Lifetime of the project (in yr)

The annuity factor translates the annual costs and benefits into total values for the total lifetime of the project, as there is a discrepancy between the time of spending and earning money (Blok, 2006; Blok & Nieuwlaar, 2016). Literature indicates that the costs indicated in the formula consist out of aspects such as land, cultivation system investment, CO_2 delivery, general machinery, supply of water/nutrient/electricity (Norsker et al., 2011; Richardson et al., 2012). As the design of the cultivation methods differ, the costs also differs between them (Norsker et al., 2011; Richardson et al., 2012). The processing route costs will also differ a little, depending on the amount of biomass processed and thus the energy and nutrients needed for fractionation and processing (Richardson et al., 2012; Ruiz et al., 2016).

A product with a positive NPV is considered an attractive investment (Blok, 2006). However, the NPV is only an absolute value and does not give a clear indication into the profitability of a project compared to the costs (Blok, 2006). Therefore, the pay-back period (PBP) is used for a cost benefit analysis (Blok, 2006). The formula used is (Blok, 2006):

$$PBP = \left(\frac{I}{B - C}\right)$$

I= Initial investment
B= Annual benefits
C= Annual costs

An average payback period for large production facilities and or mining projects is between the 10 to 20 years (Leanmanufacture.net, 2009).

The goal of this research is to indicate whether it is commercially feasible to create HEFA from microalgae. Therefore, the revenue created by selling the full biomass is important. The production of the algae biomass per cultivation system is calculated by the following formula (Norsker et al., 2011):

$$PE\ (\%solar) = \frac{P_a \cdot h}{I} \cdot 100$$

PE = Photosynthetic efficiency (in %)
P_a = Areal dry weight (DW) productivity (in g/(day*m^2), ash free or total DW)
h = Combustion enthalpy (in MJ/g total or ash free DW)
I = Intercepted solar irradiation (MJ/(day*m^2))

Here, the enthalpy is specific to the algae strain selected. The photosynthetic efficiency is based on the solar radiation and productivity of the cultivation system in other regions. The intercepted solar irradiation value is based on the monthly irradiation values of the Netherlands. Hence, the productivity can be calculated.

3.0 Method

The theory indicates the steps of the algae to jet fuel cycle. It indicates the four cultivation systems, the processing route and its considerations. From the theory, four different algal to jet fuel cycles can be formed. An overview of these routes is given below (figure 2).

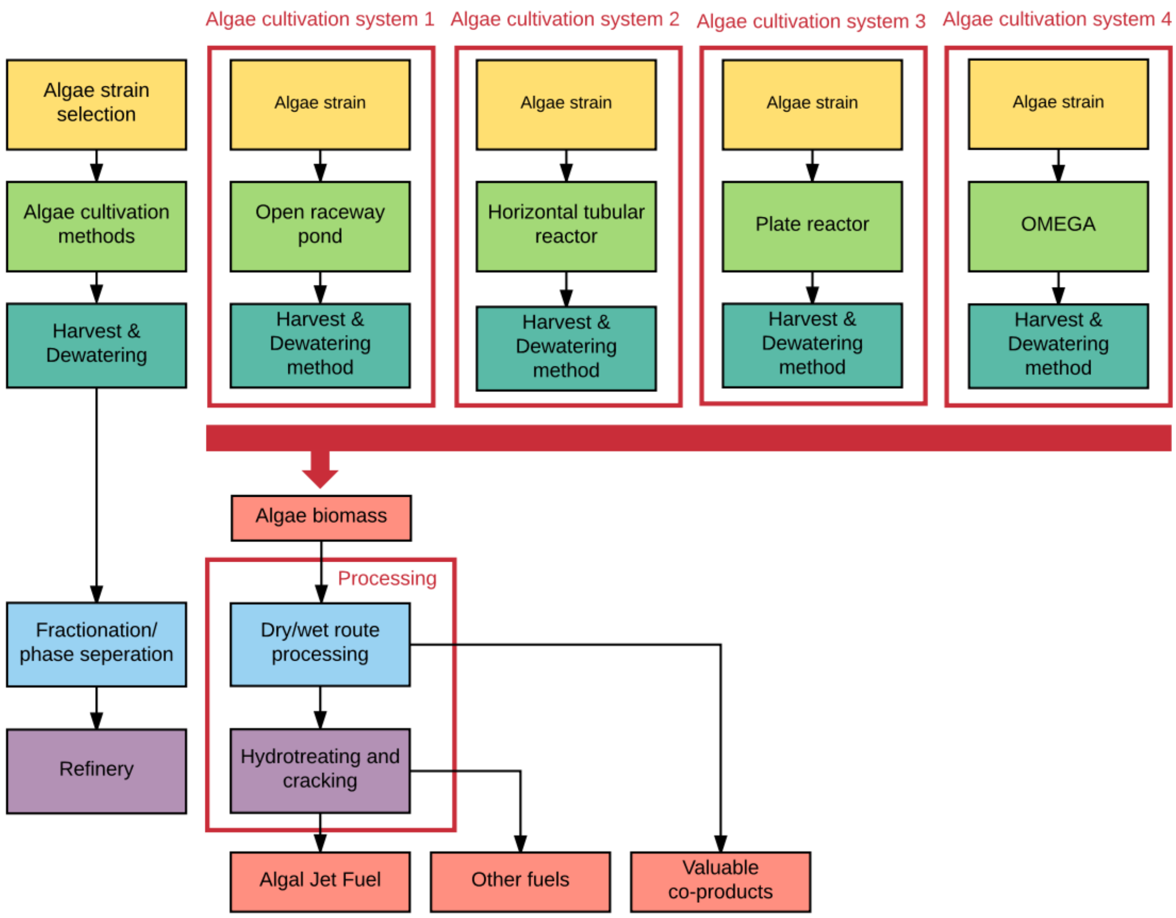


Figure 2: Overview of the 4 algae biomass to AJF cycles

Figure 2 gives a general overview of the different aspects of the fuel cycle. However, the exact technical design of each step is still unclear. In order to form a base case for the four algae cultivation methods and an answer to the main research question, some more information is needed about scale of the research, the exact cultivation system design and costs, the algae biomass processing design and costs, the final costs and revenues and the current and future commercial feasibility. To address these aspects, the steps below are followed.

3.1 General considerations and preliminary assessment

The first step of this research is to investigate the scale of the research. In this step, the algae strain is selected based on the considerations given in the theory section above, such as oil content, production rate and whether it is able to survive in the Dutch environment. Both a freshwater algae strain and a marine algae strain are selected to compare the both of them. It is thus assessed whether the marine and fresh water microalgae differ significantly in their potential. The same algae strain will be used for each of the cultivation methods to ensure a fair comparison between the methods. Therefore, this consideration is discussed in the preliminary assessment rather than the cultivation system as mentioned by the theory. Furthermore, the specific productivity of the algae in the Netherlands is calculated for each of the four cultivation methods. For this calculation the photosynthetic efficiency of each of the cultivation methods and the monthly irradiance data of the Netherlands are used, obtained from a source based on KNMI data (Allesoverzonnepanelen.nl, 2012). Finally, the total land occupation for each cultivation method is assessed. This is based on the 1 Mton CO_2 target of Tata Steel and the productivity, as the productivity ensures the CO_2 mitigation during photosynthesis. This CO_2 originates from flue gas from Tata Steel, which consists of roughly 20% CO_2, 70% Nitrogen and 10% other gases, which could be fully used for the algae cultivation (Louwerse & Koelemeijer, 2016). By estimating the exact CO_2 uptake efficiency of each of the different cultivation methods, the exact size of the plant can be identified.

The size and productivity of each of the cultivation methods is given with a margin which is based on an optimistic scenario and a pessimistic scenario. The optimistic scenario takes only optimistic production values into account and the pessimistic only pessimistic production values. This range is used as this research topic is not yet widely performed on a large scale and therefore each value has a high uncertainty.

3.2 Cultivation cost assessment

After the general considerations and preliminary assessment, the cultivation system for each of the four cultivation methods is described in more detail. Literature indicates different cultivation designs per method. However, only one design per method is considered in this research, due to time constraints. The design will be selected based on: costs, large scale viability (commercial readiness) and the amount of data available, as specific data are needed for the cost calculations.

As mentioned by the theory, harvesting and dewatering are also part of the cultivation system and therefore these parts are also described and modelled in this part of the research. To ensure a fair comparison between the different cultivation methods, the same harvesting and dewatering system is used. The selection of the harvesting/dewatering technique is based on literature and an interview with Mr. Jongbloed, who runs a commercial algae cultivation company in the Netherlands (Jongbloed, R., personal communication, 22 March, 2017)[2]. Again the costs, large scale viability and data availability play a key role in this process. The cultivation system describes a path from nutrients and microalgae medium preparation to a 15 to 20% dry based algae biomass, which is then processed in the processing part of this research. The most important part of this research step is that the technical design is described so that it is clear which data are needed for the next step of the research. This design is used for the cost calculation of the algae biomass per cultivation method. The design assumptions can be found in appendix B.

The algae biomass production costs are calculated by using the optimistic and pessimistic scenario of the general growth dimensions and the technical description of the cultivation systems section. The optimistic and pessimistic scenario are further elaborated with optimistic and pessimistic values for the dewatering and harvesting section, as these values can vary a lot (Brennan & Owende, 2010; Rhea, 2016). The assumptions for the calculation are obtained from literature or by personal communication with specialists. All the data is integrated in an excel model which is added as supplementary data to this research.

The costs are divided into major equipment costs, operating costs and other fixed costs. All the costs are given on an annual basis by multiplication with the annuity factor whenever needed. The total annual costs are added together and then divided by the annually produced biomass to obtain the

[2] Jongbloed, R. is a director at Algaspring, a commercial micro algae cultivation company situated in the Netherlands

costs estimate per kg dry weight. Thereafter, the costs are allocated to check which aspect mainly influences the costs of the biomass. Furthermore, the costs per different method are compared to find out if there is a difference in cost division for each of the cultivation methods.

3.3 Processing cost assessment

The 15% to 20% dry biomass that is obtained from the cultivation system are processed in order to obtain valuable products that can create revenue. Algae jet fuel is the most important product in this research, as this is the fuel that is used at Schiphol. The algae biomass processing research section consists of two main steps, the algae biomass composition and process design (including fraction extraction and algae jet fuel refining). The algae biomass composition gives the general composition of microalgae, which is needed to gain insight in how much of the algae biomass can be used for which market. In this research, advanced fractionation and refinery is used to fully use the algae mass for viable products. The process design gives a detailed overview of the different steps of the microalgae biomass processing from 15% to 20% dry algae biomass to the final products. The fractionation of algae biomass is a very complex and costly process (Ruiz et al., 2016).

The complete biorefinery design (complete fractionation) presented in Ruiz et al., 2016 is used in this research, as this design is backed with a lot of data and ensures the full use of the algae (Ruiz et al., 2016). The design is explained in more detail in section 7.0. For the refinery of the extracted lipid fraction, the approved HEFA processing route is used. The refinery process is discussed in more detail in section 7.2.2 of the research. The processing design assumptions are given in appendix D.

The costs for the processing are assessed in the same way as the cultivation. The costs are divided into major equipment costs, operating costs and other fixed costs. For algae oil refining, a fixed upgrading costs for kerosene production is used (M. Pearlson, Wollersheim, & Hileman, 2013). These fixed upgrading costs are used, as the factories that can perform these processes are already installed and available (AltAir Fuels, 2016) and therefore do not need to be constructed. These costs are added and divided by the total amount (in kg) of valuable fractions obtained. Again, these calculations are performed in the supplementary excel model, making use of the described assumptions. Finally, the main cost drivers and the difference in processing costs between the different cultivation systems are assessed.

3.4 Current and future commercial feasibility

In this section of the research the total costs and revenues from algae biomass cultivation and processing are calculated. The total processing costs are added to the biomass cultivation costs to obtain a final product cost. The total revenue is calculated by multiplying the total obtained biomass fractions with the price (€/ton) of each fraction. For this calculation, the total market size is taken into account, as the market for some of the fractions could be saturated. Furthermore, the revenue created by CO_2 mitigation is considered. All these calculations are performed in the supplementary excel file.

In order to assess the commercial feasibility of each of the different cultivation methods the NPV is used, as mentioned in the theory. All the values obtained from the other sub-questions are used and filled into the NPV formula. Furthermore, the PBPs are calculated. The NPV and PBP together can indicate whether it is currently commercially feasible to invest in the microalgae for the production of AJF while taking valuable co-products into consideration.

The last step of this research is to explore the future potential of AJF. This is assessed by looking at the current commercial feasibility and examine whether future potential improvements could increase commercial feasibility. Examples of these alternatives are HTL, improvement in harvesting methods and improvement in algae strains (Chen et al., 2011; Chiaramonti et al., 2017; Hannon et al., 2010).

Furthermore, it is assessed how much the price of the AJF and the CO2 permits have to rise until each project is commercially feasible and whether this rise is likely to occur.

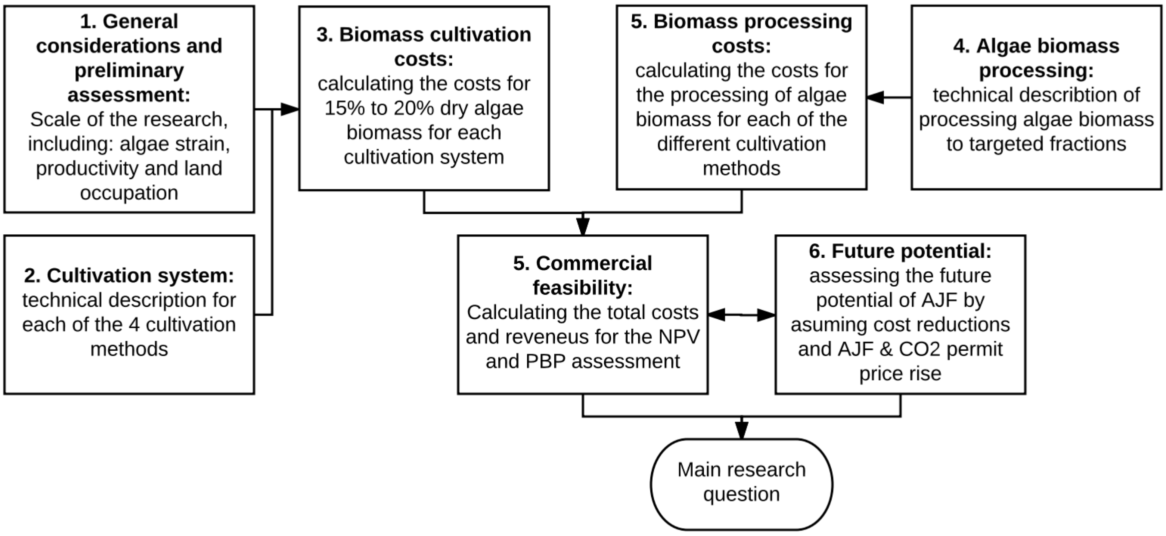


Figure 3: Overview of the research method

An overview of all the different research steps is given in figure 3. The figure indicates how the main research question is answered. For each of the design steps, literature and/or personal communication is used as indicated by the method.

4.0 General considerations and preliminary assessment

Before the four cultivation methods are investigated, a preliminary assessment of the scale of the research is performed, as this gives an insight in the design and scale of the cultivation systems. The aspects treated in this section are the algae strain, productivity and the land occupation.

4.1 Algae strain

As mentioned before, the algae strain selection in this research mainly depends on the biomass productivity and the lipid content (Rodolfi et al., 2009). However, especially for the open pond systems a strong algae strain is needed that can survive contamination of other algae species and microorganisms (Yen, Hu, Chen, & Chang, 2014). An overview of the characteristics of some of the most used algae strains is given in the table below.

Table 1:Lipid content and productivity of microalgae species (Mata et al., 2010)

Marine and freshwater microalgae species	Lipid content (% dry weight biomass)	Lipid productivity (mg/L/day)	Volumetric productivity of biomass (g/L/day)	Areal productivity of biomass (g/m^2/day)
Ankistrodesmus sp.	24.0–31.0	–	–	11.5–17.4
Botryococcus braunii	25.0–75.0	–	0.02	3.0
Chaetoceros muelleri	33.6	21.8	0.07	–
Chaetoceros calcitrans	14.6–16.4/39.8	17.6	0.04	–
Chlorella emersonii	25.0–63.0	10.3–50.0	0.036–0.041	0.91–0.97
Chlorella protothecoides	14.6–57.8	1214	2.00–7.70	–
Chlorella sorokiniana	19.0–22.0	44.7	0.23–1.47	–
Chlorella vulgaris	5.0–58.0	11.2–40.0	0.02–0.20	0.57–0.95
Chlorella sp.	10.0–48.0	42.1	0.02–2.5	1.61–16.47/25
Chlorella pyrenoidosa	2.0	–	2.90–3.64	72.5/130
Chlorella	18.0–57.0	18.7	–	3.50–13.90
Chlorococcum sp.	19.3	53.7	0.28	–
Crypthecodinium cohnii	20.0–51.1	–	10	–
Dunaliella salina	6.0–25.0	116.0	0.22–0.34	1.6–3.5/20–38
Dunaliella primolecta	23.1	–	0.09	14
Dunaliella tertiolecta	16.7–71.0	–	0.12	–
Dunaliella sp.	17.5–67.0	33.5	–	–
Ellipsoidion sp.	27.4	47.3	0.17	–
Euglena gracilis	14.0–20.0	–	7.70	–
Haematococcus pluvialis	25.0	–	0.05–0.06	10.2–36.4
Isochrysis galbana	7.0–40.0	–	0.32–1.60	–
Isochrysis sp.	7.1–33	37.8	0.08–0.17	–
Monodus subterraneus	16.0	30.4	0.19	–
Monallanthus salina	20.0–22.0	–	0.08	12
Nannochloris sp.	20.0–56.0	60.9–76.5	0.17–0.51	–
Nannochloropsis oculata.	22.7–29.7	84.0–142.0	0.37–0.48	–
Nannochloropsis sp.	12.0–53.0	37.6–90.0	0.17–1.43	1.9–5.3
Neochloris oleoabundans	29.0–65.0	90.0–134.0	–	–
Nitzschia sp.	16.0–47.0			8.8–21.6
Oocystis pusilla	10.5	–	–	40.6–45.8
Pavlova salina	30.9	49.4	0.16	–
Pavlova lutheri	35.5	40.2	0.14	–
Phaeodactylum tricornutum	18.0–57.0	44.8	0.003–1.9	2.4–21
Porphyridium cruentum	9.0–18.8/60.7	34.8	0.36–1.50	25
Scenedesmus obliquus	11.0–55.0	–	0.004–0.74	–
Scenedesmus quadricauda	1.9–18.4	35.1	0.19	–
Scenedesmus sp.	19.6–21.1	40.8–53.9	0.03–0.26	2.43–13.52
Skeletonema sp.	13.3–31.8	27.3	0.09	–
Skeletonema costatum	13.5–51.3	17.4	0.08	–
Spirulina platensis	4.0–16.6	–	0.06–4.3	1.5–14.5/24–51
Spirulina maxima	4.0–9.0	–	0.21–0.25	25
Thalassiosira pseudonana	20.6	17.4	0.08	–
Tetraselmis suecica	8.5–23.0	27.0–36.4	0.12–0.32	19
Tetraselmis sp.	12.6–14.7	43.4	0.30	–

The table and literature indicates that Chlorella sp. and Nannochloropsis sp. are among the most used algae strains and have a high growth rate, lipid content and are able to survive in both in and outdoor areas (Hulst, 2012; Mata et al., 2010; Rodolfi et al., 2009). However, Chlorella is a freshwater and Nannochloropsis a marine algae species (Rodolfi et al., 2009). In a report by Hulst, it becomes clear that the productivity of Chlorella is larger than Nannochloropsis in the Netherlands (Hulst, 2012). However, this difference is rather small and hard to measure, as the productivity of each of the different reactor designs also influence the growth speed (Hulst, 2012). Furthermore, it becomes clear from table 1 that the productivity also has a large range of 0.02-2.5 g/L/day for Chlorella and 0.17-1.43 g/L/day for Nannochloropsis. However, the difference in productivity between the two algae species is not extremely large. Therefore, it is assumed that the productivity is the same for both Chlorella sp. and Nannochloropsis sp. in this research.

Table 1 indicates a lipid content of 10.0-48.0% for Chlorella sp. and 12.0-53.0% for Nannochloropsis. In this research, the lipid content of Chlorella and Nannochloropsis is assumed to be 20%, which is an average value for algae with a high lipid content in the Netherlands (Jongbloed, R., personal communication, 22 March, 2017). Besides, this percentage lies between the boundaries mentioned in table 1. Finally, literature indicates that stress on the algae lowers the growth speed but enlarges the oil accumulation (Chiaramonti et al., 2017). As the growth conditions for algae are not optimal in the Netherlands (Slegers et al., 2013), a 20% oil content seems a fair assumption for both algae strains.

During an interview with Jongbloed it became apparent that there are still a lot of unknown algae species and that it is dependent on the final product which algae is best to grow (Jongbloed, R., personal communication, 22 March, 2017). Furthermore, it is hard to find out which algae is most suitable for a certain location, as there are so many aspects that influence the growth. Examples are the cultivation system itself, the temperature or the exact amount of pathogens in the air that can contaminate the algae batch (Jongbloed, R., personal communication, 22 March, 2017). Due to these uncertainties, the Nannochloropsis sp. and Chlorella sp. which can survive in open systems as mentioned before are used as an example in this research. More in depth research is needed to find out which algae strain can grow best in the Netherlands.

Multiple sources indicate that both marine and freshwater microalgae can be used for the production of biofuels and that salts are used for the algae fractionation (Brennan & Owende, 2010; Chiaramonti et al., 2017; Cox et al., 2014; Ruiz et al., 2016). Therefore, it is assumed that the salt content in the marine algae do not cause a problem for the processing of the algae. Thus, the same processing route is used for both marine and freshwater algae.

4.2 Algae productivity

As mentioned in the theory, the algae growth mainly depends on the photosynthetic efficiency (of the cultivation system), the location specific irradiation and the combustion enthalpy (the amount of energy an algae strain takes up) (Norsker et al., 2011).

Ph and nutrient supply also influence algae growth (S. P. Singh & Singh, 2015). However, these aspects are considered optimal in this research by sufficient nutrient supply and degassing. The last aspect which influences growth is temperature (Malakootian, Hatami, Dowlatshahi, & Rajabizadeh, 2016; S. P. Singh & Singh, 2015), as mentioned by the theory. Temperature influences the growth, as it indicates an optimal range in which the algae can grow (Baicha et al., 2016; S. P. Singh & Singh, 2015). Outside of this range the algae can be damaged or even destroyed (Baicha et al., 2016; S. P. Singh & Singh, 2015). In general a lower temperature causes a lower growth rate, depending on the optimal temperature (Hulst, 2012). The exact influence of temperature on algae growth is hard to measure; hence only the absolute temperature boundaries are taken into account in this research. When looking at the temperature, chlorella can grow at environment temperatures of 5 to 30 degrees Celsius, with an optimal temperature of 25 degrees (S. P. Singh & Singh, 2015). The absolute minimum temperature at which Nannochloropsis sp. can be grown is 5 degrees Celsius, which is assumed to be the same as Nannochloropsis Oculata (Malakootian et al., 2016). In general the optimum temperature for microalgae to grow is between 20-30 degrees Celsius (Chisti, 2008).

For the calculation of the growth rate, the average temperature (both on and offshore) and irradiation, data of the Netherlands are used. These values are given in table 2-4.

Table 2: Average solar irradiance at Schiphol (Allesoverzonnepanelen.nl, 2012)

Average solar irradiance data Schiphol (in KWH/(m2*yr.))													
Month	**Jan**	**Feb**	**Mar**	**Apr**	**may**	**Jun**	**Jul**	**Aug**	**Sep**	**Oct**	**Nov**	**Dec**	**Unit**
Value	20	36	74	119	158	159	158	132	85	51	23	15	KWH/(m^2*yr.)

Table 3: Average onshore temperature in the Netherlands (KNMI, 2017)

Average onshore temperature Netherlands (in degrees centigrade)													
Month	**Jan**	**Feb**	**Mar**	**Apr**	**may**	**Jun**	**Jul**	**Aug**	**Sep**	**Oct**	**Nov**	**Dec**	**Unit**
Value	3.1	3.3	6.2	9.2	13.1	15.6	17.9	17.5	14.5	10.7	6.7	3.7	˚C

Table 4: Average offshore temperature North Sea (Gemiddeldgezien.nl, 2017)

Average temperature North Sea offshore (in degrees centigrade)													
Month	**Jan**	**Feb**	**Mar**	**Apr**	**may**	**Jun**	**Jul**	**Aug**	**Sep**	**Oct**	**Nov**	**Dec**	**Unit**
Value	5.5	4.5	5.7	7.3	11.7	14.8	17.5	20	19	17.2	13	8.8	˚C

The months that are highlighted in red are too cold for the microalgae to grow and are not considered in the calculation. Therefore, the onshore growth season is 275 days/yr and offshore 337 days/yr. Furthermore, the average solar irradiation is calculated by adding all the solar irradiation data of the months that have the sufficiently high temperature and then divide this number by the total amount of days in the growth season. Afterwards, this amount is multiplied by 3.6 to receive a value in MJ/(m^2*day). The average onshore solar irradiance is 12.6 MJ/(m^2*day) and offshore 10.6 MJ/(m^2*day), for the total growth season.

Furthermore, the enthalpy and PE of the raceway open pond, tubular and plate reactor are (Norsker et al., 2011):

- Enthalpy of combustion 0,022 MJ/(G(DW)) (Dillschneider, Steinweg, Rosello-Sastre, & Posten, 2013; Ruiz et al., 2016; Weyer, Bush, Darzins, & Willson, 2009)
- PE raceway open pond: 1.5%
- PE horizontal tubular reactor: 3%
- PE plate reactor: 5%

The PE of each of the different cultivation methods includes shading and orientation (Norsker et al., 2011). Therefore, these aspects are also taken into account in this research.

