# Electronic Supplementary Materials for

# Towards industrial products from microalgae

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**Methods:**

Two interdependent techno-economic models are developed for biomass production and biorefinery processing. Both models are based on available empirical information and literature, allowing to project scenarios from different assumptions and to perform sensitivity analysis on the processes. Projections are done for a 1 hectare and 100 hectares production scales in different cultivation systems. The produced biomass is subsequently processed via simulated biorefinery chain to a variety of products. It should be noted that the assumptions used in the models involve an inherent uncertainty, like scalability of the results or extrapolation to different locations.

- **GENERAL INPUTS:**

Production and biorefinery models work on the basis of a set of shared inputs described here:

Locations and climatology:

Six different locations are included in this study: The Netherlands (52°17´ N 4°46´ E), Canary Islands (27°55´ N 15°22´ W), South of Spain (37°15´ N 6°56´ W), Turkey (38°30´ N 27°01´ E), Saudi Arabia (24°42´ N 46°47´ E) and Curaçao (12°7´ N 68°56´ W). Parameters that change with location are included in this study; i.e. climatic conditions, energy cost, cost of labor, employer's contribution to labor costs and standard workweek hours (Supplementary Table 1). In case of Curaçao climatic data from the Coast of Venezuela (10°36´ N 66°58´ W; ~200 km away) was used.

Labor:

Manpower cost derives from the estimated number of workers (assuming a standard workweek of 40 hours), qualification and cost of working hour (from salary and number of hours per workweek) (Supplementary Table 1). Salaries are based on minimum wages per location, being assigned salaries of a plant manager, supervisor and operator 6.7, 4.3 and 3 times minimum wage respectively (from [1], assuming the occupation titles "Industrial production managers", "First-Line Supervisors of Mechanics, Installers, and Repairers" and "Installation, Maintenance, and Repair Occupations" from the study for plant manager, supervisor and operator respectively). The employer's contribution is added to the manpower cost to cover for the liability of work-related accidents and occupational illness (Supplementary Table 1). Labor cost is finally increased by 20% for labor supervision activities.

Electricity supply and wastewater:

Industrial prices for electricity supply at the location of production are considered (Supplementary Table 1). Treatment of wastewater performed at a cost of 0.43 $€\cdot m^{-3}$ [2], by an external party is assumed (the energy to perform it is excluded from the study).

**Supplementary Table 1: Specific parameters affecting the economic analysis for different locations**

| **Location** | **Energy cost** ($€\cdot kWh^{-1}$) | **Minimum wages** ($€\cdot Yr^{-1}$) | **Employer's contributions** (% of labor cost) | **Standard workweek** (hours) |
|---|---|---|---|---|
| **Canary Islands** | 0.122[a] | 9,080[d] | 23.6[g] | 40[d] |
| **The Netherlands** | 0.096[a] | 18,021[d] | 18.8[h] | 40[d] |
| **Saudi Arabia** | 0.029[b] | 6,936[e] | 11[i] | 48[d] |
| **Curaçao** | 0.307[c] | 7,608[f] | 10.9[j] | 45[l] |
| **Turkey** | 0.093[a] | 5,091[d] | 22.5[k] | 45[d] |
| **South of Spain** | 0.122[a] | 9,080[d] | 23.6[g] | 40[d] |

[a] [3]

[b] Source: Saudi Electricity Company. For a currency conversion 1 € = 5.19 Saudi Arabian Riyal

[c] [4] For a currency conversion 1 € = 1.89 Netherlands Antillean Guilder

[d] Source: EUROSTAT

[e] Source: U.S. Department of State

[f] [5]

[g] Source: Ministerio de Empleo y Seguridad Social

[h] Source: Netherlands Foreign Investment Agency

[i] Source: HSBC Expat
[j] Source: [6]
[k] Source: Expat Guide Turkey
[l] Source: Curaçao chronicle

Capital costs:

Major equipment is depreciated over 15 years with an 8% interest rate. Major equipment costs (MEC) are not location dependent.

Lang factors are used for estimation of capital investment, by multiplying the major equipment cost by specific Lang factors to obtain the weight of the different items in this cost (Supplementary Table 2). This technique is used frequently to obtain cost estimates of a process plant, varying these factors upon the type of product or process.

In the model for algae cultivation, Lang factors used for the breakdown of the capital investment are similar as used in [7] for a microalgae production plant, with the exception of instrumentation and control, land cost, construction expenses, contingency, contractor's fee and purchase tax (Supplementary Table 2). In our experience, the cost of instrumentation and control is higher for closed cultivation systems than considered in the previous study [7] and, consequently is increased by a factor ten (150% of major equipment cost). In addition, construction expenses and contingencies are also increased; the former changes to the average for ordinary chemical process plants (10% of the total direct costs) [8] and the latter to the highest value commonly used due to the novelty of the process, i.e. 15% of the direct and indirect plant costs [8]. Contractor's fee is estimated as 5% of the direct plant cost, average number given by [8].