When comparing these values to the values obtained by the ALgaePARC facility, these PE values seem a little high (Ruiz et al., 2016). Therefore, the values from Ruiz et al. (2016) are used as a pessimistic scenario. The values given in this paper are (Ruiz et al., 2016):

- PE raceway open pond: 1.2%
- PE horizontal tubular reactor: 1.5%
- PE plate reactor: 2.7%

The PE of the omega cultivation method is calculated by using the PE formula and filling in the production data of May and April of the San Francisco region obtained from the Trent (2013) paper and linking this to the irradiation data of this region (NREL, 2016). The average production rate is 14.1 (g/m^2*day) (Trent, 2013) and irradiation data is 5.75 Kwh/(m^2*day) (NREL, 2016), resulting in an PE of 1,4%. The full calculation can be found in the supplementary excel file. As, there are no other data available for this method, the OMEGA method will have the same PE for both the pessimistic and optimistic scenario.

Using all the data and filling it into formula 1 gives the following daily production data per annual growth season:

Table 5: Daily dry weight production data of the different algae cultivation methods in the Netherlands

Method	Optimistic	Pessimistic	Unit
Productivity raceway pond	23.5	18.8	Ton/ha/yr
Productivity tubular reactor	47.1	23.5	Ton/ha/yr
Productivity plate reactor	78.5	42.4	Ton/ha/yr
Productivity OMEGA	23.3	23.3	Ton/ha/yr

These optimistic values differ a little when looking at the production data indicated in the paper of Norsker et al. (table 6 below) and Zittelli et al. (Zittelli, Biondi, Rodolfi, & Tredici, 2013). The difference in value is probably caused by the difference in irradiation data between the papers.

Table 6:Microalgae biomass production per algae cultivation method in the Netherlands (Norsker et al., 2011)

	Open pond	Tubular bioreactor	Flat panel bioreactor
PE (solar)	1.5%	3%	5%
Productivity (ton DW) per ha)	21	41	64

The pessimistic scenario indicates a significant lower productivity than the values given in table 6. This is probably the case because lab studies generally have higher efficiencies than the systems that are positioned outside, as is the case for the AlgaePARC (Ruiz et al., 2016). However, these production values lay within the margin given (Table 5).

4.3 Land occupation

The amount of occupied land is calculated by the amount of biomass needed for the remediation of 1Mton of CO_2. To find out how much CO_2 is taken up per amount of algae (CO_2 ratio), the composition of the microalgae has to be investigated. This can be done by looking at the biochemical composition of the biomass and the molar fractions of the substrates and products, resulting in the following reaction (Hulst, 2012):

$$CO_2 + 0.148HNO_3 + 0.014H_2SO_4 + 0.012H_3PO_4 + 0.751H_2O \rightarrow CH_{1.715}O_{0.427}N_{0.148}S_{0.014}P_{0.012} + 1.437\ O_2$$

And the following table:

Table 7: Required amount of nutrients per unit of algae (Hulst, 2012)

Name	Formula	Molecular weight (g/mole)	Ratio with algae
Algae	$CH_{1.715}O_{0.427}N_{0.148}S_{0.014}P_{0.012}$	23.499	1.00
Carbon dioxide	CO_2	44.009	1.87
Oxygen	O_2	31.999	1.36
Nitrate	HNO_3	63.013	2.68
Ammonia	NH_3	17.031	0.72
Urea	$CO(NH_2)_2$	60.056	2.56
Phosphate	H_2PO_4	96.988	4.13

From this table, it becomes clear that for every gram of algae biomass 1.87 grams of CO_2 are needed. Furthermore, the CO_2 uptake efficiency of each of the different cultivation methods is considered. The uptake efficiencies are 60% for the tubular and plate reactor (Sobczuk et al., 2000; Zhang, 2015), 40% for raceway open pond as the respiration rate of the basin is very high (Zhang, 2015) and 50% for the OMEGA method (Trent, 2013).

When combining the daily production rates, days per growth season, CO_2 ratio and CO_2 uptake efficiency, the total amount of land needed per cultivation method can be calculated. This calculation gives the total amount of irradiated land. The total amount of land needed is then assumed to be 1.25 times the irradiated area (Norsker et al., 2011). For the offshore an irradiated area efficiency of 70% is assumed (Trent, 2013) which translates into a total area of 1.43 times the irradiated area. Again, the optimistic and pessimistic productivity scenarios are used to calculate the amount of land needed. The final values are given below.

Table 8: Total amount of land occupation for each cultivation method

Total amount of land needed per cultivation method					
	Optimistic		Pessimistic		
	Irradiated	Total land	Irradiated	Total land	Unit
Raceway open pond	90.9	113.6	113.6	142.0	Km^2
Tubular reactor	68.2	85.2	136.3	170.4	Km^2
Plate reactor	40.9	51.1	75.7	94.7	Km^2
OMEGA	140.8	201.2	140.8	201.2	Km^2

When looking at the amount of land needed, it becomes clear that the onshore possibilities are unlikely to be used on the 1 million ton CO_2 mitigation scale, because the amount of land available in the Netherlands is limited (Slaa, 2014). However, they are taken into account for the cost calculation as it is interesting to see what the cost difference will be between the onshore and offshore cultivation.

5.0 Cultivation

In the next step, the general cultivation systems are described. The cultivation system is build out of the cultivation methods, harvesting and dewatering of the algae. In this research, the cultivation systems contain: a medium filter unit, medium preparation tank, medium feed pump, algae cultivation method, a sedimentation tank, a centrifuge for dewatering and a weighing station. The general cultivation system design used in this paper can be seen in figure 4. This system design is chosen as it allows for a constant and continues cultivation medium (Critical Process Filtration Inc., n.d.) and large scale algae operations (Brennan & Owende, 2010). The overview is a simplified version of the total cultivation system, but the main aspects are taken into account. In this research, an optimal medium composition is assumed in the algae cultivation method, as it is unclear to what extent inequalities such as an O_2 surplus could influence the growth of the microalgae. Furthermore, costs are taken into account to maximize the medium composition. Finally, the flue gas pipeline and pumping energy for both the onshore and offshore are also included in the process.

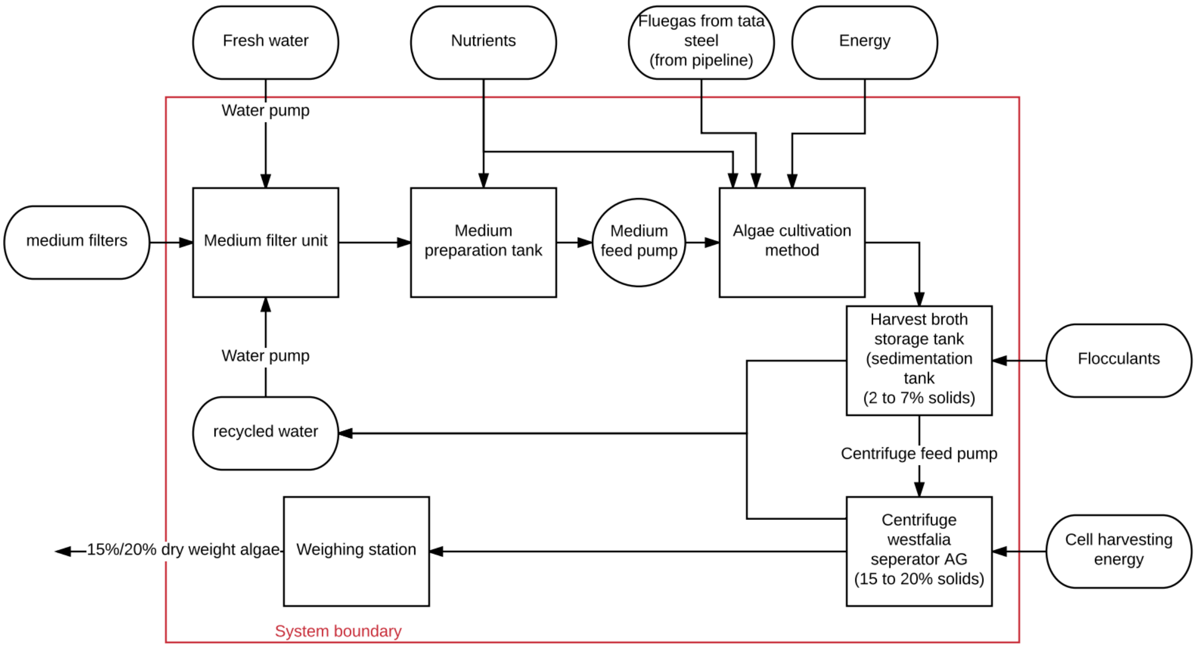


Figure 4: Overview cultivation system design including harvesting and dewatering based on (Critical Process Filtration Inc., n.d.; Norsker et al., 2011; Wijffels, n.d.) *and Jongbloed, R., personal communication, 22 March, 2017).*

As an extra validation of the large-scale algae cultivation system, an interview with René Jongbloed (Director at Algaspring) was performed. The different aspects of the algae cultivation system are mentioned below.

5.1 Medium filtration and preparation

The first step of an algae cultivation system is the medium preparation filter and tank. The medium preparation filter is used to obtain a sterile medium that does not contain pathogens and therefore ensures optimal growth (Critical Process Filtration Inc., n.d.). The filter unit that is used in this research is membrane filtration, which is also used in Norsker et al. (2011). After the filtration unit, the sterile medium is inserted into a medium preparation tank where the clean nutrients and the algae are inserted and mixed. For this research the cheapest water tank is used, one made from concrete D-115 (see figure 5) (Rettew, n.d.). The size of the tank is based on a residence time of 12 hours as algae can double their biomass in 24 hours (Chisti, 2007) and do not grow during night time. After the preparation tank, the premix culture is inserted in the cultivation method.

Figure 5: D-115 concrete premix tank (Rettew, n.d.)

5.2 Cultivation method

As mentioned before, in this research four different cultivation methods are considered: raceway open pond, tubular reactor, plate reactor and the OMEGA. Each of these systems has a specific design. The design that is used in this research is given below. These designs are selected as information of the different aspects of the design is available and already used.

5.2.1 Raceway open pond

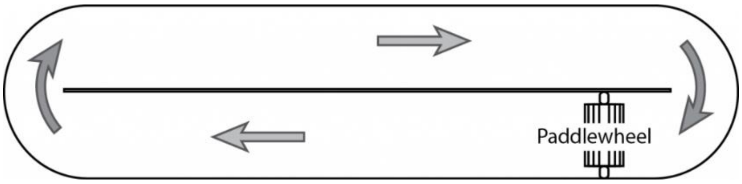


Figure 6: Raceway open pond design

In this paper the raceway open pond design is 20 cm deep and built from clay covered with a PVC layer (Norsker et al., 2011). Furthermore, it is a single circuit design (as is seen in figure 6) which is 10m wide and 100 metres long (Norsker et al., 2011). From other papers it becomes apparent that this is a cheap option and enough data is available for this cultivation design to make a cost estimate (Belay, 2013; Chisti, 2007; Darzins, Pienkos, & Edye, 2010; Norsker et al., 2011). Finally, it is assumed that the cultivation system works for 24h/day. The 24/7 strategy is chosen as the operating systems maintain an optimal medium, so that the algae can flourish and do not die. In this design the peddle wheel is used, which ensures a sufficient concentration and mixing of nutrients in the ponds (Davis et al., 2016). Furthermore, it is assumed that the biomass only grows 12 hours a day, during daytime. The algae concentration in the open pond is assumed to be 0.32 kg m^{-3} (Norsker et al., 2011).

The algae culture is continually fed in front of the paddlewheel and harvested behind the wheel after a circulation loop is fulfilled (Chisti, 2007). The raceway system is used since the 1950s (Chisti, 2007). Therefore, extensive experience exist about how to engineer and operate these systems (Chisti, 2007).

Raceway open ponds are less expensive then photobioreactors, because the ponds are made from less expensive materials and require less energy for mixing (Jorquera et al., 2010). However, the open ponds need a relative large area of land, have low efficiency and light utilization due to inefficient mixing, poor conditions to uptake CO_2, almost no temperature control (temperature is maintained only be evaporation), high contamination risks and a concentration of the algae which remains small (Chisti, 2007; Jorquera et al., 2010). Therefore, not all algae species can survive in a not fully controlled open system (Jorquera et al., 2010). For the calculation of the production costs per dry weight a lot more assumptions are made, which are used in an excel model. The assumptions are given in appendix B.

5.2.2 Tubular reactor

The tubular photobioreactor design that is considered in this research is a single horizontal layer photobioreactor made from cheap polyethylene film with a one year lifetime (Norsker et al., 2011). The single layer photobioreactor is chosen because the productivity of a stacked tubular reactor is only slightly better in the Netherlands (Hulst, 2012) and a data set is available which uses the single layer photobioreactor (Norsker et al., 2011). An overview of such a system, including a degasser, is shown in figure 7. The degasser is used to blow off the accumulated oxygen that is present in the system. The cooling and heating of the system is not considered in this research as it is hard to predict what the exact temperature inside the reactor will be. Moreover, the absolute temperature is already used for the assessment of the exact length of the growth season. Furthermore, a research performed by Girdhari, indicates a neglectable amount of energy needed for the cooling of a tubular reactor system in the Netherlands (Girdhari, 2011).

The diameter of the tubular reactor tube is assumed to be 0.057 m with a flow rate 4.59 $m^3\ h^{-1}$ (Norsker et al., 2011). The biomass concentration obtained in this system is 1.7 kg m^{-3} (Norsker et al., 2011). The circulation of the culture is performed by a circulation pump. The assumptions made for this system are given in appendix B.

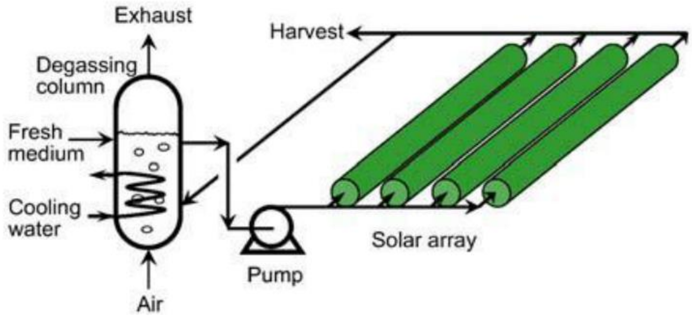


Figure 7: Overview single layer horizontal tubular reactor (Chisti, 2007)

In general, tubular reactors are made of transparent glass or plastic tubes that are aligned horizontally or vertically (some are even inclined or in a helix form) (Brennan & Owende, 2010; Jorquera et al., 2010). The tubes are around 0.1 m in diameter and circulated by a mechanical pump or airlift system (Brennan & Owende, 2010). In this case a circulation pump is used. The diameter of the solar receiver tubes is selected so that the dark zone (i.e. the area with light intensity below saturation in the middle of the tube) is as small as possible and that the fluid is evenly mixed between the dark and light zone (Molina et al., 2001). This system is dependent on the circulation velocity of the culture and the O_2 removal, as this influences the growth rate (Molina et al., 2001). The challenge of this system is that the geometry of the solar receiver needs to maximize the capture of the incoming solar radiation while minimizing the land area used (Molina et al., 2001). In the solar receiver part of the system the photosynthesis occurs. The O_2 that is produced by this process is stripped by air in the airlift zone (Molina et al., 2001). Furthermore, the airlift system is used to exchange the supplied CO_2 throughout the medium (Eriksen, 2008). Finally, a gas-liquid separator prevents air from going into the solar receiver (Molina et al., 2001).

5.2.3 Plate reactor

The plate reactor is a flat design reactor with a depth of 3 cm. The plates have a vertical surface area of around 38.500 m3 ha^{-1} (both sides) and are 1.5 metres high (Norsker et al., 2011). The reactors are made from cheap polyethylene, with an one year lifetime which is supported by a steel mesh casing (Norsker et al., 2011). The circulation of the system is performed by air sparging from blowers. The biomass concentration obtained is assumed to be 2.1 kg/m^3 (Norsker et al., 2011). An overview of such a system is given in figure 8. The amount of energy needed for cooling a 3 cm wide plate reactor in the Netherlands is taken into account, which is approximately 1 KWH/kg(DW) algae biomass (Girdhari, 2011). The assumptions made for the calculation of this system are given in appendix B.

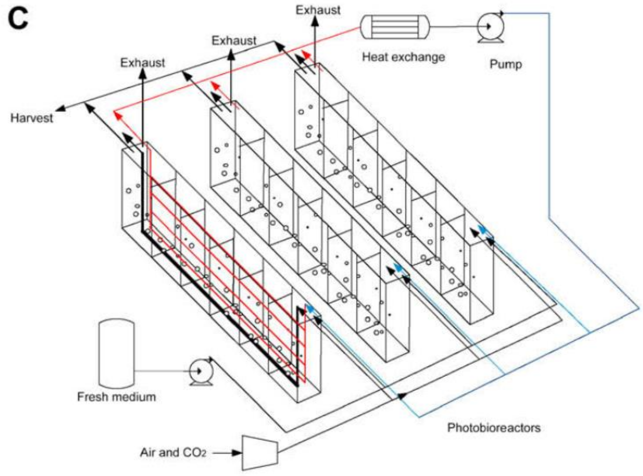


Figure 8: Overview plate reactor design (Jorquera et al., 2010)

In general, flat plate reactors can both be made of plastic or glass, and can be aligned horizontally or vertically (Sierra et al., 2008). The development of the plate reactors started in the 1950s and many system designs exist (Cheng-Wu et al., 2001; Sierra et al., 2008). However, there is a strong preference for east/west facing vertical deployment of plate reactors, as this reduces the land needed and increases the disperse radiation which is more favourable for algae systems instead of direct radiation (Sierra et al., 2008), this design is therefore also used in this research. Flat plate reactors are potentially sensitive to aeration stress problems (Sierra et al., 2008). However, the path that the dissolved O_2 takes through the medium is very short. This system therefore prevents O_2 build-up, which is often a problem with other closed systems (Cheng-Wu et al., 2001). Besides, the plate reactors can easily be cleaned as the plates are easy accessible (Cheng-Wu et al., 2001). Essentially flat plate reactors are the same as bubble columns, mixed efficiently by a flow of compressed air (Cheng-Wu et al., 2001).

5.2.4 OMEGA

OMEGA stands for: offshore membrane enclosures for growing algae (Trent, 2013). This is the only offshore microalgae cultivation system that allows for CO2 aeration (Trent, 2013). However, this system is still in the pilot phase (Trent, 2013). Therefore, only relatively small systems are currently tested in seawater tanks (Wiley et al., 2013). In general, the system is made from inexpensive, flexible plastic (LLDPE) (Trent, 2013). The system is build out of: tubular PBRs, Gas Exchange and Harvesting Column (GEHC) and a control box with sensors (Trent, 2013). The microalgae grow in the tubular PBRs that are situated in the seawater and are mixed by a circulation pump and wave energy (Trent, 2013). Furthermore, swirl vanes in the tubes are often used to enhance the mixing (Trent, 2013). The GEHC part manages the dissolved oxygen content by stripping the oxygen at the top of the column (Trent, 2013). Furthermore, CO2 is delivered at the bottom of the GEHC (Trent, 2013). At the bottom of the GEHC a pinch valve is situated to maintain a constant liquid level (Trent, 2013). This constant liquid level is needed to sediment the algae and harvest them through a harvest valve, see figure 9 (Trent, 2013). Finally, the control box contains sensors that try to maintain the optimal level for aspects such as PH, temperature and liquid level (Trent, 2013). Proposed is to use wastewater in this system, as it lowers the costs for nutrients (Carney et al., 2014). An overview of the system is given in figure 9.

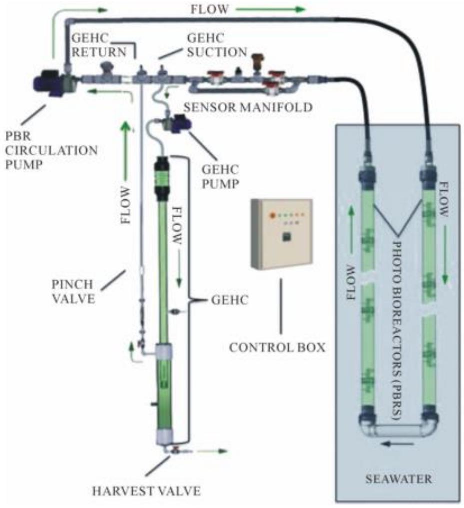


Figure 9: Overview test phase OMEGA method (Trent, 2013)

However, for the 1Mton CO_2 mitigation a large scale offshore algae cultivation plant is needed. Trent (2013) proposes a possible cheap cultivation plant, which is adopted in this research. The proposed design is made from plastic sheets that are welded into tubes, which improves the radiated surface area efficiency from 22 percent to 70 percent and reduces the cost significantly. By improving the radiated surface area efficiency, less area and thus less floating docks are needed. To reduce costs the swirl vanes are also left out. Furthermore, by this new design the flow rate and thus the energy needed for pumping is lowered. The design is an improvement of the design given in figure 10. The assumptions used are given in appendix B.

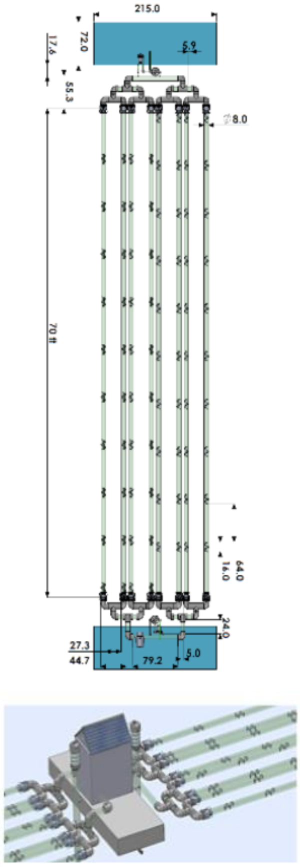


Figure 10: Overview experimental large-scale OMEGA system (Trent, 2013)

In this research only salt water algae are used in this system, as it is illogical to pump fresh water offshore. Therefore, only a CO_2 pipeline offshore and an algae broth pipeline back to shore is considered. Furthermore, it is assumed that the plant is situated 50 kilometres offshore as there are windmill parks situated in this range which could be used as anchor points.

5.3 Harvesting & dewatering

As mentioned in the theory the harvesting and dewatering is often done by bulk harvesting and thickening (Brennan & Owende, 2010). These two steps are generally used because the initial bulk harvesting improves the algae concentration and therefore ensures a lower energy use for the energy intensive thickening (Brennan & Owende, 2010). In this research sedimentation with flocculation is used as the bulk harvesting step, as this method is relatively fast and used for other large scale cases (Rhea, 2016). The sedimentation tank design is similar to the ones used for wastewater treatment, as shown in figure 11.

Flocculants are inserted in the influent pipe which causes sufficient mixing (Rhea, 2016). The algae sediment on the bottom of the sedimentation tank are moved to the middle by a underflow created by a slow moving rake mechanism (Rhea, 2016). Clear water should overflow into a weir and is then again sterilized and recycled to finally move back to the reactors (Rhea, 2016). In the sludge pipe a final algae concentration of about 2 to 7 weight ton% is obtained (Rhea, 2016).

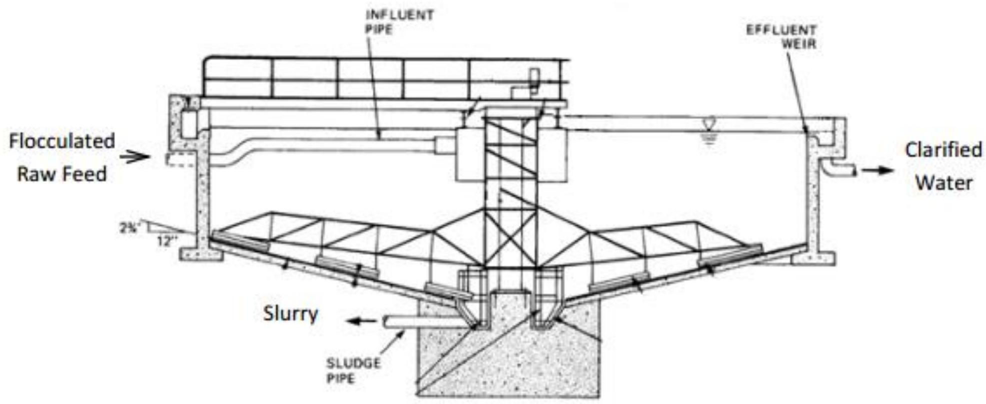


Figure 11: Gravity sedimentation thickener tank design (Rhea, 2016)

The flocculant used in this research is the biodegradable rice starch, as it is relatively cheap and is biodegradable compared to inorganic toxic flocculants (Rakesh, Saxena, Dhar, Prasanna, & Saxena, 2013). Only 120 mg/L is needed for the Chlorella culture, with a price of 0.7 USD/kg (Rakesh et al., 2013), which is 0.932*0.7= 0.65 euro/kg (value obtained 27-3-2017). The flocculants ensure a faster and more efficient sedimentation phase (Rhea, 2016). For easy comparison, the same is assumed for Nannochloropsis.

After the initial thickening, the algae broth is moved to a centrifuge. For this research, a Westfalia Separator AG centrifuge is used. This centrifuge can obtain a biomass concentration of roughly 15 to 20% (Brennan & Owende, 2010; Davis et al., 2016; Norsker et al., 2011). Furthermore, this centrifuge is specially used for substances with a high viscosity and enough technical data is available to calculate centrifugation costs in the next section of this research (Norsker et al., 2011). An image of the Westfalia Separator AG is given in figure 12.

Figure 12: Westfalia Separator AG (GEA, 2016)

6.0 Algae biomass production costs

After the technical design described in chapter 5, the costs for the production of 15% to 20% dry algae biomass are calculated. This calculation is performed in a self-made excel model. Excel is used as it allows for easy adjustment in the calculation. The excel model can be found as a supplementary document attached to this research. For the calculation of the production costs, the optimistic and pessimistic scenario are used. In addition to the optimistic and pessimistic values indicated in chapter 4.1.2, the values of table 9 are used. These values mainly focus on the dewatering and harvesting of biomass, as this will greatly influence the price of the biomass (Lammerink, R., personal communication, 16 March 2017)[3].