In the model for biorefinery, the adopted Lang factors are closer to the value reported for fluid processing chemical plants [9] (Supplementary Table 2). Since utilities are more intensely used in the downstream section (steam and cooling agent Freon), service facilities are increased (40% of MEC). Furthermore, due to higher temperatures in the process for the extraction and fractionation of lipids additional insulation is needed (8% MEC). The facility dependent costs (depreciation, maintenance, local insurance and property taxes) are calculated from the total fixed capital per year following the same procedure as [7]. Cost of laboratory, R&D and quality control are not taken into account in the operational cost.

Purchase tax is neglected; since this is recoverable (a profitable company would get tax return). Land is rented at a not location-specific cost established as 1.100 $€·Ha^{-1}·Yr^{-1}$ (data based on price of rented agricultural land in The Netherlands). The extra land required, such as space to place major equipment, buildings or roads, was considered as 20% of the total photobioreactor area (1 or 100 hectares), the total land for the facility being 1.2 or 120 hectares.

With the method of Lang factors, total capital investment for the biomass production facility becomes 501% the major equipment cost for closed systems and 324% for raceway ponds. For algal biomass refinery, total capital investment resulted in 497% the major equipment cost.

**Supplementary Table 2: Procedures for estimating CAPEX and OPEX**

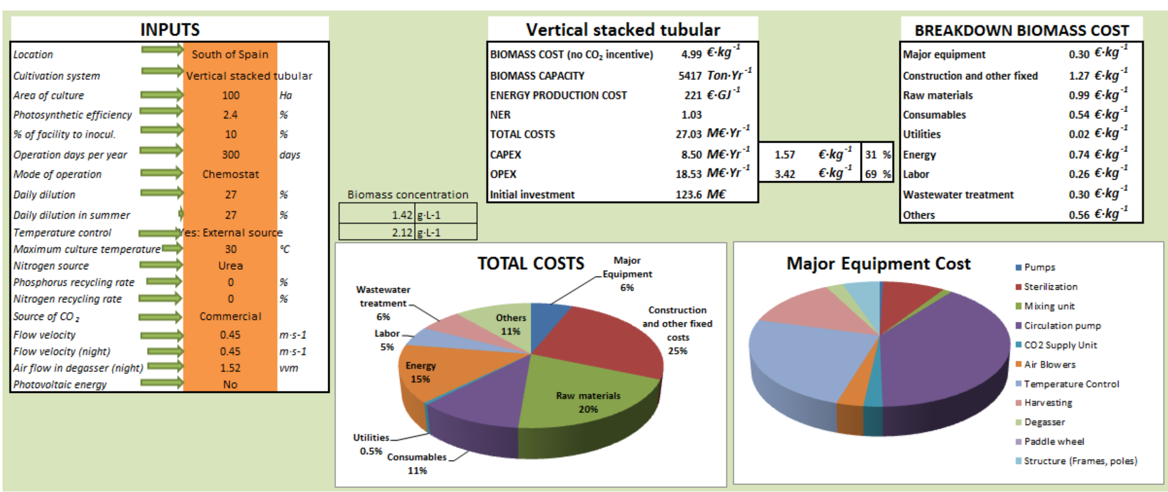
| | | | ***Cultivation*** | ***Biorefinery*** |
|---|---|---|---|---|
| **FIXED CAPITAL INVESTMENT** | Direct Cost (DC) | Major equipment | MEC | MEC |
| | | Installation costs | 20% MEC | 47% MEC |
| | | Instrumentation and control | 15[a] - 150%[b] MEC | 35% MEC |
| | | Piping[c] | 20% MEC | 40% MEC |
| | | Insulation | 0% MEC | 8% MEC |
| | | Electrical | 10% MEC | 10% MEC |
| | | Buildings | 23% MEC | 18% MEC |
| | | Land improvements | 12% MEC | 10% MEC |
| | | Service facilities | 20% MEC | 40% MEC |
| | Indirect Cost (IC) | Construction expenses | 10% DC | 15% DC |
| | | Engineering and supervision | 30% MEC | 10% DC |
| | Other Cost (OC) | Contractor's fee | 5% (DC) | 5% (DC+IC) |
| | | Contingency (Major equipment) | 15% (DC+IC) | 15% (DC+IC) |
| | Working capital | | OPEX first three months of operation | |
| **CAPEX** | Depreciation | | (DC+IC+OC)/15 years | |
| | Interest | | 8% of depreciation | |
| | Property tax | | 1% of depreciation+interest | |
| | Insurance | | 0.6% of depreciation+interest | |
| | Purchase tax | | Excluded | |
| | Land | | 1.100 €·$Ha^{-1}$·$Yr^{-1}$ | |
| **OPEX** | Energy | | Calculated from MEC consumption | |
| | Labor | | Salaries + Employer's contribution +Supervision | |
| | Raw materials | | Calculated from mass balances | |
| | Utilities | | Calculated from MEC consumption and mass balances | |
| | Wastewater treatment | | Calculated from mass balances | |
| | Consumables | | Calculated from MEC design | |
| | Others | Maintenance | 4% MEC | |
| | | Operating supplies | 0.4% (Electricity + Raw materials + Utilities) | |
| | | Contingencies | 15% (Raw materials + Utilities) | |
| | | Overheads | 55% (Labor + Maintenance) | |