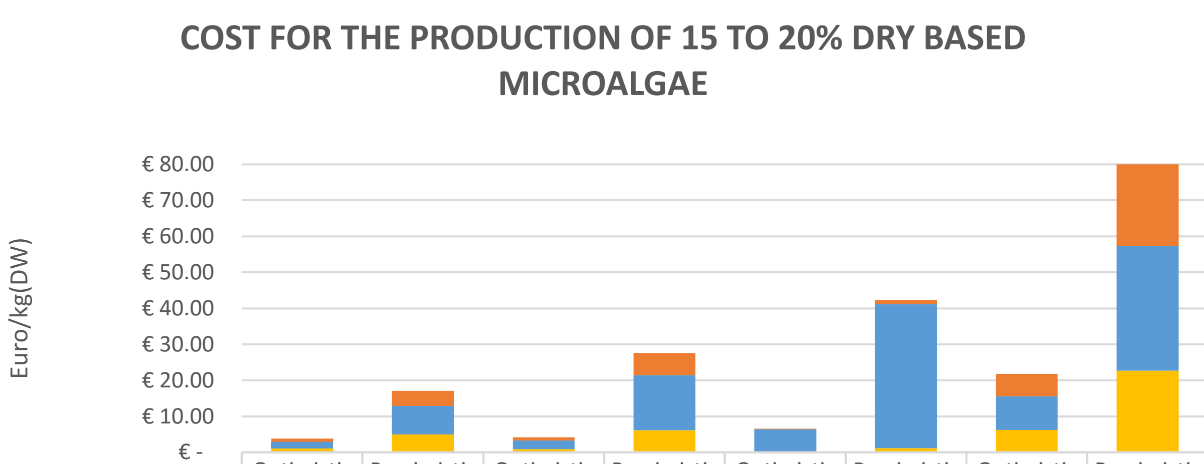
Table 9:Optimistic and pessimistic assumptions algae biomass dewatering and CO2 supply

Factor	Pessimistic	Optimistic	Unit	Source
Sedimentation efficiency	80	98	% (biomass obtained)	(Rhea, 2016)
Solid biomass concentration (after sedimentation)	2	7	% (solids in algae broth)	(Rhea, 2016)
Centrifugation efficiency	40	95	% (biomass obtained)	(Brennan & Owende, 2010)
Solid biomass concentration (after centrifugation)	15	20	% (solids in algae broth)	(Brennan & Owende, 2010)
Heat efficiency of the dryer	60	90	%	(Tang, Feng, & Shen, 2003)
Amount of biomass obtained that is sellable and clean without contamination	80	100	%	(Jongbloed, R., personal communication, 22 March, 2017)
CO_2 pipeline operation and maintenance	8	3	% of initial investment	(Noothout, Wiersma, Hurtado, Roelofsen, & Macdonald, 2014)

These different scenarios influence the costs per kg of dry biomass considerably, as the amount of water that needs to be subtracted and thus the energy used by the centrifuge and the dryer is influenced. Especially, the final amount of biomass differs between the optimistic and pessimistic scenario, which has a large influence on the final costs per unit of biomass. The detailed cost division and final cost price for each of the four cultivation methods and both the optimistic and pessimistic scenario are given in appendix C. An overview of the final cost division is shown in figure 13 below.

The costs are divided in the major equipment costs (MEC), operating costs and other fixed costs. The fixed costs an MEC are translated to costs per year by the annuity factor of 0.11, by assuming a discount rate of 10% and a lifetime of 25 years.

[3] Lammerink is a sales engineer at Flowserve. He is responsible for the water supply in the Dutch market, including public tenders.



	Raceway		Tubular		Plate		OMEGA	
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
Total	€ 3.84	€ 17.12	€ 4.21	€ 27.56	€ 6.61	€ 42.36	€ 21.84	€ 79.99
Other fixed costs	€ 0.90	€ 4.22	€ 0.91	€ 6.15	€ 0.19	€ 1.13	€ 6.23	€ 22.67
Operating costs	€ 1.88	€ 7.94	€ 2.39	€ 15.26	€ 6.24	€ 40.11	€ 9.37	€ 34.65
MEC	€ 1.06	€ 4.96	€ 0.91	€ 6.15	€ 0.19	€ 1.13	€ 6.23	€ 22.67

Scenario

Figure 13: Costs for the production of 15% to 20% dry algae biomass for each of the 4 cultivation methods

From this figure, it becomes apparent that the raceway pond is the cheapest and the OMEGA method the most expensive for both the optimistic and the pessimistic scenario. However, when the total land area occupation is taken into account, it is very unlikely that an area of approximately 135 km^2 onshore (which is needed for the raceway open pond) is available for the production of these microalgae in the Netherlands. Furthermore, the figure indicates a large cost range between the optimistic and pessimistic scenario. This is the case due to the smaller photosynthetic efficiencies in the pessimistic scenario (see section 4.2 algae productivity), which influence the biomass production and therefore the land area needed for the cultivation system. Furthermore, due to the assumptions in table 9, the final value of dry algae biomass obtained from the cultivation methods is influenced significantly, which causes higher costs per kg for the pessimistic scenario.

The cost division indicates that the operating costs are the highest for all the different cultivation methods and scenarios. Especially for the plate reactor the operating costs are high. These high costs are caused by the energy needed for the air blowers that power the medium circulation of the system, see cost division appendix C. These air blowers are less efficient then for example the circulation pump used in the tubular reactor. Further, the detailed cost division in Appendix C shows that the centrifugation costs for all the scenarios are remarkably low, as other papers indicate high costs for the cell harvesting (Brennan & Owende, 2010; Rhea, 2016). However, by including the sedimentation step, a lot of water is already lost before centrifugation which results in lower centrifugation cost/energy than for example in the Norsker et al. (2011) paper, which does not include this step.

For each of the cultivation methods, the aspects that have the highest influence on the costs are indicated in red in Appendix C. For the raceway pond the factors that influence the costs the most are: medium filters, culture medium premix, PVC liner, piping, buildings, flocculation and land occupation. For the Tubular reactor, the largest cost contributors are: the circulation pump power, the circulation pump itself, culture medium premix, the installation/piping and buildings, the medium filters, maintenance and the occupied land. For the plate reactor, the largest cost contributors are: by far the air blowers power as mentioned before, the culture medium premix, the medium filters, the occupied land, the polyethylene film for the reactors and the steel frames (used as suspension for the polyethylene film). Finally, for the OMEGA method the largest cost contributors are: by far the total bioreactor including GEHC pontoons etc., installation, piping, buildings, the PBR pump energy, instrumentation and control, maintenance, the sludge pipeline back to shore, CO_2 feed pipeline, labour and the medium filters.

The high costs for the air blower power in the plate reactor and the circulation pump power for the tubular reactor are explained by the fact that they are operational 24/7 and ensure mixing for the total bioreactor system. These, blowers and circulation pumps have an energy use which is a lot higher than the paddle wheel in the raceway pond system.

The OMEGA method has by far the highest costs (figure 13). The costs division (appendix C) indicates that these costs are mainly due to the investment costs of the offshore bioreactor and the offshore pipelines. The costs for the bioreactor are probably high as this method is still in the pilot phase and the method uses a floating infrastructure, which for example needs pontoons to float and expensive harvesting columns as a first sedimentation step to enlarge the concentration (Trent, 2013). However, for the cost calculation of the OMEGA method the conceptual commercial scale design proposed in Trent (2013). Therefore, the costs shown in this research take into account scaling and are therefore lower than the current test OMEGA reactor costs. However, this is still conceptual and therefore the exact cost reductions that could be obtained by research and development are uncertain.

The exact calculations of the different cost aspects can be found in the supplementary excel file. The success of the algae cultivation systems greatly depends on their business model and thus the targeted products. The processing of the biomass into products is treated in the next part of this research.

7.0 Algae biomass processing

After the dry algae mass is obtained, the biomass is processed. As mentioned in the introduction and the theory, the resulting product is HEFA. However, other valuable products can be obtained. This research aims to use the full potential of the biomass by advanced biorefinery, as this increases the value and thus the competitiveness of the microalgae (Ruiz et al., 2016).

7.1 Algae biomass composition

In order to find out what kind of products can be made from the dry algae, first the algae composition itself has to be investigated. The supplementary data from Ruiz et al. (2016) give insight into the composition of the Nannochloropsis sp.. When looking back at table 1, the lipid content fluctuates a lot for both Nannochloropsis sp. and Chlorella sp.. This occurs because the biomass composition is influenced by light intensity and frequency, nutrient level of the culture, temperature, feed gas composition (Blair, Kokabian, & Gude, 2014) and the amount of time that the algae do not obtain sufficient nutrients (starvation time), which triggers lipid accumulation (Vaičiulytė, Padovani, Kostkevičienė, & Carlozzi, 2014). Therefore, the exact biomass composition is hard to measure. As the lipid content of both Chlorella sp. and Nannochloropsis sp. is roughly the same (table 1), it is assumed that the total biomass composition is also roughly the same. The biomass composition used in this research is given below.

Table 10:Biomass composition microalgae, table copied from supplementary data (Ruiz et al., 2016)

Main component	Sub-component	Percentage	Physicochemical properties
Lipids (20%)	Glyco (GL), phospho-lipids (PL)	12%*	Membrane lipids, water insoluble, reference molecule: Phosphatidyglycerol. Volatility and density estimated by group contribution method**.
	Triacylglycerides (TAG)	2%	Intracellular lipids, water insoluble, reference molecule: Triolein. Volatility and density estimated by group contribution method**.
	Waxes	3%	Membrane and intracellular lipids, water insoluble, reference molecule: squalene. Volatility and density estimated by group contribution method**.
	Sterols	3%	Intracellular lipids, water insoluble, reference molecule: cholesterols. Volatility and density estimated by group contribution method**.
Proteins (50%)	Water soluble	20%	Cytosolic proteins, water soluble, refrence molecule: Rubisco, negligible volatility.
	Non-water soluble	30%	Structural proteins, mainly present in cell debris, water insoluble, negligible volatility. No specific reference compund, since they do not participate in thermodynamic equilibrium.
Carbohydrates (20%)	Monosaccharides	5%	Water soluble, negligible volatility, reference component: glucose.
	Polysaccharides	15%	Partially water soluble, negligible volatility, reference component: starch made by 70% amylose and 30% amylopectin.
Pigments (3%)	Total	3%	Poorly water soluble, negligible volatility, reference component: lutien.
Ashes (7%)	Total	7%	Water soluble, negligible volatility, no reference component.

*of which 35% saturated fatty acids, 30% (SFA) monounsaturated fatty acids (MUFA) and 35% polyunsaturated fatty acids (PUFA)

** based on a method described in (Ceriani & Meirelles, 2004)

In the next part, the algae biomass process design will be explained in more detail, from 15% to 20% dry based algae biomass to the final products. The technical process design is divided into the fraction extraction and the algae jet refinery.

During the processing, it assumed that there is no difference between the marine microalgae and the fresh water microalgae. This assumption is made as there is limited data available for the advanced refinery process specific for fresh or salt water algae. Besides, salts are used in the first part of the fractionation and are therefore present in the mixture. Furthermore, other papers do not mention other processing steps when handling fresh or saltwater microalgae (Klein-Marcuschamer et al., 2013; Ruiz et al., 2016).

7.2 Process design

The process to extract all the different fractions of the biomass is a difficult process (Ruiz et al., 2016). Furthermore, the refining of the lipid fraction to kerosene (HEFA) consists of many steps (M. N. Pearlson, 2007; Wang et al., 2016). The factories that can produce HEFA from bio-oil are already present in the Netherlands (in Rotterdam) and therefore these factories are used as reference in this research (EBTP, 2016). The fractionation method used in this research is based on the fractionation method presented in the supplementary data file of the Ruiz et al. (2016) paper.

7.2.1 Fraction extraction

The fraction extraction mainly consists out of 4 steps: cell disruption, extraction of water soluble compounds (proteins and mono/poly-saccharides), extraction of hydrophobic compounds (including the lipids) and exploitation of residual cell debris (non-water soluble proteins) (Ruiz et al., 2016). Total fraction extraction flowsheet is shown in figure 14.

7.2.1.1. Cell disruption

For the extraction of the intercellular fractions of the microalgae biomass, pre-treatment or cell disruption is needed (Homsy, 2012). The highly structured glycoprotein cell walls of the microalgae make direct extraction difficult (Homsy, 2012). Therefore, the extraction efficiency increases by performing pre-treatment (Homsy, 2012). When looking at the scale, the water content of the algae (15% to 20%), the efficiencies and the goal of the pre-treatment step, bead milling seems the most suitable option (Dong et al., 2016). Bead milling is a relative energy intensive process, but the disintegration efficiency and thus the yields are also high (Dong et al., 2016). During the disruption, the temperature is maintained at a constant 25 degrees centigrade and all the energy used in the bead mill is assumed to be dispatched in the form of heat (Ruiz et al., 2016). Therefore, it is assumed that the system is cooled to maintain a constant temperature. The disruption efficiency of the biomass is assumed to be 95% (Ruiz et al., 2016). The assumptions used for the processing cost calculation are given in appendix D.

7.2.1.2 Extraction water soluble compounds

The first fractions that are extracted are the carbohydrates and the water soluble proteins (Ruiz et al., 2016). The extraction process consists out of two stages: direct aqueous extraction and back aqueous extraction, also known as the aqueous two phase system (ATPS) (Mistry, Kaul, Merchuk, & Asenjo, 1996; Ruiz et al., 2016). For this extraction system a mixture of 26% polyethylene glycol (PEG4000) and 15% weight fraction potassium phosphate is added (Ruiz et al., 2016). At these conditions two phases arise, a light aqueous phase containing mainly PEG and a heavier phase containing potassium phosphate, which are separated (II.1 in figure 14)(Ruiz et al., 2016). The partition coefficient of the proteins during the ATPS process is assumed to be 10 at a PH of 7 (Ruiz et al., 2016). A partition coefficient of 10 means that the top phase has a ten times higher concentration than the bottom phase, or in other words a ratio of 10:1 between the top and the bottom phase (Schmidt, Ventom, & Asenjo, 1994).

First the direct extraction is performed, where the carbohydrates and soluble proteins are extracted by PEG4000 from the algae broth (Ruiz et al., 2016). The lipids, pigments, non-water-soluble proteins and ash are separated from the aqueous phase by a mixer-settler in the direct extraction due to sufficient difference in density between the both phases (see II.1 figure 14) (Ruiz et al., 2016). The residence times in the mixer and the settler are 5 and 30 minutes respectively for both the direct and back extraction, which results in a extraction efficiency of 95% (Ruiz et al., 2016).

For the direct extraction, it is assumed that the water volume is evenly distributed between the two different phases created in II.1 (see figure 14) (Ruiz et al., 2016). The back extraction is used as a second extraction unit which transports the proteins and saccharides to a salt-rich phase, with a salt concentration seven times higher compared to the direct extraction (Ruiz et al., 2016). The total energy used for mixing in both extraction units is assumed to be 0.5 KW/m^3, which is assumed to fully dissipate to heat (Ruiz et al., 2016). The heat produced in the extraction units is cooled by cooling water to maintain a 25 degree Celsius medium (Ruiz et al., 2016).

To reduce costs, the chemicals used in the direct extraction unit are recycled. First the phosphate rich phase is concentrated by ultra-filtration(UF) units, with 1 kDa membranes (II.4 see figure 14) (Ruiz et al., 2016). This UF unit recovers 80% of the phosphates by concentrating 5 times. After the UF unit the phosphates from the direct extractor and the PEG rich phase from the back extractor are dried by a spray drier (II.5 + II.6) to fully recycle the chemicals (Ruiz et al., 2016). These spray dryers work on an air/evaporated water ratio equal to five, with a water loss of at least 99%. In the spray driers, the evaporation rate is set to $100kg/(h*m^3)$, the temperature to a constant 40 degrees Celsius and the absorbed power to 0.02 kwh/kg of feed (Ruiz et al., 2016). An average heat efficiency of 50% is used in this research (McCabe, Smith, & Harriott, 1993; Wisniewski, 2015). Finally, after being processed in the backward extractor, the phosphate is concentrated 40 times and recycled by a UF unit (II.3), with a 95% efficiency (Ruiz et al., 2016).

After that, the proteins and carbohydrates are fractionated further by two diafiltration (DF) units, with different filter cut-offs (Ruiz et al., 2016). One DF unit with a 300kDa cut off value is used to separate large polysaccharides (II.7 figure 14) (Ruiz et al., 2016; Safi, Zebib, Merah, Pontalier, & Vaca-Garcia, 2014). In this DF unit the polysaccharides are washed out with water from the monosaccharides and protein fraction (Ruiz et al., 2016). The filter prices are assumed to be the same as the ultrafiltration costs, as these methods are both scaled under ultrafiltration (Koch membrane systems, 2012).

The second DF unit uses a cut-off membrane of 10 kDa, which retains proteins and let through monosaccharides (Safi et al., 2014). During this process, a phosphate buffer solution is used in the DF unit (II.8) to fractionate the monosaccharides from the proteins (Ruiz et al., 2016). The DF unit operation conditions are: a 40 $L/(m^2*h)$ process speed, 0.2 KW/m^2 power consumption of which 10% of the power is dissipated to heat and a lifetime of the membranes of approximately 1000 hours (Ruiz et al., 2016).

The polysaccharides and soluble proteins are dried by a spray drying to obtain pure proteins (II.10) and polysaccharides (II.9 figure 14) (Ruiz et al., 2016). The spray dryer operates in the same way as the earlier mentioned spray dryers.

7.2.1.3 Extraction hydrophobic compounds

After the extraction of the water-soluble proteins and saccharides, the lipid fraction is extracted from the salt rich aqueous phase obtained from ultrafiltration unit II.4 (see figure 14). The lipid extraction takes place by adding a mixture of hexane and isopropanol in a 1:4 ratio to the lipid extractor (III.1) (Halim et al., 2011; Ruiz et al., 2016). This mixture is added in a 10:1 volume ratio with respect to the water phase. The extraction is operated at an optimum temperature of 50 degrees Celsius (Halim et al., 2011; Ruiz et al., 2016).

In the lipid extractor (III.1), a mixer settling tank is used with the same dimensions as the settling tanks used for the protein extraction. A minimum lipid extraction of 85% is assumed during this process (Ruiz et al., 2016). The Hexane and isopropanol are recovered after the lipid extraction by a distillation column (III.2). In this distillation column temperatures are kept below 250 degrees Celsius, as higher temperatures influence the lipid quality (West, Posarac, & Ellis, 2008). The distillation unit operates with a boiler reflux ratio of 1.25 (R/R^{min}), vacuum conditions p= 0.4 atm and condensing temperature of 60 degrees at the top and 200 degrees Celsius at the bottom (Ruiz et al., 2016). To calculate the amount of energy needed for this recycling method, the enthalpy of evaporation for both the hexane (15.7 kj/mol) (Chickos & Acree, 2003) and isopropanol (16.5 kj/mol) are used at 200 degrees Celsius (Wormald & Vine, 2000). The assumed efficiency of the recycle stages is 80% (Ruiz et al., 2016).

After the lipid extraction, the lipid fraction is dewaxed (III.3). The waxes are extracted from the lipid fraction by cooling down all the lipid fractions to 2 degrees Celsius for 5 to 10 hours, after which the solid waxes are separated from the liquid oils by decanting the total oil-solid lipid mixture (Ruiz et al., 2016).

After the de-waxing step, the saponifiable fractions (TAG. GL and PL) are converted into fatty acids methyl esters (FAME) by a transesterification process (III.4) (fig. 14), which operates under alkaline conditions at 60 degrees Celsius and uses 1% NaOH concentration and an excess of 300% methanol (Ruiz et al., 2016). To convert 99% of the lipids into fatty acids, a fixed residence time of 60 minutes is assumed (Dimian & Bildea, 2008; Ruiz et al., 2016). The methanol used during the transesterification is recovered for 95% by evaporation and condensation (III.5) (Ruiz et al., 2016). The evaporation is performed under vacuum conditions and below 150 degrees Celsius to prevent the degradation of glycerol (Dimian & Bildea, 2008; Ruiz et al., 2016). The enthalpy of evaporation for methanol at 150 degrees Celsius is 876 kj/kg (Yerlett & Wormald, 1986).

Finally, the glycerol is washed out in an extraction unit (III.6), with an equal amount of water, obtaining separate glycerol, sterols, pigments and FAME fractions (Ruiz et al., 2016).

The further processing of the bio-oil is given in 7.2.2. algae jet refinery. As HEFA is a hydrotreated fuel, transesterification is not needed for the production of HEFA. However, by using transesterification and the glycerol washer a separate pigment fractions is obtained, which has the highest value of the whole algae (Ruiz et al., 2016). Therefore, the transesterification is still included in this research.

7.2.1.4 exploitation of residual cell debris

The residual cell debris that remains, consists of ash and non-water-soluble proteins. In a DF (IV.1) the salts are washed out of the mixture and the concentration of the non-water-soluble proteins is increased (Ruiz et al., 2016). Again, the same operation conditions for the DF and spray dryer (IV.2) are used. During the DF the ash content is reduced by washing with water (Ruiz et al., 2016).

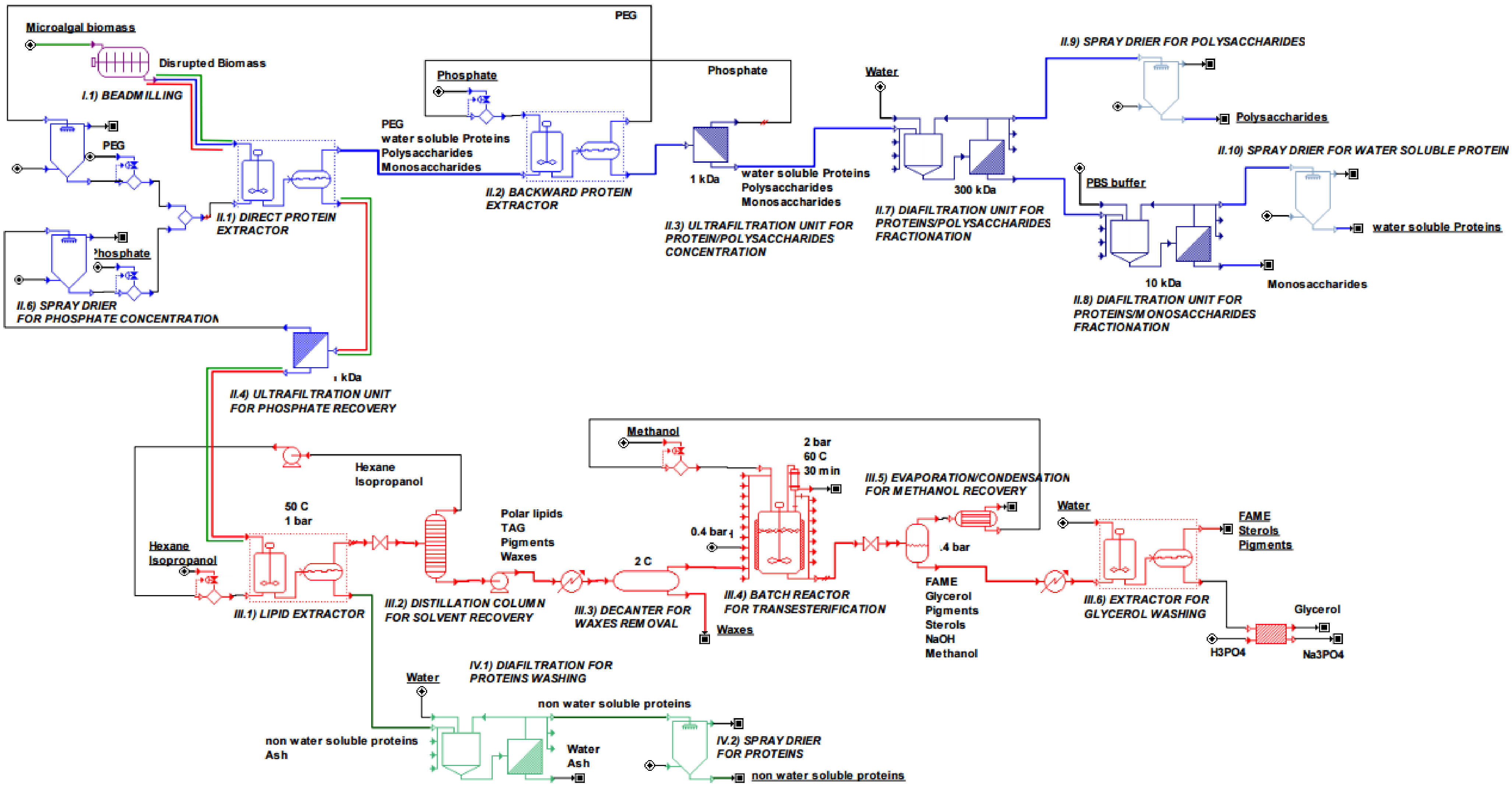


Figure 14: Overview complete fractionation process (Ruiz et al., 2016)

7.2.2 Algae jet refinery

The lipid fraction obtained from the oil extraction in 7.2.1 is further processed to algae jet fuel and other fuels. Due to the transesterification, the composition of the lipid is altered slightly, but this change in composition can be influenced by hydrotreatment (Hulsman et al., 2011). Therefore, the conventional lipid to HEFA process is used. An overview of a HEFA representative process diagram is given below (Wang et al., 2016). It is assumed that the whole lipid fraction is used for the jet refinery.