[a] Raceway pond [b] Closed systems [c] Piping used to channel cooling water from the sea is considered major equipment and calculated as part of MEC (see below)

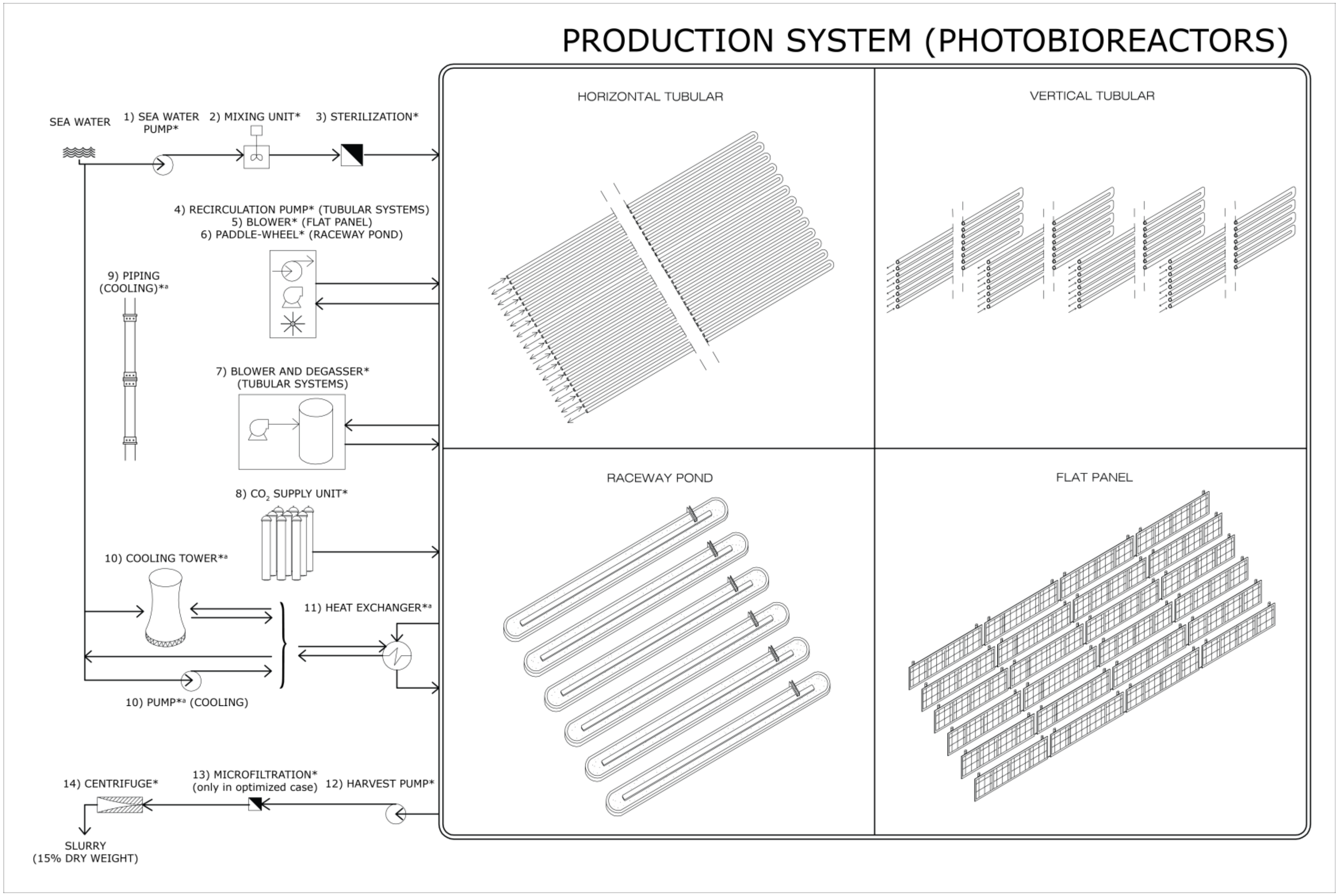
## - MICROALGAL CULTIVATION:

Basis:

Four types of microalgae cultivation systems are analyzed: horizontal tubular photobioreactor, vertically stacked horizontal tubular photobioreactor, flat panel photobioreactor and raceway pond. Each projection is based on one of these four systems. The algal production chain (Supplementary Fig. 1) starts with natural seawater, which is pumped and enriched in nutrients in a mixing unit. This seawater based medium is sterilized by filtration and added to the selected cultivation system. Medium addition to the systems takes place only during daylight hours, while the broth leaves the reactors continuously. Carbon dioxide supply units add inorganic carbon to the culture. The culture is mixed via a pump, blower or paddle wheel depending on the system. The harvest is continuously pumped from the culture systems to centrifuges, obtaining algal slurry (15 % w/w) as end product of cultivation. The slurry at this concentration can be pumped and used in the biorefinery process. In the projection for the optimized case a microfiltration unit pre-concentrates the culture prior to dewatering by centrifuges. A combination of heat exchangers, pipes and pumps is installed to control temperature in closed systems. Deep sea water is directly used as cooling water and then discharged back to the sea. The use of a hypothetical cooling tower as alternative source of cooling water is also studied as an option.




**Supplementary Picture 1: Main screen of the model developed for microalgal cultivation**




**Supplementary Figure 1: Scheme of the microalgal production chain.**
***Major equipment.**
**[a] Only in closed systems.**

The facility produces microalgal biomass as slurry with 15% solids (dry weight). The amount of biomass produced per year is calculated from the total annual irradiation for the selected location, the photosynthetic efficiencies obtained outdoors at AlgaePARC pilot facility in The Netherlands (fraction of total light energy converted into chemical energy during photosynthesis) and the chemical energy stored in the biomass (biomass combustion enthalpy, considered constant for non-stressed biomass at a value of 22.5 $kJ \cdot g^{-1}$ [10–12]). AlgaePARC (www.algaeparc.com) is a research pilot facility in The Netherlands aiming to fill the gap between fundamental research on algae and full-scale production facilities.

The following formula is used to calculate the biomass productivity:

$$Annual\ Productivity = \frac{Light\ intensity\ x\ Photosynthetic\ efficiency}{Biomass\ combustion\ enthalpy}\ x\ \frac{Days\ of\ operation}{365\ days}$$

For The Netherlands, 270 days of operation per year are considered, while for other locations with more favorable climatic conditions the simulation is done for 300 operational days per year. Downtime is required for maintenance, starting new cultures, possible contingencies or because the weather does not allow production. For cleaning and maintenance selective unit operations will be out of order without affecting the overall operation. An exception will be The Netherlands as frost limits the operational timeframe, hence a longer downtime. The number of operational days affects annual plant productivity, volumes processed and energy consumption.

Input data for projections are experimental results from the pilot plant facility AlgaePARC, climatological database (http://www.energy.gov/; http://www.soda-is.com/eng/index.html; http://www.windguru.cz), suppliers of equipment, suppliers of raw materials and literature, as shown in more detail in other sections.

Climatologic information used in this study is based on average hourly data. Irradiation values influence biological parameters such as productivity and oxygen production, which combined with day length and dilution rate determine the major equipment needed. Data on temperature, irradiation, relative humidity, wet bulb temperature, dew point temperature and wind speed are used to estimate the requirements for temperature control of the culture.

Cost analysis:

Total annual costs divided by total dry biomass annual production yields the biomass cost ($€ \cdot kg^{-1}$). Since the biomass is produced as a slurry with water, the unit production cost is based on the dry weight of the biomass in the slurry and not the volume of the entire slurry. Total annual costs are calculated by summing annual capital expenditures (CAPEX) and operating expenses (OPEX).

Cost of major equipment, consumables and materials is obtained directly from suppliers when possible; otherwise prices are derived from standard engineering estimates or literature. In case the retrieved cost for a certain component is not from the current year, the price is then updated to the base year using the Harmonized Indices of Consumer Prices (HICP) for the European Union (Source of Data: Eurostat). These costs are listed in Supplementary Tables 3 and 4.