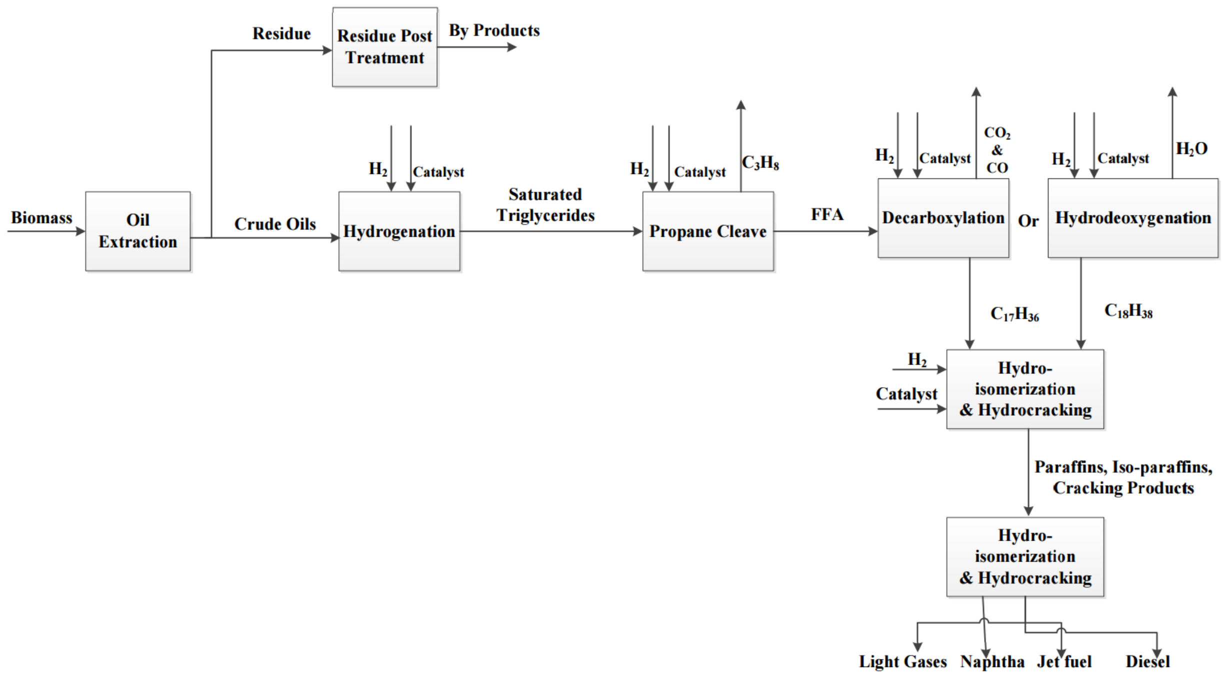


Figure 15: Progress diagram of the biomass to jet process (Wang et al., 2016)

In this flow diagram the first step is to produce saturate double bond glycerides by addition of hydrogen and a catalyst in a process that is called hydrogenation (M. N. Pearlson, 2007). Afterwards, three free fatty acids (FFA) are separated and converted into propane by adding hydrogen in the propane cleaver (M. N. Pearlson, 2007). Aviation fuels need certain specifications including high flash point and cold flow properties (Wang et al., 2016). Therefore, the fuel is deoxidised in a decarboxylation or hydrodeoxygenation process (M. N. Pearlson, 2007). An overview of the process from triglyceride to deoxygenation/decarboxylation is given below.

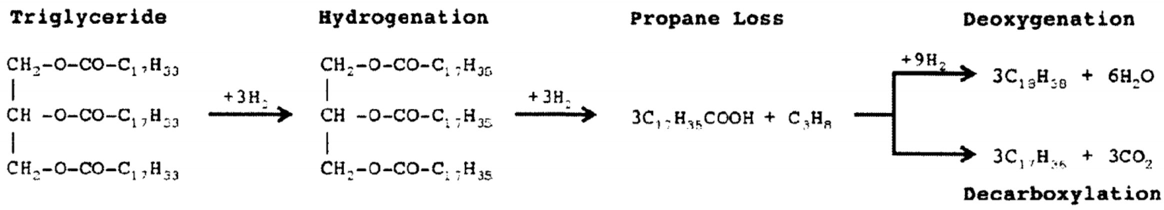


Figure 16: Renewable deoxidation (M. N. Pearlson, 2007)

Thereafter, hydrocracking and hydro isomerisation is used to convert the intermediate fuel into synthetic paraffinic kerosene (SPK) (Wang et al., 2016). Due to the isomerization the straight carbon structures are branched, which lowers the freeze point of the fuel (M. N. Pearlson, 2007). Furthermore, hydrocracking reduces the carbon chain length and therefore influences the type of fuel produced (M. N. Pearlson, 2007; Wang et al., 2016). An example of the isomerization and cracking process is given below.

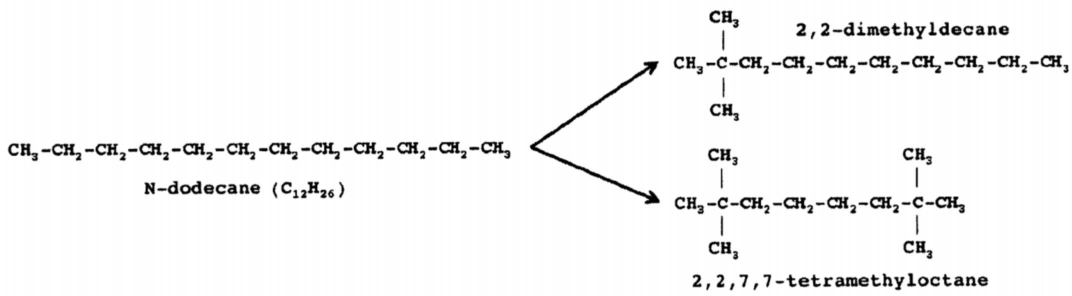


Figure 17: Example isomerization process (M. N. Pearlson, 2007)

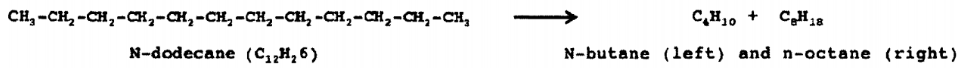
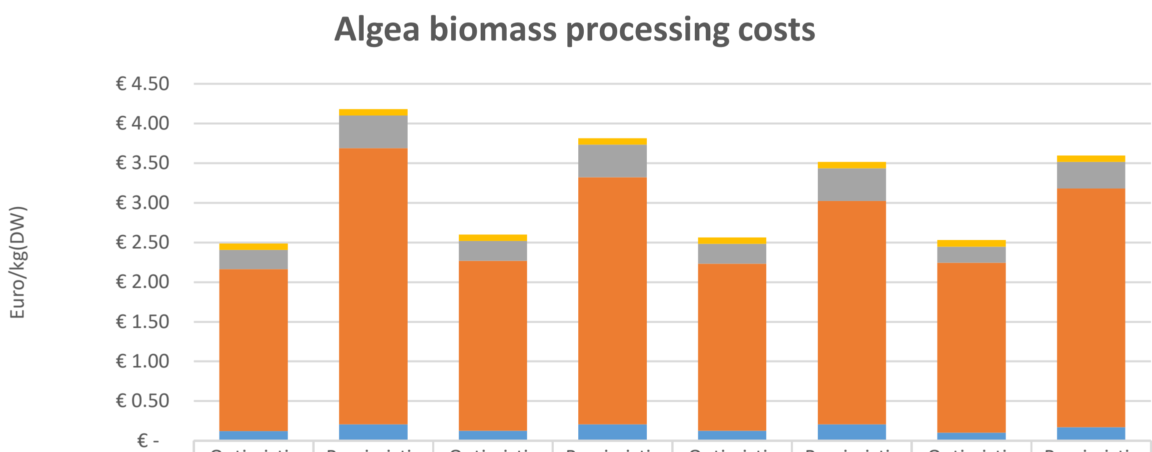


Figure 18: Example cracking process (M. N. Pearlson, 2007)

The goal of the cracking is to obtain as much paraffin (kerosene) as possible and not to crack for too long, which results in a large amount of light gases that cannot be used for the aviation industry (Wang et al., 2016). Catalyst are used to influence the yield of jet fuel (Wang et al., 2016). Therefore, the selection of the correct catalyst is important in this part of the process (Wang et al., 2016). Finally, the different fuels are separated by a fractionation process (for example distillation) (Klein-Marcuschamer et al., 2013), which results into mostly paraffinic kerosene, but also light gasses, naphtha and paraffinic diesel (Wang et al., 2016).

8.0 Algae biomass processing costs

After the technical description in chapter 7, the processing costs of the algae biomass are calculated. As the detailed fractionation process is not yet performed on a large scale, the calculations are done using argued vales, based on best available estimates or averages. The same assumptions are therefore used in the optimistic and the pessimistic scenario. These assumptions are given in Appendix D. The outputs of the algae biomass processing are the valuable fractions which can be sold on the market. Therefore, the cost division in figure 19 below is given per kg(DW) valuable products.



	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
	Open pond		Tubular reactor		Plate reactor		OMEGA	
Total	€ 2.49	€ 4.18	€ 2.60	€ 3.81	€ 2.56	€ 3.52	€ 2.53	€ 3.60
Jet refining	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08	€ 0.08
Other fixed costs	€ 0.24	€ 0.41	€ 0.25	€ 0.41	€ 0.25	€ 0.41	€ 0.20	€ 0.34
Operating costs	€ 2.04	€ 3.48	€ 2.14	€ 3.11	€ 2.11	€ 2.82	€ 2.14	€ 3.01
MEC	€ 0.12	€ 0.21	€ 0.13	€ 0.21	€ 0.13	€ 0.21	€ 0.10	€ 0.17

Scenario

Figure 19: Algae biomass fractionation costs per cultivation method for both the optimistic and pessimistic scenario

The cost divisions per kg (DW) (given in figure 19) does not differ significantly between the different cultivation methods, as many assumptions are the same for the different methods. The small difference in value is caused by the labour costs, which are linked to the total land area needed. Furthermore, the labour costs are connected to the payroll and general plant overheads, which also differ amongst the methods. The difference between the optimistic and pessimistic scenario is caused by the biomass input of the processing system. In the case of the optimistic scenario the input is 20% dry weight algae and in the pessimistic scenario case the input is 15% dry weight algae. Therefore, in the pessimistic scenario more water is processed and therefore more equipment and energy is needed for the processing, causing higher costs.

In this research only a detailed calculation of the fractionation process is performed, as this process does not yet exist in large scale and is not yet fully developed (Ruiz et al., 2016). Therefore, for this process again the MEC, operating costs and the fixed costs are calculated. The exact calculation can be found in the supplementary excel file.

For the processing of the lipid fraction to HEFA explained in 7.2.2, a constant upgrading cost of 0.29 €/L[4] and maximum jet fuel production cost of 0.075 €/L is used (M. Pearlson et al., 2013). This leads to

[4] The refining costs in dollars is converted to euros with a rate of 0.932 Euro/dollar (obtained at 27-03-2017)

a total refining cost of 0.37 €/L for maximizing the amount of HEFA produced from the lipid fraction. The density of the HEFA is assumed to be 900 kg/m^3 (Noureddini, Teoh, & Davis Clements, 1992). It is assumed that the full lipid fraction is sent to these large scale factories (AltAir Fuels, 2016). The assumptions made for the biomass processing calculation are given in appendix D.

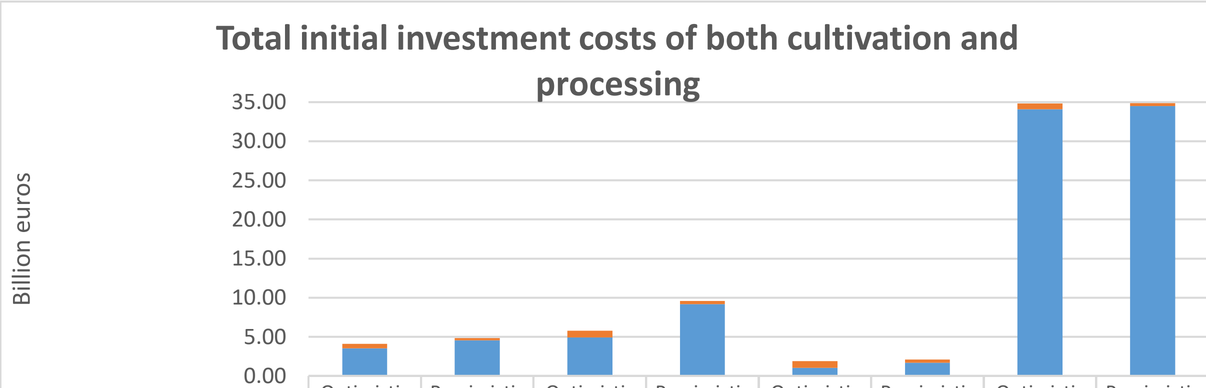
The cost division of the algae biomass processing in Appendix E indicate that the highest costs in the total algae processing are the bead mill energy, spray dry energy, utilities (such as PEG400, Isopropanol, hexane), distillation column for methanol recovery, cooling energy (especially for the bead mill), the labour costs and the jet refinery. As mentioned before the same assumptions are used for each cultivation method, so these components have the highest cost share in each cultivation system and scenario. The exact total costs of both the cultivation/processing and revenues created are treated in the next section.

9.0 Commercial feasibility

The last part of this research is to assess the current and future commercial feasibility of the algae to jet fuel cycle. This assessment is made by calculating the NPV and the PBP along the lines set out in the theory. However, for these formulas the total initial investment costs, annual (operating) costs and revenues are needed. Therefore, first the total costs of both cultivation and processing are calculated. Secondly, the revenue is assessed by looking into market price per kg and size of the HEFA and other valuable biomass fractions. Furthermore, the revenues created by the CO_2 mitigation permits are taken into account. Finally, the current commercial feasibility and an assessment of the potential future commercial feasibility are presented.

9.1 Total costs

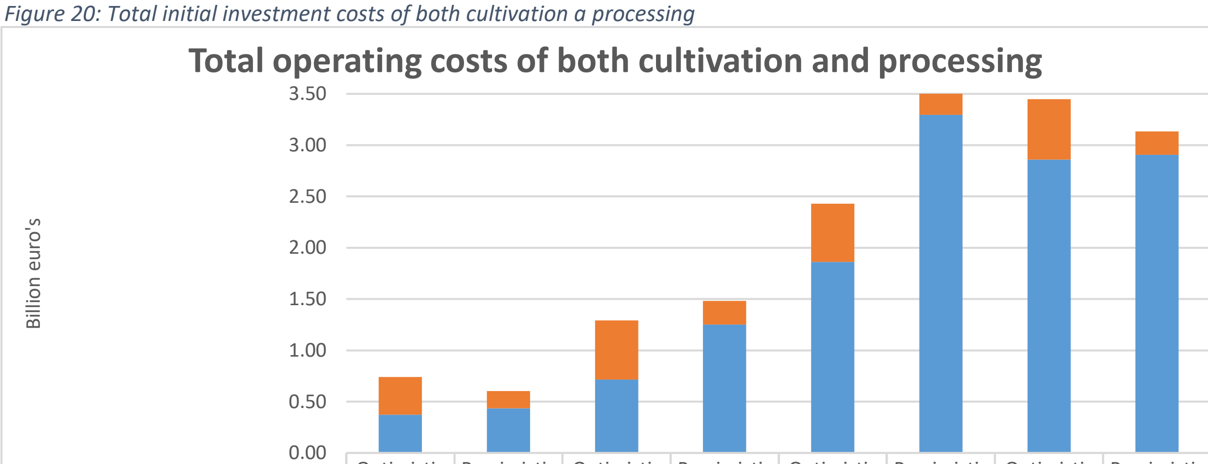
The final costs are divided in the total initial investment costs (MEC and other fixed costs) and the operating costs (annual costs including algae jet refinery). The initial investment costs are paid once at the start of the project and the operating costs are paid every year for the total lifetime of 25 years. The total investment costs and the operating costs are given below (figure 20/21):



	Open pond		Tubular		Flat plate		OMEGA	
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
Total	4.12	4.83	5.80	9.57	1.91	2.08	34.83	34.85
Investment costs processing	0.58	0.27	0.88	0.40	0.88	0.40	0.73	0.33
Investment costs cultivation	3.54	4.56	4.92	9.17	1.02	1.68	34.10	34.52

Scenario

Figure 20: Total initial investment costs of both cultivation a processing



	Open pond		Tubular		Flat plate		OMEGA	
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
Total	0.74	0.60	1.29	1.48	2.43	3.50	3.45	3.13
Operating costs processing	0.37	0.17	0.58	0.23	0.57	0.21	0.59	0.23
Operating costs cultivation	0.37	0.44	0.71	1.25	1.86	3.29	2.86	2.91

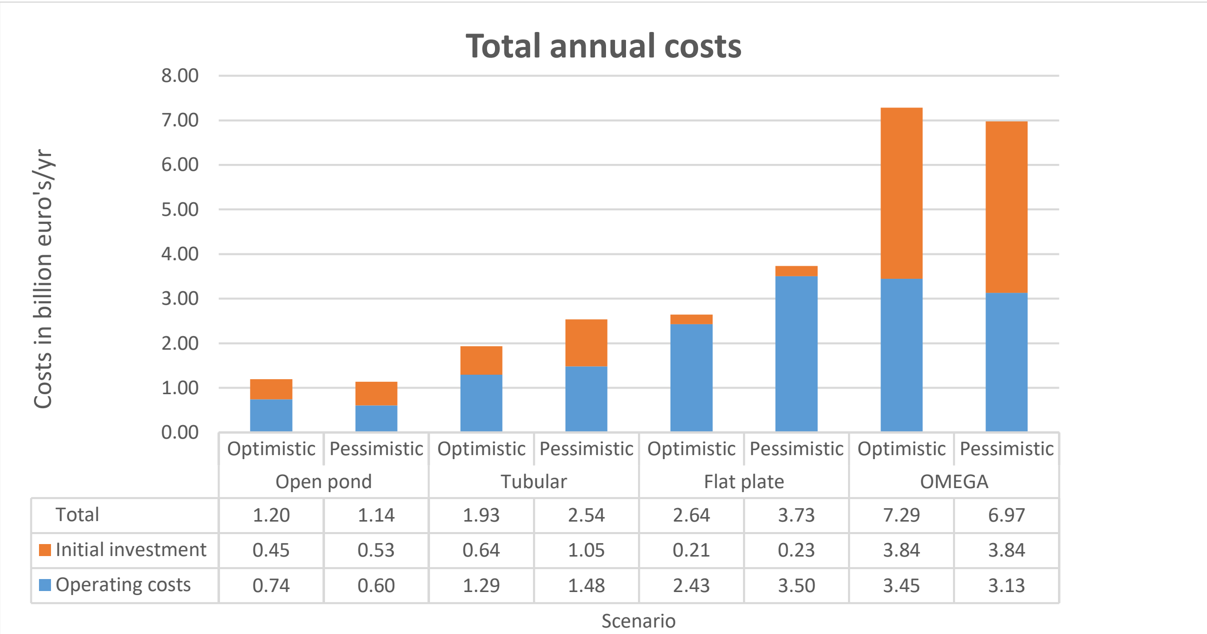
Scenario

Figure 21: Total operating costs of both cultivation and processing

From these figures, it becomes clear that the highest costs are caused by the algae cultivation. Especially when looking at the initial investment costs (figure 20). The initial investment costs in the case of algae cultivation are a lot higher than the initial investments for algae processing. Overall the plate reactor shows the lowest initial investment and the OMEGA reactor the highest.

However, when looking at the operating costs (figure 21), the plate reactor is shown to have significantly higher operating costs than the open pond and the tubular reactor. Furthermore, processing in general shows a higher cost contribution in the total operating costs. The pessimistic scenario of the plate reactor shows the highest operating costs and the pessimistic scenario of open pond the lowest. The pessimistic values for the operating costs of algae processing are overall lower due to the lower amount of biomass processed (see figure 21) and therefore a lower amount of energy and utilities needed in this step. However, the operating costs of algae cultivation are higher in the pessimistic scenario. These higher costs are caused by the larger medium volume during cultivation in the pessimistic scenario. Due to the high costs for circulation of the medium in the case of the tubular and plate reactor, a total higher operating cost for the pessimistic scenario of these methods is obtained (see figure 21). This is not the case for the open pond and OMEGA reactor with lower circulation costs. So overall, the operating costs for the cultivation are higher and the operating costs for processing lower for the pessimistic scenario. For the optimistic scenario, this is the other way around.

Below, the final annual costs are given. The annual costs are calculated by multiplying the initial investment costs with the annuity factor and adding them to the operating costs. From figure 22 it becomes clear that the OMEGA remains the most expensive option and the open pond the cheapest.



	Open pond		Tubular		Flat plate		OMEGA	
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
Total	1.20	1.14	1.93	2.54	2.64	3.73	7.29	6.97
Initial investment	0.45	0.53	0.64	1.05	0.21	0.23	3.84	3.84
Operating costs	0.74	0.60	1.29	1.48	2.43	3.50	3.45	3.13

Figure 22: Total annual costs

9.2 Total revenue

In the next section, the revenues from the algae fractions are calculated. The revenues of all the separate fractions are assessed by investigating the market size and price. Furthermore, the revenue created by the CO_2 mitigation permits is incorporated.

9.2.1 Market size and price

Table 10 in section 7.1 shows a microalgae composition of 20% lipids, 50% proteins, 20 carbohydrates, 3% pigments and 7% ashes. Each of these fractions have value, except for the ashes (Ruiz et al., 2016). Figure 23 shows an overview of the targeted products. From this figure can be concluded that the biofuel market has the lowest market value of all the different markets (Ruiz et al., 2016). The revenues of the co-products are therefore very important in making algae commercially viable.

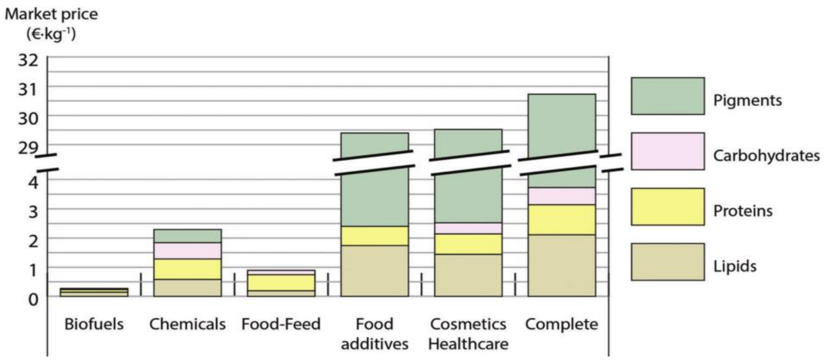


Figure 23: Market value and total selling price per market scenario. Each scenario aims to maximize revenue (Ruiz et al., 2016)

This figure suggests the lowest price for biofuels. However, the total market size of commodities, including biofuels, is almost unlimited and the market for the high-value products is more restrained (Ruiz et al., 2016). The pigments for example, which are sold as antioxidants, have a worldwide total market volume of around 1 billion dollars or 1000 ton/yr (Ruiz et al., 2016). Whereas, the pigments have the highest value in the food additives and cosmetics industry (see fig. 23). The total useful mass per fraction obtained from the biomass cultivation and processing is given below (Table 11).

Table 11: Total final useful biomass obtained per fraction from the cultivation and processing steps

Fraction	Open pond		Tubular reactor		Plate reactor		OMEGA		Unit
	Opt	Pes	Opt	Pes	Opt	Pes	Opt	Pes	
Proteins	92.7	25.5	139.1	38.2	139.1	38.2	142.0	39.0	kton/yr
Water soluble proteins	35.9	9.9	53.9	14.8	53.9	14.8	55.1	15.1	kton/yr
Non-water-soluble proteins	56.8	15.6	85.1	23.4	85.1	23.4	86.9	23.9	kton/yr
Carbohydrates	37.8	10.4	56.8	15.6	56.8	15.6	58.0	15.9	kton/yr
Monosaccharides	9.5	2.6	14.2	3.9	14.2	3.9	14.5	4.0	kton/yr
Polysaccharides	28.4	7.8	42.6	11.7	42.6	11.7	43.5	12.0	kton/yr
Lipids	29.3	8.0	43.9	12.1	43.9	12.1	44.8	12.3	kton/yr
Jet	16.1	4.4	24.1	6.6	24.1	6.6	24.6	6.8	kton/yr
Diesel	7.6	2.1	11.4	3.1	11.4	3.1	11.6	3.2	kton/yr
Naphtha	2.3	0.6	3.4	0.9	3.4	0.9	3.5	1.0	kton/yr
LPG	1.4	0.4	2.1	0.6	2.1	0.6	2.1	0.6	kton/yr
Pigments	13.2	3.6	19.9	5.5	19.9	5.5	20.3	5.6	kton/yr
Total useful biomass	173.0	47.6	259.6	71.4	259.6	71.4	265.1	72.9	kton/yr

The final fraction mass is linked to the total market size and price for each of these fractions. The market size and price per fraction is given in table 12. To calculate a total annual revenue for each of the co-products and the HEFA the following formula is used:

$$Total\ revenue\ fraction\ method = \sum_{i=x}^{n} Fraction\ mass(x) * Fraction\ price(y)$$

X= the specific fraction

*Y= the specific price of the fraction**

**this price can differ if the market with the highest value is saturated and the fraction is sold for a lower price in the other market segments*

Table 12: Market size and average price of the different fractions (for complete biorefinery with the focus on jet fuel)

Fraction	Targeted market	Market size (kton/yr)	Price (€/ton)	Source
Proteins	Food additives (Water soluble protein)	2,300	3,300	(Arla Foods Ingredients Inc., 2013; Ruiz et al., 2016). market share in 2012
	Biopolymers (Non-water-soluble proteins)	1,446	1,400	(PRNewswire, 2014; Ruiz et al., 2016), is the total biopolymer market size, not only proteins, but is big enough for all.
Carbohydrates	Biopolymer	1,446	2,000	(PRNewswire, 2014; Ruiz et al., 2016)
Lipids	Jet (HEFA)	53,700	480	(de Jong et al., 2015; FuelsEurope, 2015), based on market size of 2014
	Diesel	270,300	446	
	Naphtha	48,700	363	
	LPG	26,700	288	
	Propane		288	(de Jong et al., 2015), propane and LPG are taken together is they are very similar.
Pigments	Food additives (natural antioxidants)	1	900,000	(Ruiz et al., 2016)
	Pigment for paint (Synthetic pigment market)	155	15,000	(Grand View Research, 2017; Ruiz et al., 2016)

When looking at table 11 and 12, it is clear that only the natural antioxidants (pigments) market is saturated, for all cultivation methods and scenarios. As the natural antioxidants market size is 1 kton/yr (table 12) and the obtained pigment biomass is 3.6 to 20.3 kton/yr (table 11).