**Supplementary Table 3: Details about the major equipment considered in the study**

| NUMBER IN SUPPLEMENTARY FIG. 1 | SCALE (Ha) | MAJOR EQUIPMENT | Capacity | €·unit$^{-1}$ | Power |
|---|---|---|---|---|---|
| 1, 12 | 1 | **Pumps** | 2 $m^3 \cdot h^{-1}$ | 455 | 0.18 kW |
| 1, 12 | 1 | **Pumps** | 4 $m^3 \cdot h^{-1}$ | 1,035 | 0.40 kW |
| 1, 12 | 100 | **Pumps** | 200 $m^3 \cdot h^{-1}$ | 13,544 | 5.9 kW |
| 3 | 1 | **Sterilization** | 5.99 $m^3 \cdot h^{-1}$ | 16,665 | - |
| 3 | 100 | **Sterilization** | 59.9 $m^3 \cdot h^{-1}$ | 117,979 | - |
| 2 | 1 | **Mixing unit*** | 0.1 $m^3$ | 14,000 | 0.05 kW |
| 2 | 100 | **Mixing unit*** | 4 $m^3$ | 220,000 | 2.07 kW |
| 2 | 100 | **Mixing unit*** | 8 $m^3$ | 243,000 | 4.15 kW |
| 2 | 100 | **Mixing unit*** | 25 $m^3$ | 291,000 | 12.96 kW |
| 4 | 1 | **Culture circulation pump** | 700 $m^3 \cdot h^{-1}$ | 28,105 | See section "Power requirement for liquid circulation" |
| 4, 10 | 100 | **Culture circulation pump and temperature control** | 28,000 $m^3 \cdot h^{-1}$ | 595,600 | See section "Power requirement for liquid circulation" |
| 5 | 1 | **Air blower** | 1,000 $m^3 \cdot h^{-1}$ | 5,653 | 3.96 kW |
| 5, 7 | 100 | **Air blower** | 2,499 $m^3 \cdot h^{-1}$ | 11,182 | 11.15 kW |
| 6 | 1 and 100 | **Paddle wheel** | 1 Ha of ponds | 13,679 | 0.36 $W \cdot m^{-2}$ (Flow 0.25 $m \cdot s^{-1}$) |
| 7 | 1 | **Air blower** | 200 $m^3 \cdot h^{-1}$ | 3,027 | 0.99 kW |

| | | | | | |
|---|---|---|---|---|---|
| 7 | 1 | **Degasser** | 0.66 $m^3$ | 1,214 | - |
| 7 | 100 | **Degasser** | 6.6 $m^3$ | 2,503 | - |
| 8 | 1 and 100 | **$CO_2$ supply unit** | 1 Ha | 4,717 | Insignificant |
| 9 | 1 and 100 | **Piping (cooling)** | 1 m | 350 | - |
| 11 | 1 and 100 | **Heat exchanger** | - | - | See section "Temperature control" |
| 10 | 1 and 100 | **Cooling tower** | - | - | See section "Temperature control" |
| 13 | 1 and 100 | **Microfiltration unit** | 32 $L \cdot m^{-2} \cdot h^{-1}$ | 71 $€ \cdot m^{-2}$ membr | 0.375 $kW \cdot m^{-3}$ |
| 14 | 1 | **Centrifuge** | 0.13 $m^3 \cdot h^{-1}$ | 27,000 | 1.1 kW |
| 14 | 1 | **Centrifuge** | 2.1 $m^3 \cdot h^{-1}$ | 51,000 | 4.0 kW |
| 14 | 1 and 100 | **Centrifuge** | 16.3 $m^3 \cdot h^{-1}$ | 115,000 | 22 kW |
| 14 | 100 | **Centrifuge** | 65 $m^3 \cdot h^{-1}$ | 300,000 | 55 kW |
| - | 1 and 100 | **Steel mesh casing (flat panel)** | 0.875 $kg \cdot m^{-2}$ | 650 $€ \cdot ton^{-1}$ | - |
| - | 1 and 100 | **Metal poles (vertical tubular)** | 3.8 $kg \cdot m^{-1}$ | 621 $ton^{-1}$ | - |

*30 seconds retention time [13]

**Supplementary Table 4: Information about the raw materials and consumables used in the simulation**

| **RAW MATERIALS AND CONSUMABLES** | Price | Lifetime (Years) |
|---|---|---|
| **Commercial $CO_2$** | 184 $€ \cdot ton^{-1}$ | - |
| **$CO_2$ from flue gas** | 29 $€ \cdot ton^{-1}$ | - |
| **Nitrogen from urea** | 633 $€ \cdot ton^{-1}$ | - |
| **Phosphorus from triple superphosphate** | 1,155 $€ \cdot ton^{-1}$ | - |
| **Polyethylene tubes (horizontal tubular)*** | 0.20 $€ \cdot m^{-1}$ | 1 |
| **Glass tubes (vertical tubular)*** | 4.13 $€ \cdot m^{-1}$ | 20 |
| **Plastic lining (raceway pond)*** | 102,000 $€ \cdot Ha^{-1}$ | 25 |
| **Polyethylene film (flat panel)*** | 0.19 $€ \cdot m^{-2}$ | 1 |
| **Microfiltration membranes** | 26 $€ \cdot m^{-2}$ | 3 |
| **Chemical cleaning** | 668 $€ \cdot m^{-3}$ | - |

| | | |
|---|---|---|
| **Plastic granulates (cleaning in tubular systems)** | 22.9 €·kg$^{-1}$ | 3 |
| **Rental of cleaning device (cleaning in raceway ponds)** | 409 €·day$^{-1}$ | - |

*20% installation cost is included

The optimum equipment in terms of performance and capacity for the location, scale, system and operation is selected in each specific case among those in Supplementary Table 3. Number of units of major equipment for each specific case is based on mass balances for the peak capacity, i.e. for the month with the highest irradiation. To achieve a conservative economic estimate the number of processing units is calculated considering operation at 90% of the maximum capacity of equipment. The number of units needed is rounded to the next larger integer.