9.2.2 CO_2 permit revenue

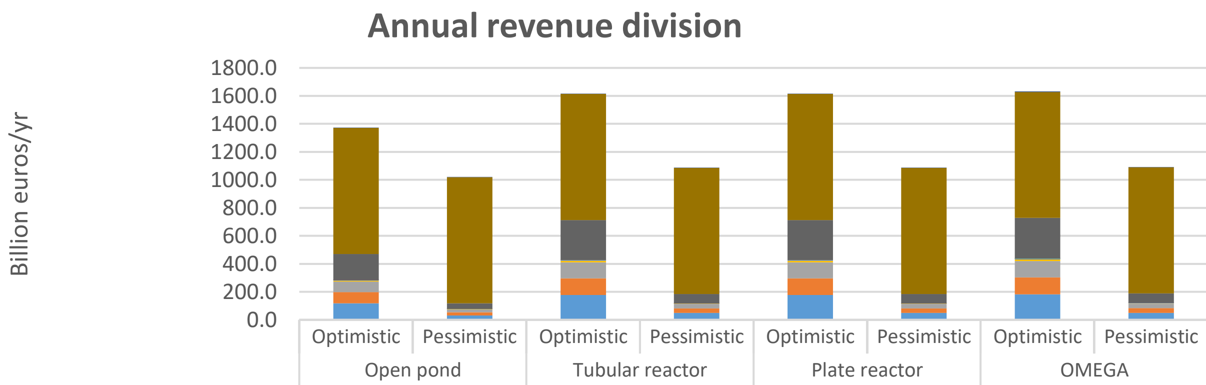
The revenue created from the CO_2 mitigation is assumed to be €5 per tonne CO_2 mitigated (Parry, Shang, Wingender, Vernon, & Narasimhan, 2016; Zechter et al., 2016). In section 4.3 (Land occupation) of this research, the assumptions used for the total CO_2 mitigation calculation are given. These assumptions result in the following mitigation mass, including the assumption that all the other energy used in the system is produced from zero carbon energy sources (table 13).

Table 13: CO_2 mitigation mass

Cultivation method	Optimistic	Pessimistic	Unit
Total mitigated CO2 (open pond)	400	400	kton/yr
Total mitigated CO2 (tubular)	600	600	kton/yr
Total mitigated CO2 (plate)	600	600	kton/yr
Total mitigated CO2 (OMEGA)	500	500	kton/yr

9.2.3 Total revenue division

From the market size, price and the CO_2 permit revenue a total revenue division is made. This revenue division is made by multiplying each market size with the market costs and adding the obtained CO_2 mitigation revenue to this value. The total annual revenue division is given below (figure 24).



	Open pond		Tubular reactor		Plate reactor		OMEGA	
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic
Total	1372.3	1020.4	1615.9	1088.2	1615.9	1088.2	1630.9	1091.9
CO2 permits	2.0	2.0	3.0	3.0	3.0	3.0	2.5	2.5
Pigments high value	900.0	900.0	900.0	900.0	900.0	900.0	900.0	900.0
Pigments low value	183.6	39.6	283.0	66.9	283.0	66.9	289.3	68.7
Propane	0.6	0.2	0.8	0.2	0.8	0.2	0.9	0.2
LPG	0.4	0.1	0.6	0.2	0.6	0.2	0.6	0.2
Naphtha	0.8	0.2	1.2	0.3	1.2	0.3	1.3	0.3
Diesel	3.4	0.9	5.1	1.4	5.1	1.4	5.2	1.4
Jet	7.7	2.1	11.6	3.2	11.6	3.2	11.8	3.3
Carbohybrates	75.7	20.8	113.5	31.2	113.5	31.2	115.9	31.9
Non-water soluble proteins	79.5	21.8	119.2	32.8	119.2	32.8	121.7	33.5
Water soluble proteins	118.6	32.6	177.9	48.9	177.9	48.9	181.7	50.0

Scenario

Figure 24: Annual revenue division

Figure 24 indicates that most revenue is obtained from the high and low value pigments. Furthermore, it is clear the amount of revenue created from the jet fuel and other fuels is almost negligible. Similarly, the revenue created by the CO_2 permits is also very small. These low values are obtained due to the lower selling price of these products with respect to that of the pigments (see table 10). Finally, the carbohydrates and proteins show a significant contribution to the total cost division, especially in the optimistic scenario. All the other fractions have no significant contribution to the total revenue.

9.3 Current commercial feasibility

As indicated in the theory, the commercial feasibility is assessed by the NPV and the PBP. For this calculation, the total costs and revenue calculated in section 9.1 and 9.2 are used and filled into the NPV and PBP formula. The formulas used are given again below (Blok, 2006):

I= Initial investment
B= Annual benefits
C= Annual costs
α= The capital recovery factor (or annuity factor)
r= Discount rate
L= Lifetime of the project

$$NPV = -I + \left(\frac{B-C}{\alpha}\right)$$

$$\alpha = \frac{r}{1-(1+r)^{-L}}$$

$$PBP = \left(\frac{I}{B-C}\right)$$

The input values for this formula are given below for clarity:

Table 14: Total revenue per cultivation method

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	1.37	1.02	billion€/yr
Tubular	1.62	1.09	billion€/yr
Plate	1.62	1.09	billion€/yr
OMEGA	1.63	1.09	billion€/yr

Table 15: Total initial investment per cultivation method

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	4.13	4.83	billion€
Tubular	5.80	9.57	billion€
Plate	1.91	2.08	billion€
OMEGA	34.83	34.85	billion€

Table 16: Total operating costs per cultivation method

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	0.74	0.60	billion€/yr
Tubular	1.29	1.48	billion€/yr
Plate	2.43	3.50	billion€/yr
OMEGA	3.45	3.13	billion€/yr

By filling in these values into the NPV and PBP formula, the current commercial feasibility can be assessed. Again, the lifetime is assumed to be 25 years and the discount rate to be 10%. The final values of the NPV and PBP are given below. Note that a positive NPV indicates a commercially feasible project (Blok, 2006).

Table 17: Current NPV and PBP for each cultivation method for the whole project

	Open pond		Tubular		Plate		Omega		
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Unit
NPV	1.60	-1.05	-2.86	-13.14	-9.30	-23.98	-51.34	-53.38	Billion Euros
PBP	6.54	11.61	17.91	-24.35	-2.34	-0.86	-19.16	-17.08	Years

From this table, it becomes clear that only the optimistic scenario of the Raceway pond cultivation method is commercially feasible, as this scenario gives a positive NPV. Furthermore, for this scenario the PBP lies within the 10 and 20 years estimate which is an average value for large scale projects (Leanmanufacture.net, 2009). However, all the other scenarios and cultivation methods show a negative NPV, which indicates that it is not profitable to invest in the system (Blok, 2006). Some scenarios even show a negative PBP which is caused by higher annual operating costs than revenues, which can be seen in table 14 and 16.

Besides, the amount of produced HEFA is small compared to the total amount of kerosene used at Schiphol each year (3.5 million tons) (van der Wielen et al., 2014). The exact market share is given below, based on the jet fuel values of table 11:

Table 18: Share of AJF compared to the total amount of jet fuel used at Schiphol

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	0.46	0.13	Share (in %)
Tubular	0.69	0.19	Share (in %)
Plate	0.69	0.19	Share (in %)
OMEGA	0.70	0.19	Share (in %)

9.4 Future commercial feasibility

For the assessment of the future commercial feasibility the gap towards the current commercial feasibility is assessed. In other words, are there options that could lead to commercial feasibility? To answer this, firstly it is assessed which products are or could be commercially feasible in comparison to the costs per kg of valuable products. Furthermore, cost reductions indicated by literature are included in this overview to find out if algae production for AJF might become feasible in the future. Besides, it is investigated how much the price of HEFA and or the CO_2 permits would have to increase in order to make the projects commercially feasible (i.e. the moment where the NPV becomes 0).

9.4.1 Revenue vs costs

Figure 25 shows the revenue for each fraction and the total costs per cultivation method. The figure indicates the relation between the revenues created by fraction and the costs per kg of dry valuable product for each of the cultivation methods. All the valuable fraction above each cost line have higher revenues than costs. The values below the cost lines have higher costs then revenues. Furthermore, the value of the products is given from high to low. On the left-hand side, the pigments with the highest revenue per kg are given (€900). On the right-hand side, the jet fuel (0.48 €/kg) and other fuel fractions (0.446 to 0.288 €/kg) with the lowest value are given. This means that only the residues of the algae biomass fractions that could not be sold in higher market segments will be sold at the fuel market. An example are sterols, they are part of the lipid fraction and could be sold as feed for fuel. However, the sterols themselves have a much higher value and therefore will probably be sold as sterols instead of feed for fuel (see figure 25).

In this research, the full microalgae are used, even the fractions that show lower revenue than costs. This approach is used as the highest costs are made for the cultivation and production of 15% to 20% dry based biomass (see figure 20 and 21). The highest costs are thus already made before the processing of the algae biomass into valuable fractions. Therefore, each fraction which could produce revenue is taken into account in this research.

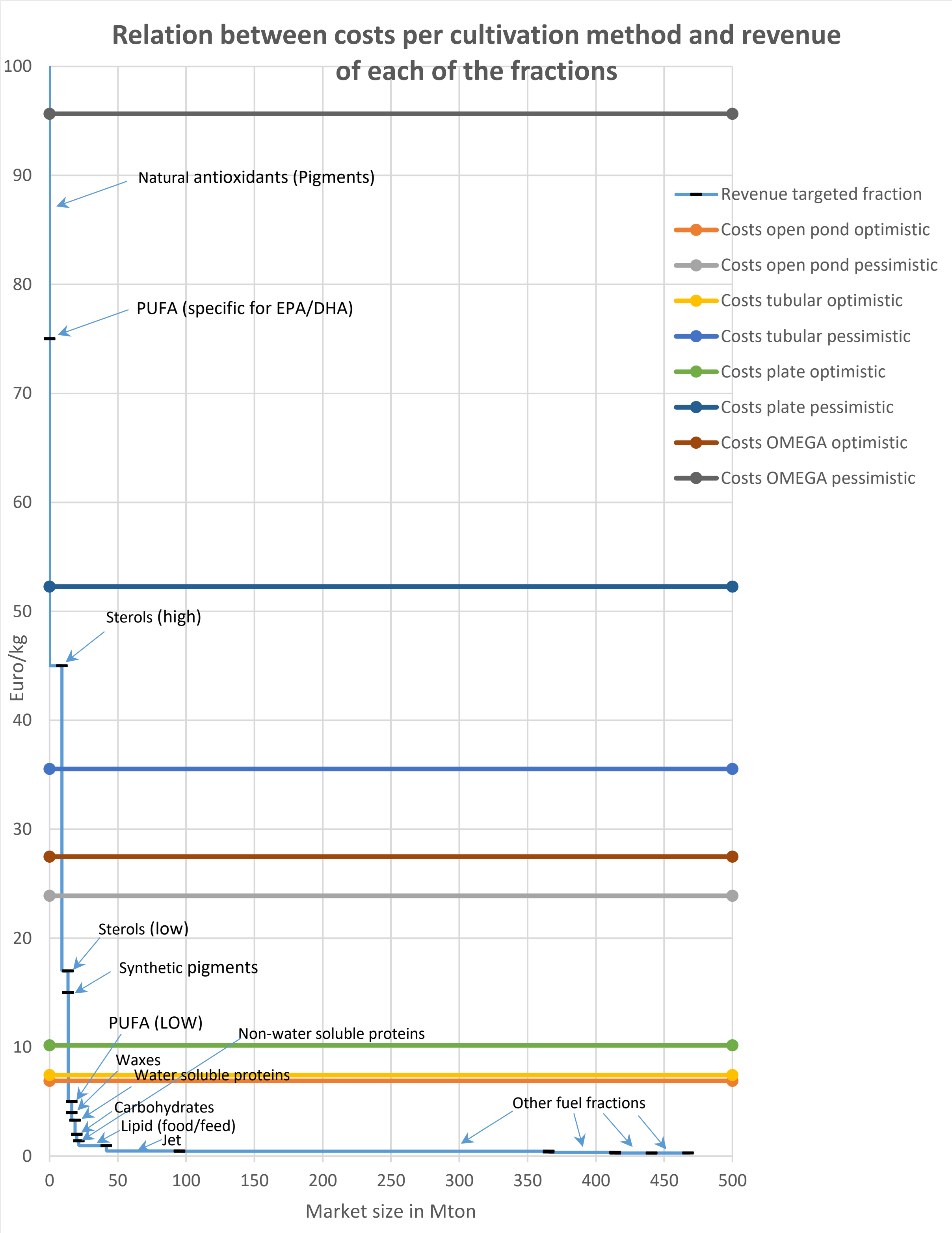


Figure 25:Overview of revenue of each fraction compared to the costs per kg of each cultivation method

As this research focusses on jet fuel production, the other markets for the lipid fraction were not considered earlier in this research. However, part of the lipids could also be sold at other markets. These markets are added to figure 25. These other market sections have higher values then the fuel markets and sometimes also a large market size, such as the food and feed lipids with a value of 0.95 €/kg (Ruiz et al., 2016). Furthermore, other products such as waxes (4 €/kg), sterols (17-45 €/kg) and

PUFA (5-75 €/ton) are part of the lipid fraction and have even higher values (Ruiz et al., 2016). Therefore, it would be illogical to focus on the fuel market (including jet fuel) in the future, as these products have the lowest value.

The assumptions for the extra markets used in figure 25 are given below:

Table 19: Overview market price and size of other possible markets

Fraction	Market size (kton/yr)	Price (€/ton)	Source
Waxes	42	4	(CBI, 2015; Ruiz et al., 2016), this is the total European market size
Lipid (food/feed)	20,000	0.95	(Akoh, 2005; Ruiz et al., 2016), estimation based on global revenue
PUFA (EPA/DHA)	47.5	75	(Global Market Insights, 2016; Ruiz et al., 2016), assumed 50% of 2022 values (see figure source)
PUFA (food additives)	2500	5	(Menayang, 2016; Ruiz et al., 2016), estimation based on global revenue
Sterols (high) as food additives	9000	45	(Grand View Research, 2015; Ruiz et al., 2016; Transparency Market Research, 2016), 50% of total market size sterols.
Sterols (low) for cosmetics	4500	17	(Grand View Research, 2015; Ruiz et al., 2016; Transparency Market Research, 2016), assumed 25% of total market size sterols.

The market size in figure 25 indicates that the fuel market is by far the biggest market of all the fractions. The market sizes are an approximation, as their exact market sizes are often unclear. In this research, the highest amount of lipid fraction obtained is 44.8 kton in the optimistic scenario of the OMEGA method. When looking at table 19, it can be seen that all the higher markets for the lipid fractions are saturated at approximately 35 Mton. It is very unlikely that this amount of lipid biomass is obtained in the near future. Therefore, the fuel market will probably not be used in the near future.

9.4.2 Conventional jet fuel vs algae jet fuel prices

Conventional jet fuel prices will increase in the future (Carter, 2012). However, as the oil price dropped in 2014/2015, the conventional jet fuel price also dropped (IATA, 2017). To assess if it is likely that the conventional jet fuel is going to be priced equally to the AJF in the future, the difference in price is investigated. The price of the AJF is calculated by filling in NPV=0 and all the costs and revenues into the NPV formula, except for the HEFA price. The prices and difference compared to conventional jet fuel are given below:

Table 20: Prices HEFA for future commercial feasibility

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	-11	27	Thousand Euro/ton
Tubular	14	219	Thousand Euro/ton
Plate	43	399	Thousand Euro/ton
OMEGA	230	869	Thousand Euro/ton

Table 20 shows what prices HEFA has to be to make each scenario commercially feasible (NPV=0). The prices are extremely high compared to the current jet fuel price of 480 €/ton (de Jong et al., 2015). Only for the optimistic scenario of the open pond a negative value is indicated, as the open pond is already commercially feasible (NPV>0) without higher revenue from HEFA (see table 17). The high price of the other scenarios is caused by the fact that the complete deficiency of the NPV is corrected by the HEFA. These HEFA prices will probably never be met.

9.4.3 CO_2 pricing

The current CO_2 price is around 5 €/ton CO_2 mitigated (Parry et al., 2016; Zechter et al., 2016). This price will probably increase as the amount of CO_2 in the air will increase (IPCC, 2008). The price needed to obtain a NPV of 0 is calculated in this section. The prices and difference compared to the current CO_2 prices are given below:

Table 21: Prices CO_2 permits for future commercial feasibility

Cultivation method	Optimistic	Pessimistic	Unit
Open pond	€ -438	€ 295	Euro/ton
Tubular	€ 530	€ 2,418	Euro/ton
Plate	€ 1,713	€ 5,291	Euro/ton
OMEGA	€ 11,316	€ 11,766	Euro/ton

Table 21 shows the CO_2 permit prices for a NPV of 0. Literature indicates that only in a very optimistic case the CO_2 permit price could increase to 120 €/ton in 2050 (Luckow et al., 2015). If this price is compared to the prices displayed in table 21, it is clear that the current prices are way too high and will probably never be met without strong cost reductions. Again, a negative value is indicated for the open pond optimistic scenario, as this scenario is again already commercially feasible without higher revenue from CO_2 pricing.

9.4.4 Optimal scenario

Finally, both an improvement in the AJF and CO_2 permit price is taken into account in an optimal scenario. Afterwards, the NPV and PBP are calculated one more time to check the commercial feasibility in the future, while using this optimal scenario. This optimal scenario consist of a cost reduction of 30% for both initial investment and operating costs, as this reduction seems achievable from literature (Ruiz et al., 2016). Furthermore, the CO_2 permit price of 120 €/ton (Luckow et al., 2015), used earlier. Furthermore, a HEFA price of 1200 €/ton is assumed, as another paper indicates a needed price premium of around 762 €/ton on top of the current jet fuel price (480 €/ton) to improve the transition to alternative jet fuels (de Jong et al., 2017). The NPV and PBP are given below.

Table 22: Future NPV and PBP of the optimal scenario

	Open pond		Tubular		Plate		Omega		
	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Optimistic	Pessimistic	Unit
NPV	5.39	2.49	3.18	-5.57	-1.33	-13.16	-30.81	-33.82	Billion Euros
PBP	3.16	5.23	5.09	53.58	2215.43	-1.13	-34.43	-23.50	Years

This final table indicates that only the optimistic and pessimistic scenario of the open pond method and the optimistic scenario for the tubular reactor have now a positive NPV. The pessimistic scenario of the tubular reactor still shows a negative value, thus indicating that it is not sure if this cultivation method will be profitable. Only the open pond method could be commercially feasible with these extremely optimistic assumptions. The plate reactor and OMEGA method are still not commercially feasible without even more drastic cost reductions or revenue improvements.

There are thus strong indications that it will not become commercially feasible to use microalgae for the production of AJF in the Netherlands in the future. Only the open pond method could be commercially feasible in the future, in an optimal scenario. However, when looking at the land occupation of this method this system will probably never be used in the Netherlands. Furthermore, a lot more improvements have to be made to make the other methods commercially feasible in the future; especially for the offshore OMEGA method, which is by far the most expensive option.

10. Discussion

Limitations

In this research assumptions are used to make a general cost calculation for the cultivation and processing of microalgae. However, these assumptions only give an indication of the real values. This is the case as the assumptions used in this research are often obtained from lab research, which is more controllable then the open pond system and other cultivation reactors that are situated in the outdoors. For example, for cultivation the same average value for the annual productivity is used each year. However, in real life these productivities will change, as there will be years with good and bad harvests. Furthermore, temperature will influence the productivity of the system. In this research however, it is assumed that the temperature is always optimal during the cultivation step, this is naturally very unlikely as for example cloud cover will influence the radiation and thus the temperature. For the processing of microalgae, a lot more research is needed, especially for the fractionation process, as this is not yet performed on a large scale. Therefore, the assumed values could differ a lot from the actual large-scale situation.

Marine and fresh water algae are assumed the same in this research, as there is no clear evidence that they are treated differently during processing and/or cultivation. However, as mentioned in this research, every algae species is different and therefore the productivity and optimal growth medium will also slightly differ for each cultivation method. Therefore, the costs linked to these algae only will give an estimation.

The water costs are not taken into account in this research as the costs for water are very low and the water is assumed to be recycled. In reality, the water for algae growth cannot be recycled endlessly, as the algae will grow less efficient after recycling the water a couple of times (Jongbloed, R., personal communication, 22 March, 2017). Furthermore, it is unclear if all water sources can be used for the cultivation medium, as the water is only cleaned by the large filter units at the start of the cultivation system. Because the price of water will depend on the water source (for example ground water), these costs will differ. In this research, every system is assumed to be close to water, resulting in low costs for a water infrastructure. In reality, this could differ per location and therefore the water pump costs assumed in this research will change.

The open pond systems in this research do not include a roof or other type of cover. However, in the Netherlands with low temperatures and large amounts of pathogens in the air, a roof could be necessary (SCHOTT, 2017). This is confirmed by one of the large scale open pond system in the Netherlands (Jongbloed, R., personal communication, 22 March, 2017).

A continued algae cultivation and processing during the growth season is assumed in this research. However, the cultivation systems have to be cleaned regularly, the cultivation will be stopped periodically. Therefore, the growth season and thus the amount of biomass obtained will be lower than assumed. However, without cleaning the productivity will be lower due to fouling (Darzins et al., 2010).

The OMEGA method in this research could be damaged by for example high waves (Trent, 2013) or corrosion; yet this is not accounted for in this research. Besides that, a multitude of difficulties for the offshore method could emerge, for example that the full water cleaning facility has to be built offshore. Furthermore, the exact costs and operation for the pipeline are unclear, as this depends on the pressure, diameter, material and length of the pipeline. Therefore, general assumptions are used. Finally, in this research only sea water algae are cultivated in the OMEGA system, as it seems illogical to transport freshwater to the sea. However, in the paper of Trent (2013) only freshwater algae are considered and therefore the growth rate and for example the fouling by salt could influence the costs and revenue of this system.

Furthermore, it is assumed that all the power used in this research is obtained from CO_2 neutral sources. In the future, this will probably be the case, but in the short term this energy will in reality not fully be supplied by carbon neutral sources. This results in lower CO_2 permit revenues and thus a less commercially feasible project.

Finally, the final products of the processing are sold in different markets. Each of these markets has its own quality requirements. For example, the food/feed quality needs to be much higher than the fuel quality, as impurities could make people sick. Extra steps could be needed to assure this quality, but these steps are not taken into account in this research. Furthermore, the full global market of high value pigments is saturated in this research. However, in reality this will probably not be the case, as there could be more suppliers that already have contracts or will lower the price due to competition. Therefore, it is unlikely that the full market sizes of all the market fractions will be filled up with the valuable products of one facility.

Relevance

This research gives an indication of the current commercial feasibility of algae jet fuel. Therefore, other companies or countries could learn from this research, as it gives insights into where the highest costs of the project are and if it is feasible to pursue an algae biomass project. Due to the clear step by step approach used in this research, a cultivation and process framework is created which could be useful for other cases.

Furthermore, the future prospects are treated in this research. Therefore, this research gives insight in possible future commercial feasibility. Due to this research, it is apparent that the investment in algae cultivation and processing systems for algae jet fuel is only profitable if there are large cost reductions or if the revenue is increased. Furthermore, even with these cost reductions or increased revenue, other markets will still have higher value and thus will probably be explored first. Therefore, investment in algae jet fuel will not be performed in the short term and probably also not directly in the long term. This research thus indicates the direction for further research.

As this research is commissioned by the Schiphol group and Tata Steel, these companies can learn from these conclusions as it shows possible new markets and gives insight in the commercial feasibility of the project. Importantly, this project shows that it is currently not commercially feasible to invest in the cultivation and processing of algae jet fuel in the Netherlands. Therefore, this research may act as a guideline in determining the strategy of both companies.

Further research

In the results section, it became clear that the targeted market (algae jet fuel) of this research is not the most profitable one. Therefore, more research is needed into the higher market segments, especially for the lipid market. This includes a detailed assessment of the exact global but also local market size of different products. Furthermore, a price assessment has to be done, as a large-scale algae plant will saturate some high-end markets and therefore probably lower the price.

The productivity and thus the commercial feasibility of each system greatly depends on the solar irradiation of a specific area. Therefore, further research is needed to find out which algae can grow best in these higher solar irradiated areas. However, costs for cooling will increase in these areas and will have to be taken into account. Furthermore, more research is needed to improve the photosynthetic efficiency of the cultivation methods, as the current efficiencies are far below the optimum level of 8 to 10% (Melis, 2009).

The processing of the microalgae has to be investigated more intensively, as it is unclear which design is best to efficiently fractionate and extract all the valuable fractions. Further research must focus on the fractions with the highest values and how to extract and fractionate these fractions efficiently in order that as much mass (and thus revenue) as possible is obtained for the lowest costs. In this research, a theoretical design is used, in order to map the main costs drivers. However, these main costs drivers could be lowered by more efficient processing routes obtained by further research.

In this research only a general land occupation indication per cultivation method is given. However, a detailed analysis of the available land is not performed in this research. Some cultivation methods such as the open pond systems need a large amount of land for their cultivation. It is thus needed to identify whether the required amounts of land are available and what the exact total costs for this land will be, as this will differ per region.

As mentioned in the last part of this research, costs have to be reduced significantly to reach commercial feasibility. Therefore, any research which could lower the costs will aid in making algae jet fuel more commercially feasible. Especially the cultivation costs will have to be lowered as they have by far the highest cost share in the project. Furthermore, the benefits for the total use of the flue gas of Tata Steel has to be investigated, as the exact cost reductions for nutrient supply by using this gas is uncertain and therefore not taken into account in this research. The same holds for the use of wastewater in the system.

Finally, in general more research into microalgae strains is needed. This includes the exploration of new algae strains and the modification of current ones. Some research already indicates microalgae that produce and deposit oil straight into the medium itself. This eliminates the need for extraction, which will lower the processing costs and improves the efficiency. This further research should focus on enlarging the mass share of the most valuable fractions (if the market size allows this), in order to generate as much revenue as possible. Furthermore, characteristics such as self-flocculation capabilities have to be investigated.

11. Conclusion

As part of their CO_2 mitigation goals, Tata Steel and the Schiphol Group commissioned a research to assess the commercial feasibility of microalgae in the Netherlands for the production of AJF. Consequently, this resulted in the following main research question: *To what extent is it commercially feasible to use microalgae for the production of AJF on the short and long term in the Netherlands, while making use of 1Mton of CO_2 from Tata Steel annually?* This main research question is subdivided into four sub questions: 1.) What is the scale of the research, including algae strain, productivity and land occupation? 2.) What are the costs for the cultivation of wet algae biomass in the Netherlands? And what are the main cost drivers? 3.) What are the costs for the algae processing? And what are the main cost drivers? 4.) What is the current and future commercial feasibility of the AJF, considering the total costs, the total revenue (including co-products, AJF and CO_2 permits) and possible future improvements?

As a preliminary research step, the order of magnitude of the research was assessed. The algae strain used in this research are the Nannochloropsis sp. and the Chlorella sp., as these strains are able to survive in both open and closed systems, have a high oil content and can grow in the Dutch environment. The productivity of these algae differs a lot between the cultivation methods, resulting in a productivity of 18.8 to 23.5 Ton/ha/yr for open pond systems, 23.5 to 47.1 Ton/ha/yr for tubular reactors, 42.4 to 78.5 Ton/ha/yr for plate reactors and 23.3 Ton/ha/yr for the OMEGA reactor. Furthermore, a large amount of land is needed: 113.6 to 142.0 km^2 for open pond systems, 85 to 170.4 km^2 for tubular reactors, 51.1 to 94.7 km^2 for plate reactors and 201.2 km^2 for the OMEGA reactor. It became clear that the productivity greatly influenced to amount of land needed. Furthermore, the total land occupation is an important factor as it is unclear if this amount of land is available in the Netherlands.

The costs for cultivation are assessed by first indicating all the different cultivation steps and their design. The different steps are implemented in a final costs model which gives the following values: 3.84 to 17.12 €/kg(DW) for open pond systems, 4.21 to 27.56 €/kg(DW) for tubular reactors, 6.62 to 42.37 €/kg(DW) for plate reactors and 21.83 to 79.99 €/kg(DW) for the OMEGA system. The cost division of the cultivation indicates that the operating costs are the highest for al cultivation methods. However, the MEC and other fixed costs also show a significant influence on the final cultivation costs. The main general costs drivers for cultivation are: land occupation, medium filters, culture medium premix, piping, buildings and maintenance. Furthermore, all cultivation methods show method-specific cost drivers: the open pond reactor has very high costs for a PVC liner inside the ponds; the tubular reactor has very high investment and operating costs for the circulation pump; the plate reactor has very high operating costs for the air blowers (which power the circulation); and the OMEGA method has very high investment costs for the offshore bioreactor.

The costs for processing are assessed by first indicating all the different processing steps and their design. The processing consists of a fractionation and jet refining process in which the extracted lipid fraction is further processed by hydro processing into HEFA (jet fuel). These costs are implemented in an excel model with the following final processing costs: 2.50 to 4.18 €/kg(DW) for open pond systems, 2.60 to 3.81 €/kg(DW) for tubular reactors, 2.56 to 3.52 €/kg(DW) for plate reactors and 2.53 to 3.60 €/kg(DW) for OMEGA systems. The main cost driver for the algae processing are by far the operating costs. Furthermore, the cost division shows that the bead mill energy, spray dry energy, utilities (such as PEG400, Isopropanol, hexane), distillation column for methanol recovery, cooling energy (especially for the bead mill), the labour costs and the jet refinery have the highest costs.

To assess the present and future commercial feasibility, the cultivation and processing costs are added together in a total cost division. Furthermore, the revenue of the different fractions obtained from the algae biomass fractions and the CO_2 permits is calculated and added. All these values are used to calculate the current NPV and PBP of the total algae cultivation and processing system. Almost all values show a negative NPV, which translates to a non-commercially feasible investment. Only the optimistic scenario of the open pond system shows a NPV of 1.60 billion euros and a PBP of 6.54 years (which is sufficient for these large-scale projects). However, the pessimistic scenario of the open pond system shows a NPV of -1.05 billion euros and a PBP of 11.61 years. This is thus a high-risk investment, as it is uncertain whether the optimistic or the pessimistic scenario is more accurate.

The future potential of the AJF is assessed by investigating how much the price of the AJF and the CO_2 permits will have to rise in order to make the project commercially feasible (NPV=0). The values obtained show that the AJF price (480 €/ton) will have to rise to 12,000 €/ton in the most optimistic case, to 869,000 €/ton for the most pessimistic one. Furthermore, the CO_2 permit price (5 €/ton) will have to rise to 295 €/ton for the most optimistic case, to 11,766 €/ton for the most pessimistic one. It is clear that such costs are unlikely to ever be met in the future.

The division of the revenue and the prices of the different fractions indicate that the AJFs have a small share in the total revenue creation and have the lowest value of all the fractions. Besides, the amount of AJF obtained from the algae cultivation and processing is 0.13% to 0.70% of the total amount of jet fuel used at Schiphol annually. Furthermore, this research identifies higher markets for the lipid fraction instead of the fuel market, such as lipid for food/feed (0.95 €/ton), waxes (4 €/kg), sterols (17-45 €/kg) and PUFA (5-75 €/ton). Therefore, companies will probably focus on these other lipid markets instead of the fuel market (including jet) in the future, as they have higher prices per kg. Although, the fuel market size is significantly larger than these high value markets, the fuel market will probably only be used as these higher markets are saturated. When looking at the current lipid fractions obtained from the algae cultivation and processing (8.0 to 44.8 kton) and the amount of lipids needed for this saturation (35 Mton), it is clear that the fuel markets will probably not be utilised in the near future.

As a final assessment, the future potential of the AJF an optimal scenario is identified which consist out of: a higher cost for jet fuel (1200 €/ton), a higher CO_2 permit revenue (120 €/ton) and a 30% cost reduction of both the initial investment and the operating costs. However, these values still indicate a negative NPV for the pessimistic scenario of the tubular reactor (-5.58 billion euros) and all the plate and OMEGA scenarios. Only the optimistic scenario of the tubular reactor (3.14 billion euros) and both the scenarios of the open pond system (5.35 and 2.48 billion euros) indicate a positive NPV. Furthermore, these positive scenarios show a sufficient PBP below 6 years. It can therefore be assumed that it could be profitable to use the open pond system in the future within this optimal and yet very unlikely scenario. The tubular reactor will have too many risks for an investment company, as the pessimistic scenario shows a negative NPV and it is unclear whether the optimistic or pessimistic scenario is more accurate.

Overall it can thus be concluded that it currently not commercially feasible to use microalgae for the production of AJF. In the future, it could be commercially feasible to cultivate algae for the production of AJF in an open pond system if there are very large cost reductions and higher jet fuel and CO_2 permit prices. Even then it is more profitable to focus on other markets with higher values. Therefore, it is very unlikely that the use of microalgae for the production for AJF could be commercially feasible in the future.

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Appendix

Appendix A: Extra information harvest & dewatering methods

Flocculation
Flocculation is a process that aggregates the microalgae cells to enlarge the particle size (Brennan & Owende, 2010). It is a preparatory step which uses chemicals (metal salts) to increase the efficiency of flotation, filtration and gravity sedimentation (Molina Grima et al., 2003). Microalgae generally have a negative charge that prevents aggregation (Molina Grima et al., 2003). Therefore, flocculants such as cationic (group of ions with a positive charge) polymers or multivalent cations are added to neutralize the microalgae and enhance aggregation (Molina Grima et al., 2003). The flocculant selected should have a low toxicity level, be relatively cheap, effective in small quantities and should not affect the downstream processing negatively (Molina Grima et al., 2003). Examples of flocculants are metal salts, polyferric sulfate (PFS), chitosan and cationic polymers (polyelectrolytes) as mentioned before (Molina Grima et al., 2003). However, auto flocculation by modifying the algae production medium, bio flocculant and bacterium are also investigated and proven effective in some cases (Molina Grima et al., 2003).

Flotation
Flotation uses micro-air bubbles to trap algae cells at the surface of the medium, without using chemicals (Wang et al., 2008). However, some microalgae species naturally float due to their lipid content (Brennan & Owende, 2010). With floatation, particles up to 500 micrometres can be captured (Chen et al., 2011). The two general applications of flotation are dissolved air flotation (DAF) and dispersed flotation (Chen et al., 2011; Uduman et al., 2010)

The DAF method is most widely used (Uduman et al., 2010). This method uses a water stream that is saturated with air (which has higher pressure than the atmosphere) and depressurised (Uduman et al., 2010). The saturated stream is injected into a flotation tank through nozzles and a needle valve, which causes an air bubble rise through the liquid to the surface, taking the solids in the floatation tank with them (Uduman et al., 2010). The biomass is thereafter skimmed off the surface (Uduman et al., 2010). The mean bubble size of this process is 40 micrometre, but can range from 10 to 100 micrometre (Uduman et al., 2010). This method is often combined with chemical flocculation and can obtain an algae slurry with 6% solids (Uduman et al., 2010).

Dispersed flotation uses a high speed mechanical agitator and air injection system to form bubbles with a size of 700 to 1500 micrometre (Uduman et al., 2010). Furthermore, this method uses collectors or surfactants to improve the amount of bindings between the algae particles and the air bubbles (Uduman et al., 2010). These collector ions increase the hydrophobicity (water-repellent), which could lead to a electrostatic interaction of the algae biomass and the air bubbles and therefore an increase the yield (Uduman et al., 2010).

Electrophoresis techniques
This method uses an electric field that electrolyses water, producing hydrogen which is attached to micro algae flocs and separated from the solution (Chen et al., 2011). The separation is made by the hydrogen that brings the flocs to the surface of the medium (Mollah et al., 2004). This method is versatile, energy efficient, safe, select, environmental compatible and cost effective (Mollah et al., 2004). Other electro techniques are coagulation by metal ions, produced from reactive electrodes and electrolytic flocculation that moves negatively charged microalgae to the anode to separate them from the medium (Uduman et al., 2010).

Sedimentation
Gravity and centrifugal sedimentation is based on Stake's Law: "settling characteristics of suspended solids is determined by density and radius of algae cells (Stoke's radius) and sedimentation velocity" (Brennan & Owende, 2010, p. 567). This method is generally used for wastewater treatment due to large volumes and low biomass concentration (Brennan & Owende, 2010). Furthermore, it is only possible to use this method for large (> 70 micrometre) microalgae strains (Muñoz & Guieysse, 2006). The sedimentation process is cheap but unreliable (Uduman et al., 2010). Often flocculation is used to improve the sedimentation/separation rate and thus the reliability of the process (Uduman et al., 2010). The two general system designs to sediment algae are lamella separators and sedimentation tanks (Uduman et al., 2010).

Centrifugation
Centrifugation is generally used for high value products used in aquaculture (Heasman et al., 2000). However, most algae can be harvested by this method (Molina Grima et al., 2003). It is rapid and energy intensive method that has potential high maintenance costs due to moving parts (Brennan & Owende, 2010). However, centrifugation has an harvesting efficiency of over 95% and increases the solid content of the algae slurry by 150 times to around 15% (Brennan & Owende, 2010) to 20% (Davis et al., 2016). Centrifugation is a preferred method to obtain algae cells, especially for aquaculture purposes as this method extends the durability (Molina Grima et al., 2003). The main factors that influence the recovery of algae biomass in a centrifuge are the characteristics of the algae cells, the residence time and the settling depth of the centrifuge (Molina Grima et al., 2003). The settling depth is dependent of the centrifuge design and the residence time can be controlled by the flow rate (Molina Grima et al., 2003).

Ultrasonic aggregation
This method uses ultrasonic sound to aggregate the micro algae, thereafter the biomass is enhanced by sedimentation (Brennan & Owende, 2010). A test that used this process showed a separation efficiency of 92% and a 20 times more concentrated solution than the starting point (Brennan & Owende, 2010). The benefits of this process are that it can be used continuously without stressing the biomass (Brennan & Owende, 2010). This method is not yet widely used (Brennan & Owende, 2010).

Filtration
Finally, conventional filtration can be used for relatively large (> 70 micrometre) microalgae and cannot be used to harvest species below 30 micrometre (like chlorella) (Heasman et al., 2000). For small particles (<30 micrometre) ultra-filtration is a technical viable alternative (Brennan & Owende, 2010; Petruševski et al., 1995). Generally, filtration uses pressure or suction to improve efficiency (Brennan & Owende, 2010). Filtration is able to increase the concentration 245 times, with a 27% solids result (Brennan & Owende, 2010). In general membrane filtration is more cost effective compared to centrifugation when operating with small scales (<2m^3 per day) and centrifugation is considered more cost effective when working with larger production scales (>20m^3 per day), due to membrane replacement and pumping costs for the filtration option (Brennan & Owende, 2010). At last, a filter aid (diatomaceous earth or cellulose) is often used during the filtration process, which contaminates the algae mass and makes it very hard to extract intracellular products (such as oils in this case) (Molina Grima et al., 2003).

Appendix B: Overview of assumptions for calculation biomass production

Table 23: General assumptions cultivation methods

Factor	Open pond	Tubular	Plate	Omega	Unit	Source
Photosynthetic efficiency	Dependent on scenario	Dependent on scenario	Dependent on scenario	Dependent on scenario	%	(Norsker et al., 2011; Ruiz et al., 2016), calculation chapter 4.1.2
Enthalpy of combustion algae	0.022	0.022	0.022	0.022	MJ/(g(DW))	(Dillschneider et al., 2013; Ruiz et al., 2016; Weyer et al., 2009)
Average growth season irradiance	12.5	12.5	12.5	10.66	MJ/(m²*day)	(Allesoverzonnepanelen.nl, 2012; Gemiddeldgezien.nl, 2017; KNMI, 2017), calculation chapter 4.1.2
Number of days per growth season	275	275	275	337	Days	Calculation chapter 4.1.2
CO_2 fixation rate	1.87	1.87	1.87	1.87	$kgCO_2$/kg(DW)	(Hulst, 2012; Zhang, 2015)
CO_2 input	1	1	1	1	Mton	(Louwerse & Koelemeijer, 2016)
CO_2 uptake efficiency cultivation method	40	60	60	50	%	(Sobczuk et al., 2000; Trent, 2013; Zhang, 2015)
Total land area multiply factor	1.25	1.25	1.25	1.43	*irradiated area	(Norsker et al., 2011), calculation chapter 4.1.2
Lifetime cultivation system	25	25	25	25	Years	(Koelemeijer, van der Stel & Louwerse, personal communication, 2 March, 2017)
Discount rate	10	10	10	10	%	(Blok, 2006; Blok & Nieuwlaar, 2016)
Annuity factor	0.11	0.11	0.11	0.11		See formula in the theory
General biomass concentration	0.32	1.7	2.01	1.7	Kg/m3	(Norsker et al., 2011)
Total land area irradiated	Dependent on scenario	Dependent on scenario	Dependent on scenario	Dependent on scenario	Km^2	Calculation see chapter 4.1.3
Total land area needed	Dependent on scenario	Dependent on scenario	Dependent on scenario	Dependent on scenario	Km^2	Calculation see chapter 4.1.3
Total mitigated CO_2	400,000	600,000	600,000	500,000	Ton/yr	Calculation see excel
Biomass concentration when moving to sedimentation tank	0.32	1.7	2.01	3.4	Kg/m3	Same as general biomass concentration but OMEGA method has an internal sedimentation unit which doubles the concentration (Trent, 2013)
EURO per dollar	0.932	0.932	0.932	0.932	Euro/dollar	Picked at 27-03-2017
CO_2 pipeline width	500	500	500	500	mm	Assumed from (Grant, Morgan, & Gerdes, 2014; Noothout et al., 2014)
Sludge pipeline width from offshore algae plant	/	/	/	500	mm	Assumed to be the same as the CO_2 pipeline to give an indication
Evaporation heat water	2257	2257	2257	2257	Kj/kg	(The Engineering ToolBox, n.d.)

General Dutch electricity price	0.066	0.066	0.066	0.066	Euro/kwh	(Renzenbrink, 2014)

Table 24: MEC assumptions

Factor	**Open pond**	**Tubular**	**Plate**	**Omega**	**Unit**	**Source**
PVC liner (1044 m2/ pond)	7.15	/	/	/	Euro/m²	(Norsker et al., 2011)
Circulation pump (1000 m³/h)	/	26,100	/	/	Euro/unit	(Norsker et al., 2011)
Steel frame (88 ton/ha)	/	/	650	/	Euro/ton	(Norsker et al., 2011)
Paddle wheel (1 unit/pond)	800	/	/	/	Euro/unit	(Norsker et al., 2011)
Air blowers tubular reactor (0.03 unit/ha)	/	105,000	/	/	Euro/unit	(Norsker et al., 2011)
Air blowers plate reactor (0.32 unit/ha)	/	/	106,000	/	Euro/unit	(Norsker et al., 2011)
Total bioreactor including GEHC, pontoons etc.	/	/	/	1.19	Mil€/ha	(Trent, 2013)
Onshore CO_2 pipeline	100,000	100,000	100,000	100,000	USD/(inch diameter*mile)	(Grant et al., 2014; Noothout et al., 2014)
Offshore CO_2 pipeline	/	/	/	700,000	USD/(inch diameter*mile)	(Grant et al., 2014; Noothout et al., 2014)
CO_2 costs	0	0	0	0	Euro/ton	CO_2 costs are considered 0 as they are delivered by Tata Steel
Water costs	0	0	0	0	Euro/ton	Water costs are considered 0 as seawater does not cost anything and in the freshwater case the water is recycled so has a small impact
Sludge pipeline from offshore algae plant	/	/	/	Same costs for on and offshore CO_2 pipeline	USD/(inch diameter*mile)	Assumed to be the same as (Grant et al., 2014; Noothout et al., 2014) to give an indication
Centrifuge Westfalia separator AG (60 m3/h)	500,000	500,000	500,000	500,000	Euro/unit	(Norsker et al., 2011)
Centrifuge feed pump (80 m3/h)	8,000	8,000	8,000	8,000	Euro/unit	(Norsker et al., 2011)
Medium filter unit (6 m3/h)	13,500	13,500	13,500	13,500	Euro/unit	(Norsker et al., 2011)
Medium Feed pump (84,600 m3/h)	15,000,000	15,000,000	15,000,000	15,000,000	Euro/unit	(Lammerink, R., personal communication, 16 March, 2017). Lammerink is sales engineer at Flowserve.
Medium preparation tank. concrete type D-115	0.43 (for first 2million gallons) – 0.28 (for	0.43 (for first 2million gallons) – 0.28 (for	0.43 (for first 2million gallons) – 0.28 (for	0.43 (for first 2million gallons) – 0.28 (for	USD/gallon	(Rettew, n.d.)

(12h inflow holding capacity)	every extra gallon)	every extra gallon)	every extra gallon)	every extra gallon)		
Gravity broth sedimentation tank (94.6 MillionL/day)	3,384,194	3,384,194	3,384,194	3,384,194	Euro/unit	(Wastewater Innovations, 2014)
Water pump station (84,600 m^3/h)	15,000,000	15,000,000	15,000,000	15,000,000	Euro/unit	(Lammerink, R., personal communication, 16 March, 2017)
Weighting station 64,000 ton(DW)/(unit*yr)	40,000	40,000	40,000	40,000	Euro/unit	Based on (Norsker et al., 2011)

Table 25: General operation and maintenance costs assumptions

Factor	**Open pond**	**Tubular**	**Plate**	**Omega**	**Unit**	**Source**
Labour (0.07 people/ha)	37,100	37,100	37,100	37,100	Euro/person	(Norsker et al., 2011)
Payroll charges	25	25	25	25	% of salary	(Norsker et al., 2011)
Maintenance	2	4	4	8	% of total MEC	(Norsker et al., 2011), OMEGA maintenance costs were based on (Milborrow, 2016), which indicates that offshore windfarms have roughly double the maintenance costs of onshore methods
General plant overheads	55	55	55	55	% of salary + maintenance	(Norsker et al., 2011)
Power for cell harvest (in centrifuge) (1.1 kwh/m3)	5,490,196	4,941,176	8,235,294	7,064,385	Kwh/yr	(Norsker et al., 2011), see excel
Power requirement for paddle wheel (0.208 KW/(ha*h))	14.880.242	/	/	/	Kwh/yr	(Norsker et al., 2011), see excel
Culture medium premix (0.44 Euro/kg(DW) algae)	214	321	321	328	Kton premix	(Norsker et al., 2011), see excel
Medium filter cartridges	132	109	126	109	Cartridges/ (ha*yr)	(Norsker et al., 2011), for OMEGA the same amount of cartridges as the tubular reactor is assumed
CO_2 pipeline pump costs	Dependent on scenario	Dependent on scenario	Dependent on scenario	Dependent on scenario	See table 8	
Flocculation costs (0.00012 kg/l)	0.64	0.64	0.64	0.64	Euro/kg	(Rakesh et al., 2013)
Polyethylene liner tube (0.03 euro/m)	/	175438	/	/	m/(ha*yr)	(Norsker et al., 2011), see excel
Culture circulation pump power (for tubular design) (0.005 kwh/m^3)	/	8.07*10^11	/	/	M^3/yr	(Norsker et al., 2011), see excel
Blower power tubular reactor (55 m3/kwh)	/	636	/	/	Gwh/yr	(Norsker et al., 2011), see excel

Polyethylene film for plate reactor (38,500 m^2/ha)	/	/	0.168	/	Euro/m2	(Norsker et al., 2011), see excel
Air blowers power (110 m3/kwh)	/	/	28.1*10^9	/	Kwh/yr	(Norsker et al., 2011), see excel
Cooling power needed (1 kwh/kg(DW) algae)	/	/	321	/	Gwh/yr	(Girdhari, 2011), see excel
Polyethylene liner tube OMEGA (0.03 euro/m)	/	/	/	834,378	Km tube	(Norsker et al., 2011; Trent, 2013), see excel
PBR pump OMEGA (0.17 kj/l)	/	/	/	5,598	Gwh/yr	(Trent, 2013), see excel
GEHC pump OMEGA (0.2551 kj/l)	/	/	/	13	Gwh/yr	(Trent, 2013), see excel
Sludge pipeline back to onshore for processing (OMEGA)	/	/	/	Assumed same pump energy of CO_2 pipeline but based on scenario	See table 8	
Cell harvest power in harvesting column (OMEGA) (0.05 kj/l)	/	/	/	2.7	Gwh/yr	(Trent, 2013), see excel

Table 26: Other fixed costs

Factor	**Open pond**	**Tubular**	**Plate**	**Omega**	**Unit**	**Source**
Installation	15	30	30	30	% MEC	(Norsker et al., 2011), for omega assumed the same as tubular
Instrumentation	10	10	10	10	% MEC	(Norsker et al., 2011)
Piping	30	30	30	30	% MEC	(Norsker et al., 2011)
Buildings	30	30	30	30	% MEC	(Norsker et al., 2011)
Land	6	6	6	/	€/m^2	(Boerderij.nl, n.d.), rough estimate of area around Ijmuiden

Appendix C: Cost division of biomass production for each of the cultivation methods

Table 27: Cultivation cost division of raceway open pond

MEC	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
PVC liner	€ 74,729,552	37.53	€ 112,094,328	204.70
Centrifuge Westfalia separator AG	€ 440,672	0.22	€ 1,211,849	2.21
Centrifuge feed pump	€ 5,288	0.00	€ 14,983	0.03
Medium filter unit	€ 25,106,588	12.61	€ 25,106,588	45.85
Medium feed pump	€ 3,956,682	1.99	€ 3,956,682	7.23
Medium preparation tank	€ 9,258,534	4.65	€ 9,258,534	16.91
Gravity broth sedimentation tank	€ 9,579,781	4.81	€ 9,579,781	17.49
Water pump station	€ 1,978,341	0.99	€ 1,978,341	3.61
Automatic Weighing station with silos	€ 141,015	0.07	€ 39,661	0.07
Paddle wheel	€ 8,008,954	4.02	€ 12,013,432	21.94
CO2 pipeline/feed pipeline	€ 2,561,408	1.29	€ 2,561,408	4.68
Land	€ 75,083,708	37.70	€ 93,854,635	171.39
Total	**€ 210,850,522**	**105.88**	**€ 271,670,219**	**496.12**
Operating Costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Labour	€ 29,531,600	14.83	€ 36,877,400	67.34
Payroll charges	€ 7,382,900	3.71	€ 9,219,350	16.84
Maintenance	€ 38,277,973	19.22	€ 49,319,229	90.07
General plant overheads	€ 37,295,265	18.73	€ 47,408,146	86.58
Power for cell harvest	€ 217,412	0.11	€ 621,176	1.13
Paddle wheel power	€ 824,661	0.41	€ 1,030,826	1.88
Culture medium premix	€ 94,117,647	47.26	€ 94,117,647	171.88
Medium filters	€ 113,953,124	57.22	€ 142,441,405	260.12
CO2 pipeline pump	€ 697,500	0.35	€ 1,860,000	3.40
Flocculation	€ 52,139,037	26.18	€ 52,139,037	95.21
Total	**€ 374,437,118**	**188.02**	**€ 435,034,217**	**794.45**
Other Fixed costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Installation	€ 31,627,578	15.88	€ 40,750,533	74.42
Instrumentation	€ 21,085,052	10.59	€ 27,167,022	49.61
Piping	€ 63,255,157	31.76	€ 81,501,066	148.83
Buildings	€ 63,255,157	31.76	€ 81,501,066	148.83
Total	**€ 179,222,944**	**90.00**	**€ 230,919,686**	**421.70**
Overall total costs	**€ 764,510,585**	**383.90**	**€ 937,624,122**	**1712.26**
Final costs per kg(DW)		**€ 3.84**		**€ 17.12**
Total biomass production	**199144385**	**kg(DW)/yr**	**54759358**	**kg(DW)/yr**