CAPEX is derived from the capital investment, its depreciation and interest, while OPEX is the annual sum of raw materials, consumables, energy, utilities, labor, maintenance, operating supplies, overheads, contingencies and wastewater treatment cost. Maintenance, operating supplies and general plant overheads are calculated as factors of the purchased major equipment by following the same procedure as [7] (Supplementary Table 2). Other contingencies related to raw materials and utilities are increased to 15% (Supplementary Table 2) compared to 5% used previously [7] since this is not a mature process yet.

Tubes, polyethylene film and plastic liners for tubular systems, flat panels and raceway ponds respectively, as well as filtration membranes are considered as consumables. The annual cost of these consumables is obtained by multiplying the unit cost by the number of units, and then divided by the lifetime. Nutrients (nitrogen, phosphorus and carbon dioxide) and cost of cleaning are considered as raw materials, with quantities obtained from mass balances and prices from suppliers (Supplementary Table 4). For the base cases, all natural seawater used in the process is treated afterwards as wastewater; recycling of culture medium is not done for the base case scenarios. Energy cost is estimated as the product of the total power consumption and the location specific electricity supply cost (Supplementary Table 1).

Number of employees, standard workweek hours, employer's contribution, rank, assigned salary and cost of supervision are the factors used to estimate labor cost at each location (Supplementary Table 1). 10 workers for the operation of a 1 hectare production facility has been considered a logical value, breaking down in 1 plant manager, 1 supervisor and 8 operators of different skill levels. The relationship between labor requirements and size is not linear, therefore, according to the 0.25 power of the capacity ratio often used to scale up labor [8] 32 workers are needed in the 100 hectares facility (1 plant manager, 3 supervisors and 28 operators).

<u>Technical description and operation parameters of culture systems in the study</u>:

Design and operation of culture systems are based on AlgaePARC pilot facility, however systems in this study are meant for industrial scale and therefore the design is adapted accordingly when needed. More details about AlgaePARC pilot facility can be found in [14,15].

- Horizontal tubular photobioreactor:

This closed system is a serpentine tubular photobioreactor where a pump circulates the culture at a liquid velocity of 0.45 m·s$^{-1}$. The reactor is built up from standard units consisting of two straight transparent tubes connected to form a loop, which are placed on the ground. At the end of the loop, excess oxygen is removed from the culture by sparging ambient air in a separate vessel (degasser). Then, the broth returns to the transparent tube and carbon dioxide is added.

Disposable tubes of low density polyethylene with 0.057 m diameter are considered for the serpentine reactor. A horizontal distance of 0.05 m between tubes is selected for the design (volume:ground area ratio 23.8 $L \cdot m^{-2}$); similar to the system installed at AlgaePARC. Maximum length of units is limited by oxygen build up and depends on different factors that change with scenario. These factors are irradiation, flow velocity, dissolved oxygen concentration before the degasser and maximum photosynthetic rate value [16]. The maximum volumetric photosynthesis rate (mol $O_2 \cdot m^{-3} \cdot s^{-1}$) is calculated from the productivity for the maximum hourly irradiation (kg biomass$\cdot m^{-3} \cdot h^{-1}$) and photosynthetic quotient for the urea (1.11 mol$O_2 \cdot$mol assimilated $CO_2^{-1}$, calculated from the empirical formula for microalgae of $C_{106}H_{181}O_{45}N_{16}P$ [17]). Length of the two tubes constituting each standard unit is this maximum in order to minimize corners and elbows. Photo-inhibition is not considered in this estimation due to the existing limits to predict its effect on culture performance.

High partial oxygen pressure reduces algal growth, hence the required degassing. Maximum dissolved oxygen concentration before the degasser is set to 300% of oxygen saturation; higher concentrations are avoided at AlgaePARC. The gas exchange unit, where dissolved oxygen is released can be connected to several standard units. The volumetric gas-liquid mass transfer coefficient (kLaL) in the degassers is 0.08 $s^{-1}$ for 1.52 volume of air per degasser volume and time [18]. These values are fulfilled for both scales, i.e. 1 and 100 hectares.

- Vertical stacked horizontal tubular photobioreactor:

Similar to the aforementioned tubular system, this system also consists of straight transparent tubes containing the algae suspension. The tubes are made from rigid borosilicate glass stacked parallel to the ground in a vertical structure. A pump is circulating the culture liquid (liquid velocity 0.45 m $s^{-1}$) from the tubes to the degasser at the end of the loop and back to the tubes.

Units of tubes with 0.065 m diameter are connected to form the standard unit (one loop of two tubes in opposite directions). Mimicking the design at AlgaePARC, the distance between vertical stacks is set to 0.50 m, the height being 0.95 m, while the vertical distance between tubes is 0.05 m (volume:ground area ratio 47 $L \cdot m^{-2}$). One loop consists therefore of 8 vertically stacked horizontal tubes. The estimation of the maximum length for the standard unit is analogous to the horizontal tubular system. The degasser design and maximum oxygen concentration at the end of the standard unit are also identical to the previous system.

Metal poles of hot-dip galvanized steel are used as structure for the tubes (Supplementary Table 3). Steel angles with equal leg buried 0.95 m in the ground and reaching the same height as the system (0.95 m) are placed at a distance of 10 meters (equal to the length of the connection between tubes). The metal price from the stock exchange is used as reference to calculate cost of poles and is increased with a 50% as profit margin for the supplier.

- Flat panel photobioreactor:

This closed flat panel photobioreactor is mixed by air bubbling from the bottom, which prevents buildup of dissolved oxygen and provides mixing. The culture grows in a bag of polyethylene film enclosed in steel mesh casing (Supplementary Table 3). The dimensions of the panels are identical to those from AlgaePARC, being the light path in the panel 0.02 m, the height 0.50 m and the panels placed 0.25 m apart (volume:ground area ratio 37 $L \cdot m^{-2}$). The entire surface area is illuminated; the front surface is exposed to direct radiation, while diffuse and reflected light reach the back surface, improving the efficiency of light conversion. The aeration flow was set to 0.32 volume of air per culture volume and time, used by some authors [19] and within the range used at AlgaePARC (Supplementary Table 5).

- Raceway pond:

The raceway ponds in the study are open, ring-channel systems in the form of a single loop with a depth of 0.20 m (volume:ground area ratio 200 $L \cdot m^{-2}$), where the culture is circulated at a liquid velocity of 0.25 $m \cdot s^{-1}$; the same values as for the raceway pond installed at AlgaePARC. Selected dimensions for hectare-scale ponds in the simulations are 510 m length and 28 m total width, identical to a real demonstration plant [20]. Similarly to the raceway pond at the AlgaePARC pilot, the flow is accomplished by one paddle wheel per pond; in practice at large scale it may result in a less turbulent regime in the broth than for the system present at AlgaePARC due to increased length. The bottom of the pond is lined with reinforced and thermo-sealed PVC.

There is one carbonation sump per pond to promote the carbon transfer to the liquid phase and to ensure adequate carbon supply. The carbonation sump is 1 m deep and 0.65 m long with the same width as the channel; this design has been proven as appropriate for its purpose [21].

Empirical data:

Experimental data used for simulations were obtained in the pilot production systems (ground area ~25 $m^2$) at the AlgaePARC pilot facility in Bennekom, The Netherlands [15]. For the flat panel photobioreactor data was obtained with a smaller production system (ground area 2.5 $m^2$). The pilots were operated in continuous mode as chemostat between April and August 2013; Supplementary Table 5 shows the average photosynthetic efficiencies on sunlight (PE), associated dilution rates and amount of days that are used to calculate these values for the different systems.

**Supplementary Table 5: Experimental data used in the study; obtained outdoors at AlgaePARC in pilot plant production systems [15]**

| Reactor | Raceway pond | Horizontal tubular | Vertical stacked tubular | Flat panels[a] |
|---|---|---|---|---|
| Photosynthetic efficiency (% sunlight) | 1.2 | 1.5 | 2.4 | 2.7 |
| Daily dilution (%) | 16 | 25 | 27 | 27 |
| Days | 24 | 36 | 36 | 36 |
| Flow of culture ($m \cdot s^{-1}$) | 0.25 | 0.45 | 0.45 | - |
| Aeration (vvm) | - | - | - | 0.3-0.6 |

[a]2.5 $m^2$ pilot plant production system

Culture medium and carbon dioxide source:

Nutrients considered for the cost analysis of the culture medium are nitrogen (as urea) phosphorus (as triple-superphosphate) and carbon dioxide, since these are main components of biomass and have most impact on the economics compared to other elements.

Nutrient concentrations in the culture medium, and therefore the cost (Supplementary Table 4), are calculated separately for each case, based on biomass concentration and biomass composition. Biomass composition used for the simulation is the empirical formula for microalgae of $C_{106}H_{181}O_{45}N_{16}P$ [17].

Commercial carbon dioxide is the source of carbon in this work. The amount of carbon dioxide needed for biomass production is directly calculated from the productivity, considering a $CO_2$:biomass ratio of 1.87 derived from the considered elemental composition [17]. Carbon, nitrogen and phosphorus losses are neglected.