Table 28: Cultivation cost division of tubular reactor

MEC	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Circulation pump	€ 177,086,440	59.28	€ 354,170,004	431.18
Centrifuge Westfalia separator AG	€ 661,008	0.22	€ 1,817,773	2.21
Centrifuge feed pump	€ 7,932	0.00	€ 22,034	0.03
Medium Filter Unit	€ 7,101,709	2.38	€ 7,101,709	8.65
Medium Feed pump	€ 1,117,181	0.37	€ 1,117,181	1.36
Medium preparation tank	€ 2,636,274	0.88	€ 2,636,274	3.21
Gravity broth sedimentation tank	€ 2,704,879	0.91	€ 2,704,879	3.29
Water pump station	€ 558,590	0.19	€ 558,590	0.68
Weighting station	€ 220,336	0.07	€ 61,694	0.08
Air blowers	€ 19,850,083	6.65	€ 19,850,083	24.17
CO2 pipeline/feed pipeline	€ 2,561,408	0.86	€ 2,561,408	3.12
Land	€ 56,312,781	18.85	€ 112,625,562	137.12
Total	**€ 270,818,622**	**90.66**	**€ 505,227,191**	**615.09**
Operating Costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Labour	€ 22,148,700	7.41	€ 44,260,300	53.88
Payroll charges	€ 5,537,175	1.85	€ 11,065,075	13.47
Maintenance	€ 98,329,259	32.92	€ 183,438,697	223.33
General plant overheads	€ 66,262,877	22.18	€ 125,234,449	152.47
Polyethylene liner (annual replacement)	€ 35,870,290	12.01	€ 71,740,579	87.34
Culture circulation pump power	€ 223,558,036	74.84	€ 447,116,073	544.34
Blower power	€ 35,265,493	11.81	€ 70,530,986	85.87
Cell harvest power	€ 326,118	0.11	€ 931,765	1.13
Culture medium premix	€ 141,176,471	47.26	€ 141,176,471	171.88
Medium filters	€ 70,573,241	23.63	€ 141,146,483	171.84
CO2 pipeline pump	€ 697,500	0.23	€ 1,860,000	2.26
Flocculation costs	€ 14,721,611	4.93	€ 14,721,611	17.92
Total	**€ 714,466,770**	**239.18**	**€ 1,253,222,488**	**1525.73**
Other Fixed costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Installation	€ 81,245,586	27.20	€ 151,568,157	184.53
Instrumentation	€ 27,081,862	9.07	€ 50,522,719	61.51
Piping	€ 81,245,586	27.20	€ 151,568,157	184.53
Buildings	€ 81,245,586	27.20	€ 151,568,157	184.53
Total	**€ 270,818,622**	**90.66**	**€ 505,227,191**	**615.09**
Overall total costs	**€ 1,256,104,013**	**420.50**	**€ 2,263,676,870**	**2755.91**
Final costs per kg(DW)		**€ 4.21**		**€ 27.56**
Total biomass production	**298716578**	**kg(DW)/yr**	**82139037**	**kg(DW)/yr**

Table 29: Cultivation cost division of plate reactor

MEC	Optimistic	€cts/kg	Pessimistic	€cts/kg
Steel frame	€ 25,768,731	8.63	€ 47,719,901	58.10
Centrifuge Westfalia separator AG	€ 661,008	0.22	€ 1,817,773	2.21
Centrifuge feed pump	€ 7,932	0.00	€ 22,034	0.03
Medium Filter Unit	€ 5,995,259	2.01	€ 5,995,259	7.30
Medium Feed pump	€ 944,879	0.32	€ 944,879	1.15
Medium preparation tank	€ 2,234,433	0.75	€ 2,234,433	2.72
Gravity broth sedimentation tank	€ 2,287,709	0.77	€ 2,287,709	2.79
Water pump station	€ 472,440	0.16	€ 472,440	0.58
Weighting station	€ 211,523	0.07	€ 61,694	0.08
Air blowers	€ 15,286,261	5.12	€ 28,307,025	34.46
CO2 pipeline/feed pipeline	€ 2,561,408	0.86	€ 2,561,408	3.12
Land	€ 33,787,669	11.31	€ 62,569,757	76.18
Total	**€ 56,431,582**	**18.89**	**€ 92,424,554**	**112.52**
Operating Costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Labour	€ 13,281,800	4.45	€ 24,597,300	29.95
Payroll charges	€ 1,992,270	0.67	€ 3,689,595	4.49
Maintenance	€ 20,489,269	6.86	€ 33,557,655	40.85
General plant overheads	€ 18,574,088	6.22	€ 31,985,225	38.94
Polyethylene film for reactor	€ 26,375,364	8.83	€ 48,843,267	59.46
Air blowers power	€ 1,557,259,400	521.32	€ 2,883,813,704	3510.89
Power for cell harvest	€ 326,118	0.11	€ 931,765	1.13
Culture medium premix	€ 141,176,471	47.26	€ 141,176,471	171.88
Medium filters	€ 48,948,046	16.39	€ 90,644,530	110.35
CO2 pipeline pump costs	€ 697,500	0.23	€ 1,860,000	2.26
Cooling energy needed	€ 21,176,471	7.09	€ 21,176,471	25.78
Flocculation costs	€ 12,451,113	4.17	€ 12,451,113	15.16
Total	**€ 1,862,747,910**	**623.58**	**€ 3,294,727,096**	**4011.16**
Other Fixed costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Installation	€ 16,929,475	5.67	€ 27,727,366	33.76
Instrumentation and control	€ 5,643,158	1.89	€ 9,242,455	11.25
Piping	€ 16,929,475	5.67	€ 27,727,366	33.76
Buildings	€ 16,929,475	5.67	€ 27,727,366	33.76
Total	**€ 56,431,582**	**18.89**	**€ 92,424,554**	**112.52**
Overall total costs	**€ 1,975,611,074**	**661.37**	**€ 3,479,576,205**	**4236.20**
Final costs per kg(DW)		**€ 6.61**		**€ 42.36**
Total biomass production	**298716578**	**kg(DW)/yr**	**82139037**	**kg(DW)/yr**

Table 30: Cultivation cost division of OMEGA reactor

MEC	Optimistic	€cts/kg	Pessimistic	€cts/kg
Centrifuge Westfalia separator AG	€ 771,177	0.25	€ 1,542,353	1.84
Centrifuge feed pump	€ 8,813	0.00	€ 18,508	0.02
Medium Filter Unit	€ 5,907,432	1.94	€ 5,907,432	7.04
Medium Feed pump	€ 232,746	0.08	€ 232,746	0.28
Medium preparation tank	€ 1,116,410	0.37	€ 1,116,410	1.33
Gravity broth sedimentation tank	€ 1,127,033	0.37	€ 1,127,033	1.34
Water pump station	€ 1,652,521	0.54	€ 1,652,521	1.97
Weighting station	€ 215,929	0.07	€ 61,694	0.07
Total bioreactor including GEHC pontoons etc.	€ 1,844,541,185	604.66	€ 1,844,541,185	2198.99
CO2 pipeline/feed pipeline	€ 22,668,458	7.43	€ 22,668,458	27.02
Sludge pipeline back to shore	€ 22,668,458	7.43	€ 22,668,458	27.02
Total	**€ 1,900,910,163**	**623.14**	**€ 1,901,536,799**	**2266.94**
Operating Costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Labour	€ 52,273,900	17.14	€ 52,273,900	62.32
Payroll charges	€ 13,068,475	4.28	€ 13,068,475	15.58
Maintenance	€ 1,363,910,010	447.11	€ 1,380,826,050	1646.17
General plant overheads	€ 778,901,150	255.33	€ 788,204,972	939.67
Polyethylene liner tube	€ 21,101,029	6.92	€ 21,101,029	25.16
PBR pump energy	€ 311,442,804	102.09	€ 311,442,804	371.29
GEHC pump energy	€ 886,935	0.29	€ 886,935	1.06
Cell harvest power (INGEHC)	€ 133,849	0.04	€ 133,849	0.16
Culture medium premix	€ 144,171,123	47.26	€ 144,171,123	171.88
Medium filters	€ 145,828,175	47.80	€ 145,828,175	173.85
Cell harvest power	€ 466,249	0.15	€ 951,529	1.13
CO2 pipeline pump costs	€ 6,172,875	2.02	€ 16,461,000	19.62
Sludge pipeline power	€ 6,172,875	2.02	€ 16,461,000	19.62
Flocculation costs	€ 15,033,887	4.93	€ 15,033,887	17.92
Total	**€ 2,859,563,337**	**937.40**	**€ 2,906,844,729**	**3465.42**
Other Fixed costs	**Optimistic**	**€cts/kg**	**Pessimistic**	**€cts/kg**
Installation	€ 570,273,049	186.94	€ 570,461,040	680.08
Instrumentation and control	€ 190,091,016	62.31	€ 190,153,680	226.69
Piping	€ 570,273,049	186.94	€ 570,461,040	680.08
Buildings	€ 570,273,049	186.94	€ 570,461,040	680.08
Total	**€ 1,900,910,163**	**623.14**	**€ 1,901,536,799**	**€ 2,266.94**
Overall total costs	**€ 6,661,383,664**	**2183.68**	**€ 6,709,918,328**	**€ 7,999.29**
Final costs per kg(DW)		**€ 21.84**		**€ 79.99**
Total biomass production	**305052990**	**kg(DW)/yr**	**83881381**	**kg(DW)/yr**

Appendix D: Overview of assumptions for calculation biomass processing

Table 31: General assumptions biomass processing

Factor	Open pond	Tubular	Plate	Omega	Unit	Source
Cooling water (15 degrees Celsius)	0.037	0.037	0.037	0.037	Euro/ton	(Ruiz et al., 2016)
NaCl brine (-10 degrees Celsius)	0.185	0.185	0.185	0.185	Euro/ton	(Ruiz et al., 2016)
Formula's used for cooling calculation	$Mass\ flow\ of\ cooling\ water = \frac{energy\ to\ remove\ from\ the\ culture}{C_P \cdot \Delta T \cdot \rho_{seawater}}$ $\Delta T = T_{cooling\ water\ out} - T_{cooling\ water\ in}$ $T_{cooling\ water\ out} = T_{cooling\ water\ in} + Efficiency \cdot (\frac{T_{culture}}{T_{cooling\ water\ in}})$					(Ruiz et al., 2016)
Efficiency heat transfer	75	75	75	75	%	(Ruiz et al., 2016)
C_p	4186	4186	4186	4186	J/(kg*degree Celsius)	(Ruiz et al., 2016)
$\rho_{seawater}$	1027	1027	1027	1027	(kg/m³)	(Ruiz et al., 2016)
Density lipid fraction	900	900	900	900	Kg/m3	(Noureddini et al., 1992)
Cost rest heat	4	4	4	4	Euro/GJ	(Rooijers, 2002)
General Dutch electricity price	0.066	0.066	0.066	0.066	Euro/kwh	(Renzenbrink, 2014)
Enthalpy of vaporization Hexane (at 200 degrees Celsius)	15.7	15.7	15.7	15.7	KJ/mol	(Chickos & Acree, 2003)
Enthalpy of vaporization Isopropanol (at 200 degrees Celsius	16.5	16.5	16.5	16.5	KJ/mol	(Wormald & Vine, 2000)
Enthalpy of vaporization Methanol (at 150 degrees Celsius)	876	876	876	876	kj/kg	(Yerlett & Wormald, 1986)
Mol mass Hexane	86.17	86.17	86.17	86.17	g/mol	(ConvertUnits.com, 2017b)
Mol mass Isopropanol	60.10	60.10	60.10	60.10	g/mol	(ConvertUnits.com, 2017a)
Density lipid fraction	900	900	900	900	Kg/m³	(Noureddini et al., 1992)
Bead mill disruption efficiency	95	95	95	95	%	For all cells, so only 95% of all material can be subtracted (Ruiz et al., 2016)
Lipid extraction efficiency	85	85	85	85	%	(Ruiz et al., 2016)
Cell dry weight concentration after ultrafiltration unit (II.3- II.4)	27	27	27	27	%	(Brennan & Owende, 2010)
Lifetime of all units (if not indicated differently)	25	25	25	25	Years	(Koelemeijer, van der Stel & Louwerse, personal communication, 2 March, 2017)

Output from lipid refinery (for max kerosene production)	Jet fuel	47.5	%	(M. Pearlson et al., 2013), all the products are considered valuable products except for carbon dioxide and water.
	Diesel	22.4	%	
	Naphtha	6.7	%	
	Propane	4.0	%	
	LPG	5.8	%	
	Carbon Dioxide	5.2	%	
	Water	8.4	%	

Table 32:MEC biomass processing assumptions

Factor	Open pond	Tubular	Plate	Omega	Unit	Source
Bead milling capacity	2	2	2	2	M^3	(Ruiz et al., 2016)
Bead milling residence time	5	5	5	5	min	(Lee, Lewis, & Ashman, 2012)
Bead milling price	600,000	600,000	600,000	600,000	Euro/unit	(Ruiz et al., 2016)
ATPS mixer capacity (II.1-II.2)	9	9	9	9	M^3	(Ruiz et al., 2016)
ATPS settlers capacity (II.1-II.2)	50	50	50	50	M^3	(Ruiz et al., 2016)
Residence time mixer	5	5	5	5	Min	(Ruiz et al., 2016)
Residence time settler	30	30	30	30	Min	(Ruiz et al., 2016)
ATPS mixer-settlers costs (II.1-II.2)	400,000	400,000	400,000	400,000	Euro/unit	(Ruiz et al., 2016)
Ultrafiltration unit flow rate (II.3-II.4)	0.5	0.5	0.5	0.5	$L/(m^2*h)$	(Millipore Corporation, 2008)
Ultrafiltration price (for a 400 m2 unit)	240,000	240,000	240,000	240,000	Euro/unit	(Ruiz et al., 2016)
Diafiltration flow rate (II.7-II.8)	40	40	40	40	$L/(m^2*h)$	(Ruiz et al., 2016)
Diafiltration price (for an 80 m2 unit)	120,000	120,000	120,000	120,000	Euro/unit	(Ruiz et al., 2016)
Spray driers (II.5. II.6. II.9. II.10 and IV.2)	100	100	100	100	$Kg/(h*m^3)$	(Ruiz et al., 2016)
Spray driers capacity	60	60	60	60	M^3	(Ruiz et al., 2016)
Spray drier price	230,000	230,000	230,000	230,000	Euro/unit	(Ruiz et al., 2016)

Lipid extractor mixer capacity (III.1)	2	2	2	2	M^3	(Ruiz et al., 2016)
Lipid extractor settler capacity (III.1)	12	12	12	12	M^3	(Ruiz et al., 2016)
Residence time mixer	5	5	5	5	Min	(Ruiz et al., 2016)
Residence time settler	30	30	30	30	Min	(Ruiz et al., 2016)
Lipid extractor mixer-settler costs	160,000	160,000	160,000	160,000	Euro/unit	(Ruiz et al., 2016)
Distillation column (III.2)	50,000	50,000	50,000	50,000	Euro/unit	(Ruiz et al., 2016), assumed that only one distillation column is used
Decanter+cooler capacity decanter (III.3)	30	30	30	30	M^3	(Ruiz et al., 2016)
Residence time	5	5	5	5	hours	(Ruiz et al., 2016)
Decanter+cooler costs	40,000	40,000	40,000	40,000	Euro/unit	(Ruiz et al., 2016)
Evaporator+ condenser for methanol recovery (60 m^3 flash)	4,500	4,500	4,500	4,500	Euro/unit	(Ruiz et al., 2016), costs are very low due to relative small lipid fraction, so only one unit is assumed
Diafiltration unit (IV.1) (same flow rate as II.7-II.8) capacity	75	75	75	75	M^2	(Ruiz et al., 2016)
Diafiltration unit costs (IV.1)	120,000	120,000	120,000	120,000	Euro/unit	(Ruiz et al., 2016)

Table 33: Operation costs biomass processing

Factor	**Open pond**	**Tubular**	**Plate**	**Omega**	**Unit**	**Source**
Bead mill energy	700	700	700	700	Kwh/m^3	Based on (Lee et al., 2012)
ATPS mixer-settlers energy (II.1 – II.2)	0.5	0.5	0.5	0.5	Kw/m^3 (multiplied with number of hours per year)	(Ruiz et al., 2016), the amount of energy used in II.2 is half the amount used in II.1 due to the volumetric partitioning of the water in the top and bottom phase with a ratio of 1:1.
PEG400 (26% weight fraction II.1)	95	95	95	95	% of protein fraction is subtracted by PEG 400 and thus used in the process	(Ruiz et al., 2016)
PEG400 costs	1.35	1.35	1.35	1.35	Euro/kg	(Ruiz et al., 2016)
PEG400 recycle efficiency	91	91	91	91	%	(Ruiz et al., 2016)
PEG400 loss	9.1	9.1	9.1	9.1	%	Partitioning fraction is 10, so bottom phase which is lost is 1/11 and top phase which is recovered is 10/11 by recycling,

						obtaining a 10:1 partitioning fraction (Ruiz et al., 2016)
Sodium phosphate	0.38	0.38	0.38	0.38	Euro/kg	(Ruiz et al., 2016)
Phosphate recycle efficiency II.1	80	80	80	80	%	(Ruiz et al., 2016), the amount bought is the 20% loss, to maintain the 15% weight level.
Phosphate recycle efficiency II.2	95	95	95	95	%	(Ruiz et al., 2016), higher recycle efficiency due to higher level of phosphate in mixture (105% weight level) in II.2. Again the 5% loss is bought to maintain the weight level.
Energy for ultrafiltration (II.3-II.4)	3.57	3.57	3.57	3.57	Kwh/m^3	(Glueckstern & Priel, n.d.), energy usage is the same for both ultrafiltration units
Ultrafiltration membrane replacement (II.3-II.4)	5	5	5	5	Years lifetime	(Scholz, 2015)
Ultrafiltration membrane replacement costs (II.3-II.4)	30	30	30	30	Euro/m^2	(Glueckstern & Priel, n.d.)
Diafiltration membrane replacement costs (II.7-II.8)	30	30	30	30	Euro/m^2	Assumed the same as ultrafiltration membrane costs
Diafiltration membrane replacement (II.7-II.8)	1000	1000	1000	1000	hours	(Ruiz et al., 2016)
Energy used for diafiltration (II.7-II.8)	0.2	0.2	0.2	0.2	Kw/m^2	(Ruiz et al., 2016), can be calculated by looking at the total filter area used at the MEC section
PBS buffer usage (II.8)	1.51	1.51	1.51	1.51	Euro/kg	(Ruiz et al., 2016), assumed to react with monosaccharides 1:1
Spray drier (all together) absorbed energy	0.02	0.02	0.02	0.02	Kw/kg of feed	(Ruiz et al., 2016), assumed is that the full mass entering the biorefinery is put through the spray driers. The spray driers are thus calculated all together.
Spray drier heat transfer efficiency	50	50	50	50	%	(McCabe et al., 1993; Wisniewski, 2015), average assumption based on low efficiency and efficiencies from other papers
Spray drier heat price	Tata steal has rest heat which could be used for the drying of the algae. the amount of waste heat available is 7.14 PJ (Louwerse, G., personal communication, 16 February, 2017). This rest heat is bought for 4 €/GJ (Rooijers, 2002). all the heat above 7.14 PJ is bought at the normal price of 0.066 €/kwh (Renzenbrink, 2014) or 18.33 €/GJ .					(Renzenbrink, 2014; Rooijers, 2002), Louwerse, G., personal communication, 16 February, 2017.
Lipid extractor mixer-settler (same as II.1-II.2)	0.5	0.5	0.5	0.5	Kw/m^3	(Ruiz et al., 2016)
Hexane-isopropanol volume amount compared to water	10	10	10	10	Times higher	(Ruiz et al., 2016)
Recycle rate Hexane used for oil extraction (20% hexane in the hexane-isopropanol mixture)	80	80	80	80	% recycled	(Ruiz et al., 2016)

Hexane costs	0.93	0.93	0.93	0.93	Euro/kg	(Ruiz et al., 2016), the 20% loss is bought again every year
Recycle rate Isopropanol used for oil extraction (80% isopropanol in the hexane-isopropanol mixture)	80	80	80	80	% recycled	(Ruiz et al., 2016)
Isopropanol costs	1.34	1.34	1.34	1.34	Euro/kg	(Ruiz et al., 2016)
Distillation column energy (III.2)	$Energy\ Distillation\ column = Enthalpy\ of\ vaporization\ hexane * total\ amount\ of\ hexane\ in\ lipid\ fraction + Enthalpy\ of\ vaporization\ isopropanol * total\ amount\ of\ isopropanol\ in\ lipid\ fraction$					Efficiency of the system is 80%, so 80% of the isopropanol and hexane are evaporated and recovered
Distillation column energy price (III.2)	0.066	0.066	0.066	0.066	Euro/kwh	(Renzenbrink, 2014), this price is used instead of the rest heat price as the total amount of rest heat is used in the spray driers
Decanter + Cooler energy (III.3)	$Amount\ of\ energy\ needed\ for\ cooling = Specific\ heat\ lipid\ fraction\ (in\ Kj/(kg * K)) * total\ annual\ amount\ of\ lipid\ fraction\ (in\ kg) * (Temp\ culture\ in - Temp\ culture\ Out)$ $Mass\ flow\ of\ NaCl\ brine = \frac{energy\ to\ remove\ from\ the\ culture}{C_P \cdot \Delta T \cdot \rho_{seawater}}$ $\Delta T = T_{NaCl\ Brine\ out} - T_{NaCl\ brine\ in}$ $T_{cooling\ water\ out} = T_{cooling\ water\ in} + Efficiency \cdot (\frac{T_{culture\ out}}{T_{NaCl\ brine\ in}})$ $T_{NaCl\ in}$= -10 degrees Celsius $T_{culture\ in}$= 15 degrees Celsius $T_{culture\ out}$= 2 degrees Celsius Heat transfer efficiency assumed to be 75% As $T_{culture}/T_{cooling\ water\ in}$ is not possible for a negative cooling water in value, a value of 12 (which is 2 - -10)/1 is used resulting in $T_{NaCl\ brine\ out}$ of -1 degrees.					Based on (Ruiz et al., 2016)
Methanol recycle efficiency	95	95	95	95	%	(Ruiz et al., 2016), 5% loss is bought new
Methanol costs	0.45	0.45	0.45	0.45	Euro/kg	(Ruiz et al., 2016)
Sodium hydroxide share in reaction	1	1	1	1	% of lipid fraction	(Ruiz et al., 2016)
Sodium hydroxide costs	0.24	0.24	0.24	0.24	Euro/kg	(Ruiz et al., 2016)
Evaporator and condenser energy (III.5)	$Energy\ used\ in\ evaporator\ \&\ condenser = (Amount\ of\ methonol\ (300\%\ excess) * enthalpy\ of\ vaporisation\ Methanol) / heat\ transfer\ efficiency$ Enthalpy of vaporisation Methanol= 876 Kj/kg Efficiency heat transfer= 75% (assumed from the other heat transfer formula’s)					Based on (Ruiz et al., 2016)
Evaporator and condenser energy price (III.2)	0.066	0.066	0.066	0.066	Euro/kwh	(Renzenbrink, 2014), this price is used instead of the rest heat price as the total amount of rest heat is used in the spray driers

Cooling of the bead mill and the ATPS (II.1+II.2+III.3) systems	$Energy\ needed\ for\ cooling = Energy\ used\ for\ Beadmilling + Energy\ used\ in\ ATPS\ (II.1 + II.2 + III.1)$+ *Energy used for DF (II.7+II.8+IV.1)* $Mass\ flow\ of\ cooling\ water = \frac{energy\ to\ remove\ from\ the\ culture}{C_P \cdot \Delta T \cdot \rho_{seawater}}$ $\Delta T = T_{cooling\ water\ out} - T_{cooling\ water\ in}$ $T_{cooling\ water\ out} = T_{cooling\ water\ in} + Efficiency \cdot (\frac{T_{culture}}{T_{cooling\ water\ in}})$ Temp cooling water in 15 degrees Celsius Culture temperature= 25 for ATPS/DF units and 50 for Lipid extraction Delta T= 1.25 for ATPS and 2.5 for Lipid extraction					Based on (Ruiz et al., 2016)
Cooling water (15 degrees Celsius)	0.037	0.037	0.037	0.037	Euro/ton	(Ruiz et al., 2016)
Diafiltration power and membrane replacement (IV.1)	Same as II.7-II.8	Same as II.7-II.8	Same as II.7-II.8	Same as II.7-II.8		(Ruiz et al., 2016)
Labour (Minimum wage)	1551.6	1551.6	1551.6	1551.6	Euro/month	(Minimumloon.nl, 2017)
Number of operators	25	25	25	25	People/km^2	(Ruiz et al., 2016)
Number of Supervisors	3.9	3.9	3.9	3.9	People/km^2	(Ruiz et al., 2016)
Number of managers	1.3	1.3	1.3	1.3	People/km^2	(Ruiz et al., 2016)
Operator Salary	3	3	3	3	Times minimum wage	(Ruiz et al., 2016)
Supervisor Salary	4.3	4.3	4.3	4.3	Times minimum wage	(Ruiz et al., 2016)
Manager Salary	6.7	6.7	6.7	6.7	Times minimum wage	(Ruiz et al., 2016)
Maintenance	4	4	4	4	% of MEC	(Norsker et al., 2011)
Payroll	25	25	25	25	% of labour costs	(Norsker et al., 2011)
General plant overhead	55	55	55	55	% of labour + maintenance	(Norsker et al., 2011)

Table 34: Other Fixed costs biomass processing

Factor	**Open pond**	**Tubular**	**Plate**	**Omega**	**Unit**	**Source**
Installation costs	47	47	47	47	% of MEC	(Ruiz et al., 2016)
Instrumentation and control	35	35	35	35	% of MEC	(Ruiz et al., 2016)
Piping	40	40	40	40	% of MEC	(Ruiz et al., 2016)
Insulation	8	8	8	8	% of MEC	(Ruiz et al., 2016)
Electrical system investment	10	10	10	10	% of MEC	(Ruiz et al., 2016)

Buildings	18	18	18	18	% of MEC	(Ruiz et al., 2016)
Service facilities	40	40	40	40	% of MEC	(Ruiz et al., 2016)

Table 35: Mass flows biorefinery sections

Factor	**Open pond**		**Tubular**		**Unit**
	Optimistic	Pessimistic	Optimistic	Pessimistic	m3/day
Total volume out of centrifuge (15% to 20 % dry based)	3621	1659	5431	2489	m3/yr
Total volume out of centrifuge (15% to 20% dry based)	995722	456328	1493583	684492	m3/hr
Total volume out of centrifuge (15% to 20% dry based)	151	69	226	104	kg/hr
Total mass after centrifugation	150867	69141	226300	103711	m3/hr
Volume after II.1	75	35	113	52	kg/hr
Massa after II.1	75433	34570	113150	51855	m3/yr
Volume after II.1	497861	228164	746791	342246	m3/hr
Volume after ultrafiltration unit (II.3 and II.4) (27% dry based assumed)	56	19	84	29	kg/hr
After Lipid extractor (III.1) Lipid fraction	5294	1819	7940	2729	M3/hr
After Lipid extractor (III.1) Lipid fraction	6	2	9	3	m3/hr
After dewaxing step (III.3)	4	1	7	2	kg/yr
After dewaxing step (III.3)	25823452	8875945	38735178	13313917	m3/hr
Total amount in batch reactor (including methanol)	17	6	26	9	kg/hr
Amount really obtained	4275	1469	6412	2204	m3/hr
Volume after lipid extractor (III.1) Ash/ non-water-soluble proteins (assumed that all water moves to this phase)	50	17	75	26	kg/day
Total dry weight biomass in algae slurry out centrifuge	199144385	54759358	298716578	82139037	kg/yr
monosaccharide fraction 5%	9957219	2737968	14935829	4106952	m3/yr
Volume of algae dw	199144	68449	298717	102674	m3/yr
Total volume of water	796578	387879	1194866	581818	kg/yr
Total mass of water	796577540	387878788	1194866310	581818182	kg/hr

Volume of water evaporated (99% of total volume)	119487	58182	179230	87273	kj/kg
Evaporation heat water	2257	2257	2257	2257	m3/day
Factor	**Plate**		**OMEGA**		**Unit**
	Optimistic	Pessimistic	Optimistic	Pessimistic	
Total volume out of centrifuge (15% to 20 % dry based)	5431	2489	4526	2074	m3/day
Total volume out of centrifuge (15% to 20% dry based)	1493583	684492	1525265	699012	m3/yr
Total volume out of centrifuge (15% to 20% dry based)	226	104	189	86	m3/hr
Total mass after centrifugation	226300	103711	188584	86426	kg/hr
Volume after II.1	113	52	94	43	m3/hr
Massa after II.1	113150	51855	94292	43213	kg/hr
Volume after II.1	746791	342246	762632	349506	m3/yr
Volume after ultrafiltration unit (II.3 and II.4) (27% dry based assumed)	84	29	70	24	m3/hr
After Lipid extractor (III.1) Lipid fraction	7940	2729	6617	2274	kg/hr
After Lipid extractor (III.1) Lipid fraction	9	3	7	3	M3/hr
After dewaxing step (III.3)	7	2	5	2	m3/hr
After dewaxing step (III.3)	38735178	13313917	39556833	13596334	kg/yr
Total amount in batch reactor (including methanol)	26	9	22	7	m3/hr
Amount really obtained	6412	2204	5343	1837	kg/hr
Volume after lipid extractor (III.1) Ash/ non-water-soluble proteins (assumed that all water moves to this phase)	75	26	62	21	m3/hr
Total dry weight biomass in algae slurry out centrifuge	298716578	82139037	305052990	83881381	kg/yr
monosaccharide fraction 5%	14935829	4106952	15252649	4194069	kg/yr
Volume of algae dw	298717	102674	305053	104852	m3/yr
Total volume of water	1194866	581818	1220212	594160	m3/yr
Total mass of water	1194866310	581818182	1220211959	594159780	kg/yr
Volume of water evaporated (99% of total volume)	179230	87273	149358	72727	kg/hr

Appendix E: Costs for fractionation and refining of algae biomass

Table 36: Fractionation cost division raceway open ponds

MEC	Section	Optimistic	Pessimistic	Unit
Bead mill	I.1	€ 415,518	€ 190,427	Euro/yr
ATPS mixer/settler	II.1	€ 63,926	€ 29,296	Euro/yr
ATPS mixer/settler	II.2	€ 31,963	€ 14,648	Euro/yr
Ultrafiltration unit	II.3	€ 9,972,433	€ 4,570,252	Euro/yr
Ultrafiltration unit	II.4	€ 9,972,433	€ 4,570,252	Euro/yr
Diafiltration unit	II.7	€ 230,843	€ 79,345	Euro/yr
Diafiltration unit	II.8	€ 230,843	€ 79,345	Euro/yr
All spray driers together	II.5-II.6-II.9-II.10- IV.2	€ 126,455	€ 245,708	Euro/yr
Lipid extractor mixer/settler	III.1	€ 41,039	€ 14,106	Euro/yr
Distillation column	III.2	€ 5,508	€ 5,508	Euro/yr
Decanter+cooler	III.3	€ 4,320	€ 1,485	Euro/yr
Batch reactor	III.4	€ 29,120	€ 10,009	Euro/yr
Evaporator and condenser	III.5	€ 496	€ 496	Euro/yr
Diafiltration (DF)	IV.1	€ 220,314	€ 75,725	Euro/yr
	Total	**€ 21,345,211**	**€ 9,886,603**	Euro/yr
Operating costs		**Optimistic**	**Pessimistic**	**Unit**
Bead mill energy		€ 46,002,353	€ 21,082,353	Euro/yr
ATPS mixer-settlers energy (II.1)		€ 2,738	€ 1,255	Euro/yr
ATPS mixer-settlers energy (II.2)		€ 1,369	€ 627	Euro/yr
PEG400 costs		€ 31,772,581	€ 14,561,011	Euro/yr
Phosphate costs (II.1)		€ 11,351,230	€ 5,202,139	Euro/yr
Phosphate costs (II.2)		€ 9,932,326	€ 4,551,872	Euro/yr
Energy ultrafiltration (II.3)		€ 117,306	€ 53,760	Euro/yr
Energy ultrafiltration (II.4)		€ 117,306	€ 53,760	Euro/yr
Ultrafiltration membrane replacement (II.3)		€ 1,193,950	€ 547,174	Euro/yr
Ultrafiltration membrane replacement (II.4)		€ 1,193,950	€ 547,174	Euro/yr
Diafiltration membrane replacement (II.7)		€ 276,589	€ 95,068	Euro/yr
Diafiltration membrane replacement (II.8)		€ 276,589	€ 95,068	Euro/yr
Diafiltration power (II.7)		€ 121,699	€ 41,830	Euro/yr
Diafiltration power (II.8)		€ 121,699	€ 41,830	Euro/yr
PBS costs (II.8)		€ 15,035,401	€ 4,134,332	Euro/yr
Spray dry driers energy (all together)		€ 75,462,745	€ 28,560,000	Euro/yr
Lipid extractor mixer settler energy (III.1)		€ 1,014	€ 349	Euro/yr
Hexane		€ 12,996,791	€ 4,467,211	Euro/yr
Isopropanol		€ 74,906,239	€ 25,746,505	Euro/yr
Distillation column energy (III.2)		€ 17,314,401	€ 5,951,244	Euro/yr

Decanter+cooler energy (III.3)		€ 4,907	€ 1,687	Euro/yr
Methanol costs		€ 1,743,083	€ 599,126	Euro/yr
Sodium hydroxide costs		€ 61,976	€ 21,302	Euro/yr
evaporator and condenser energy (III.5)		€ 1,658,899	€ 570,191	Euro/yr
Diafiltration power (IV.1)		€ 108,889	€ 37,427	Euro/yr
Diafiltration membrane replacement (IV.1)		€ 247,475	€ 85,061	Euro/yr
Labour		€ 17,709,166	€ 22,136,458	Euro/yr
Maintenance		€ 853,808	€ 395,464	Euro/yr
Payroll		€ 4,427,292	€ 5,534,114	Euro/yr
General plant overhead		€ 10,209,636	€ 12,392,557	Euro/yr
Cooling of the system (total cooling)		€ 17,758,533	€ 8,136,954	Euro/yr
	Total	**€ 352,981,941**	**€ 165,644,902**	Euro/yr
Other fixed costs (direct investment)		**Optimistic**	**Pessimistic**	**Unit**
Installation costs		€ 10,032,249	€ 4,646,703	Euro/yr
Instrumentation and control		€ 7,470,824	€ 3,460,311	Euro/yr
Piping		€ 8,538,085	€ 3,954,641	Euro/yr
Insulation		€ 1,707,617	€ 790,928	Euro/yr
Electrical system investment		€ 2,134,521	€ 988,660	Euro/yr
Buildings		€ 3,842,138	€ 1,779,589	Euro/yr
Service facilities		€ 8,538,085	€ 3,954,641	Euro/yr
	Total	**€ 42,263,519**	**€ 19,575,474**	Euro/yr
Jet fuel processing		**Optimistic**	**Pessimistic**	**Unit**
Jet fuel refinery		**€ 14,023,305**	**€ 3,856,032**	Euro/yr
	Total overall	**€ 430,613,977**	**€ 198,963,011**	Euro/yr
	Total useful dry biomass	**173046896**	**47583250**	kg/yr
	Total	**€ 2.49**	**€ 4.18**	Euro/kg

Table 37: Fractionation cost division Tubular reactor

MEC	Section	Optimistic	Pessimistic	Unit
Bead mill	I.1	€ 623,277	€ 285,641	Euro/yr
ATPS mixer/settler	II.1	€ 95,889	€ 43,945	Euro/yr
ATPS mixer/settler	II.2	€ 47,944	€ 21,972	Euro/yr
Ultrafiltration unit	II.3	€ 14,958,650	€ 6,855,378	Euro/yr
Ultrafiltration unit	II.4	€ 14,958,650	€ 6,855,378	Euro/yr
Diafiltration unit	II.7	€ 346,265	€ 119,017	Euro/yr
Diafiltration unit	II.8	€ 346,265	€ 119,017	Euro/yr
All spray driers together	II.5-II.6-II.9-II.10- IV.2	€ 756,908	€ 368,562	Euro/yr
Lipid extractor mixer/settler	III.1	€ 61,558	€ 21,159	Euro/yr
Distillation column	III.2	€ 5,508	€ 5,508	Euro/yr
Decanter+cooler	III.3	€ 6,480	€ 2,227	Euro/yr
Batch reactor	III.4	€ 43,680	€ 15,013	Euro/yr
Evaporator and condenser	III.5	€ 496	€ 496	Euro/yr
Diafiltration (DF)	IV.1	€ 330,470	€ 113,588	Euro/yr
	Total	**€ 32,582,040**	**€ 14,826,902**	Euro/yr
Operating costs		**Optimistic**	**Pessimistic**	**Unit**
Bead mill energy		€ 69,003,529	€ 31,623,529	Euro/yr
ATPS mixer-settlers energy (II.1)		€ 4,107	€ 1,882	Euro/yr
ATPS mixer-settlers energy (II.2)		€ 2,054	€ 941	Euro/yr
PEG400 costs		€ 47,658,872	€ 21,841,517	Euro/yr
Phosphate costs (II.1)		€ 17,026,845	€ 7,803,209	Euro/yr
Phosphate costs (II.2)		€ 14,898,489	€ 6,827,807	Euro/yr
Energy ultrafiltration (II.3)		€ 175,959	€ 80,640	Euro/yr
Energy ultrafiltration (II.4)		€ 175,959	€ 80,640	Euro/yr
Ultrafiltration membrane replacement (II.3)		€ 1,790,925	€ 820,760	Euro/yr
Ultrafiltration membrane replacement (II.4)		€ 1,790,925	€ 820,760	Euro/yr
Diafiltration membrane replacement (II.7)		€ 414,884	€ 142,602	Euro/yr
Diafiltration membrane replacement (II.8)		€ 414,884	€ 142,602	Euro/yr
Diafiltration power (II.7)		€ 182,549	€ 62,745	Euro/yr
Diafiltration power (II.8)		€ 182,549	€ 62,745	Euro/yr
PBS costs (II.8)		€ 22,553,102	€ 6,201,497	Euro/yr
Spray dry driers energy (all together)		€ 164,364,118	€ 28,560,000	Euro/yr
Lipid extractor mixer settler energy (III.1)		€ 1,521	€ 523	Euro/yr
Hexane		€ 19,495,187	€ 6,700,816	Euro/yr
Isopropanol		€ 112,359,358	€ 38,619,758	Euro/yr
Distillation column energy (III.2)		€ 25,971,601	€ 8,926,866	Euro/yr
Decanter+cooler energy (III.3)		€ 7,360	€ 2,530	Euro/yr
Methanol costs		€ 2,614,625	€ 898,689	Euro/yr

Sodium hydroxide costs		€ 92,964	€ 31,953	Euro/yr
evaporator and condenser energy (III.5)		€ 2,488,348	€ 855,286	Euro/yr
Diafiltration power (IV.1)		€ 163,333	€ 56,140	Euro/yr
Diafiltration membrane replacement (IV.1)		€ 371,212	€ 127,592	Euro/yr
Labour		€ 13,281,875	€ 26,563,749	Euro/yr
Maintenance		€ 1,303,282	€ 593,076	Euro/yr
Payroll		€ 3,320,469	€ 6,640,937	Euro/yr
General plant overhead		€ 8,021,836	€ 14,936,254	Euro/yr
Cooling of the system (total cooling)		€ 26,637,799	€ 12,205,430	Euro/yr
	Total	**€ 556,770,520**	**€ 222,233,479**	Euro/yr
Other fixed costs (direct investment)		**Optimistic**	**Pessimistic**	**Unit**
Installation costs		€ 15,313,559	€ 6,968,644	Euro/yr
Instrumentation and control		€ 11,403,714	€ 5,189,416	Euro/yr
Piping		€ 13,032,816	€ 5,930,761	Euro/yr
Insulation		€ 2,606,563	€ 1,186,152	Euro/yr
Electrical system investment		€ 3,258,204	€ 1,482,690	Euro/yr
Buildings		€ 5,864,767	€ 2,668,842	Euro/yr
Service facilities		€ 13,032,816	€ 5,930,761	Euro/yr
	Total	**€ 64,512,439**	**€ 29,357,266**	Euro/yr
Jet fuel processing		**Optimistic**	**Pessimistic**	**Unit**
Jet fuel refinery		**€ 21,034,958**	**€ 5,784,048**	Euro/yr
	Total overall	**€ 674,899,956**	**€ 272,201,696**	Euro/yr
	Total useful dry biomass	**259570345**	**71374875**	kg/yr
	Total	**€ 2.60**	**€ 3.81**	Euro/kg

Table 38: Fractionation cost division Plate reactor

MEC	Section	Optimistic	Pessimistic	Unit
Bead mill	I.1	€ 623,277	€ 285,641	Euro/yr
ATPS mixer/settler	II.1	€ 95,889	€ 43,945	Euro/yr
ATPS mixer/settler	II.2	€ 47,944	€ 21,972	Euro/yr
Ultrafiltration unit	II.3	€ 14,958,650	€ 6,855,378	Euro/yr
Ultrafiltration unit	II.4	€ 14,958,650	€ 6,855,378	Euro/yr
Diafiltration unit	II.7	€ 346,265	€ 119,017	Euro/yr
Diafiltration unit	II.8	€ 346,265	€ 119,017	Euro/yr
All spray driers together	II.5-II.6-II.9-II.10- IV.2	€ 756,908	€ 368,562	Euro/yr
Lipid extractor mixer/settler	III.1	€ 61,558	€ 21,159	Euro/yr
Distillation column	III.2	€ 5,508	€ 5,508	Euro/yr
Decanter+cooler	III.3	€ 6,480	€ 2,227	Euro/yr
Batch reactor	III.4	€ 43,680	€ 15,013	Euro/yr
Evaporator and condenser	III.5	€ 496	€ 496	Euro/yr
Diafiltration (DF)	IV.1	€ 330,470	€ 113,588	Euro/yr
	Total	**€ 32,582,040**	**€ 14,826,902**	Euro/yr
Operating costs		**Optimistic**	**Pessimistic**	**Unit**
Bead mill energy		€ 69,003,529	€ 31,623,529	Euro/yr
ATPS mixer-settlers energy (II.1)		€ 4,107	€ 1,882	Euro/yr
ATPS mixer-settlers energy (II.2)		€ 2,054	€ 941	Euro/yr
PEG400 costs		€ 47,658,872	€ 21,841,517	Euro/yr
Phosphate costs (II.1)		€ 17,026,845	€ 7,803,209	Euro/yr
Phosphate costs (II.2)		€ 14,898,489	€ 6,827,807	Euro/yr
Energy ultrafiltration (II.3)		€ 175,959	€ 80,640	Euro/yr
Energy ultrafiltration (II.4)		€ 175,959	€ 80,640	Euro/yr
Ultrafiltration membrane replacement (II.3)		€ 1,790,925	€ 820,760	Euro/yr
Ultrafiltration membrane replacement (II.4)		€ 1,790,925	€ 820,760	Euro/yr
Diafiltration membrane replacement (II.7)		€ 414,884	€ 142,602	Euro/yr
Diafiltration membrane replacement (II.8)		€ 414,884	€ 142,602	Euro/yr
Diafiltration power (II.7)		€ 182,549	€ 62,745	Euro/yr
Diafiltration power (II.8)		€ 182,549	€ 62,745	Euro/yr
PBS costs (II.8)		€ 22,553,102	€ 6,201,497	Euro/yr
Spray dry driers energy (all together)		€ 164,364,118	€ 28,560,000	Euro/yr
Lipid extractor mixer settler energy (III.1)		€ 1,521	€ 523	Euro/yr
Hexane		€ 19,495,187	€ 6,700,816	Euro/yr
Isopropanol		€ 112,359,358	€ 38,619,758	Euro/yr
Distillation column energy (III.2)		€ 25,971,601	€ 8,926,866	Euro/yr
Decanter+cooler energy (III.3)		€ 7,360	€ 2,530	Euro/yr
Methanol costs		€ 2,614,625	€ 898,689	Euro/yr

Sodium hydroxide costs		€ 92,964	€ 31,953	Euro/yr
evaporator and condenser energy (III.5)		€ 2,488,348	€ 855,286	Euro/yr
Diafiltration power (IV.1)		€ 163,333	€ 56,140	Euro/yr
Diafiltration membrane replacement (IV.1)		€ 371,212	€ 127,592	Euro/yr
Labour		€ 7,969,125	€ 14,757,638	Euro/yr
Maintenance		€ 1,303,282	€ 593,076	Euro/yr
Payroll		€ 1,992,281	€ 3,689,410	Euro/yr
General plant overhead		€ 5,099,823	€ 8,442,893	Euro/yr
Cooling of the system (total cooling)		€ 26,637,799	€ 12,205,430	Euro/yr
	Total	**€ 547,207,570**	**€ 200,982,480**	Euro/yr
Other fixed costs (direct investment)		**Optimistic**	**Pessimistic**	**Unit**
Installation costs		€ 15,313,559	€ 6,968,644	Euro/yr
Instrumentation and control		€ 11,403,714	€ 5,189,416	Euro/yr
Piping		€ 13,032,816	€ 5,930,761	Euro/yr
Insulation		€ 2,606,563	€ 1,186,152	Euro/yr
Electrical system investment		€ 3,258,204	€ 1,482,690	Euro/yr
Buildings		€ 5,864,767	€ 2,668,842	Euro/yr
Service facilities		€ 13,032,816	€ 5,930,761	Euro/yr
	Total	**€ 64,512,439**	**€ 29,357,266**	Euro/yr
Jet fuel processing		**Optimistic**	**Pessimistic**	**Unit**
Jet fuel refinery		**€ 21,034,958**	**€ 5,784,048**	Euro/yr
	Total overall	**€ 665,337,006**	**€ 250,950,697**	Euro/yr
	Total useful dry biomass	**259570345**	**71374875**	kg/yr
	Total	**€ 2.56**	**€ 3.52**	Euro/kg

Table 39: Fractionation cost division OMEGA

MEC	Section	Optimistic	Pessimistic	Unit
Bead mill	I.1	€ 519,398	€ 238,034	Euro/yr
ATPS mixer/settler	II.1	€ 79,907	€ 36,621	Euro/yr
ATPS mixer/settler	II.2	€ 39,954	€ 18,310	Euro/yr
Ultrafiltration unit	II.3	€ 12,465,541	€ 5,712,815	Euro/yr
Ultrafiltration unit	II.4	€ 12,465,541	€ 5,712,815	Euro/yr
Diafiltration unit	II.7	€ 288,554	€ 99,181	Euro/yr
Diafiltration unit	II.8	€ 288,554	€ 99,181	Euro/yr
All spray driers together	II.5-II.6-II.9-II.10- IV.2	€ 630,756	€ 307,135	Euro/yr
Lipid extractor mixer/settler	III.1	€ 51,299	€ 17,632	Euro/yr
Distillation column	III.2	€ 5,508	€ 5,508	Euro/yr
Decanter+cooler	III.3	€ 5,400	€ 1,856	Euro/yr
Batch reactor	III.4	€ 36,400	€ 12,511	Euro/yr
Evaporator and condenser	III.5	€ 496	€ 496	Euro/yr
Diafiltration (DF)	IV.1	€ 275,392	€ 94,657	Euro/yr
	Total	**€ 27,152,701**	**€ 12,356,753**	Euro/yr
Operating costs		**Optimistic**	**Pessimistic**	**Unit**
Bead mill energy		€ 70,467,241	€ 32,294,332	Euro/yr
ATPS mixer-settlers energy (II.1)		€ 4,194	€ 1,922	Euro/yr
ATPS mixer-settlers energy (II.2)		€ 2,097	€ 961	Euro/yr
PEG400 costs		€ 39,715,727	€ 18,201,264	Euro/yr
Phosphate costs (II.1)		€ 17,388,020	€ 7,968,731	Euro/yr
Phosphate costs (II.2)		€ 15,214,518	€ 6,972,640	Euro/yr
Energy ultrafiltration (II.3)		€ 179,691	€ 82,351	Euro/yr
Energy ultrafiltration (II.4)		€ 179,691	€ 82,351	Euro/yr
Ultrafiltration membrane replacement (II.3)		€ 1,492,437	€ 683,967	Euro/yr
Ultrafiltration membrane replacement (II.4)		€ 1,492,437	€ 683,967	Euro/yr
Diafiltration membrane replacement (II.7)		€ 423,685	€ 145,627	Euro/yr
Diafiltration membrane replacement (II.8)		€ 423,685	€ 145,627	Euro/yr
Diafiltration power (II.7)		€ 152,124	€ 52,288	Euro/yr
Diafiltration power (II.8)		€ 152,124	€ 52,288	Euro/yr
PBS costs (II.8)		€ 23,031,501	€ 6,333,044	Euro/yr
Spray dry driers energy (all together)		€ 170,021,478	€ 28,560,000	Euro/yr
Lipid extractor mixer settler energy (III.1)		€ 1,554	€ 534	Euro/yr
Hexane		€ 19,908,721	€ 6,842,955	Euro/yr
Isopropanol		€ 93,632,799	€ 32,183,132	Euro/yr
Distillation column energy (III.2)		€ 26,522,514	€ 9,116,224	Euro/yr
Decanter+cooler energy (III.3)		€ 7,516	€ 2,583	Euro/yr
Methanol costs		€ 2,670,086	€ 917,753	Euro/yr

Sodium hydroxide costs		€ 94,936	€ 32,631	Euro/yr
evaporator and condenser energy (III.5)		€ 2,541,131	€ 873,428	Euro/yr
Diafiltration power (IV.1)		€ 166,798	€ 57,331	Euro/yr
Diafiltration membrane replacement (IV.1)		€ 379,086	€ 130,298	Euro/yr
Labour		€ 29,337,442	€ 29,925,897	Euro/yr
Maintenance		€ 1,086,108	€ 494,270	Euro/yr
Payroll		€ 7,334,361	€ 7,481,474	Euro/yr
General plant overhead		€ 16,732,953	€ 16,731,092	Euro/yr
Cooling of the system (total cooling)		€ 27,200,198	€ 12,463,424	Euro/yr
	Total	**€ 567,956,854**	**€ 219,514,386**	Euro/yr
Other fixed costs (direct investment)		**Optimistic**	**Pessimistic**	**Unit**
Installation costs		€ 12,761,769	€ 5,807,674	Euro/yr
Instrumentation and control		€ 9,503,445	€ 4,324,863	Euro/yr
Piping		€ 10,861,080	€ 4,942,701	Euro/yr
Insulation		€ 2,172,216	€ 988,540	Euro/yr
Electrical system investment		€ 2,715,270	€ 1,235,675	Euro/yr
Buildings		€ 4,887,486	€ 2,224,215	Euro/yr
Service facilities		€ 10,861,080	€ 4,942,701	Euro/yr
	Total	**€ 53,762,347**	**€ 24,466,370**	Euro/yr
Jet fuel processing		**Optimistic**	**Pessimistic**	**Unit**
Jet fuel refinery		**€ 21,481,154**	**€ 5,906,740**	Euro/yr
	Total overall	**€ 670,353,055**	**€ 262,244,249**	Euro/yr
	Total useful dry biomass	**265076382**	**72888887**	kg/yr
	Total	**€ 2.53**	**€ 3.60**	Euro/kg