<u>Temperature control</u>:

Maximum culture temperature in closed systems is kept at 30°C, similar to the operational strategy at AlgaePARC. Temperature control for simulations is performed by a combination of heat exchangers and cooling water from the sea. In cases where temperature control is needed, the cooling water is pumped through the in-the-culture-submerged heat exchangers. The heat flow in the photobioreactors and the expected temperature of the culture are calculated on an hourly basis. The temperature control units are active during those periods with an expected value above the setpoint.

Cooling water from the sea comes from a depth of 200 m for all locations, excepting The Netherlands where depth used is 20 m, as in this location surface water is colder. Temperature of water considered as given by the National Centers for Environmental Information, National Oceanic and Atmospheric Administration (www.nodc.noaa.gov).

Heat flows are calculated according to [22–24]. Irradiance, radiation and convection are the factors considered in the analysis; in the open system the effect of evaporation and condensation are estimated in addition. Less influential heat flows generated from algae growth and conduction from the ground or evaporation and condensation in closed systems are neglected. For this analysis, the light falling on the ground surface is assumed to be completely absorbed in all systems.

Rate of temperature change during a certain hour results from heat flow during the interval (in Watts) divided by the product of specific heat of water ($C_p$ = 4,186 $J \cdot Kg^{-1} \cdot °C^{-1}$), seawater density ($\rho$ = 1,027 $kg \cdot m^{-3}$) and total culture volume. This temperature change is added to culture temperature of the previous hour to determine the current temperature. Initial culture temperature at time zero equals dry bulb temperature of the surrounding air. Once the expected temperature is known, the energy to lower it to 30°C is calculated (in W):

$$Energy\ to\ remove\ from\ the\ culture = \frac{(Expected\ temperature\ in\ °C - 30°C) \times C_P \times \rho \times Culture\ volume}{Time} \quad \textit{(Eq. 1)}$$

The mass flow of cooling water needed to remove that energy from the culture in the heat exchanger comes from (in $m^3 \cdot h^{-1}$):

$$Mass\ flow\ of\ cooling\ water = \frac{Energy\ to\ remove\ from\ the\ culture}{C_P \times \Delta T \times \rho} \quad \textit{(Eq. 2)}$$

Where ΔT is the difference between *Temperature cooling water*$_{out}$ and *Temperature cooling water*$_{in}$.

*Temperature cooling water$_{in}$* is the temperature of the sea water. *Temperature cooling water$_{out}$* is the temperature of the cooling water leaving the heat exchanger and discharged again to the sea. It is calculated using the following equation and considering an efficiency of heat exchange of 75% [25]:

$$Temp.\,cooling\ water_{out} = Temp.\,cooling\ water_{in} + Efficiency \times \left(\frac{Temp.of\ the\ culture}{Temp.cooling\ water_{in}}\right) \qquad (Eq.\ 3)$$

The rate of heat transfer in the heat exchangers is (in $J \cdot h^{-1}$):

$$Rate\ of\ heat\ transfer = Mass\ flow\ of\ cooling\ water \cdot Specific\ heat\ of\ water \cdot \Delta T \qquad (Eq.\ 4)$$

The following equation gives the total area of heat exchangers (in $m^2$):

$$Area\ of\ heat\ exchangers = \frac{Rate\ of\ heat\ transfer}{Heat\ transfer\ coefficient \times \Delta T} \qquad (Eq.\ 5)$$

Heat transfer coefficient in heat exchangers is estimated as 852 $W \cdot m^{-2} \cdot {}^{\circ}C^{-1}$ [26].

Cooling involves CAPEX (from cost of pumps, heat exchangers and pipes) and OPEX (energy consumption, cost of chemical treatment of the water and maintenance).

The cost for heat exchanger is derived from [27], for a shell and tube U-tube type heat exchanger, made of cs/316 stainless with a working pressure below 4 bars. When the required area of heat exchangers is above 1,115 $m^2$ the estimation may not be realistic according to the source. In that case several units of 1,115 $m^2$ are used.

Area of heat exchangers and mass flow of cooling water are calculated per hour (from Eq. 2 and 5); maximum values for each case are used in the estimation of CAPEX, i.e. for summer.

Energy consumption for temperature control is the energy used by pumps. Pumps number 5 from Supplementary Table 3 are used here with a shaft power calculated assuming 3 m of water column pressure, using 1,027 $kg \cdot m^{-3}$ as seawater density and a pump efficiency of 75% [26]. Cost of chemically treating the water is 0.004 $€ \cdot m^{-3}$ of cooling water [28].

As an alternative, use of wet cooling towers as source of cooling water is also studied. In this case cooling water passes on demand through the cooling tower and evaporation lowers its temperature. Surface water is used for this, which is continuously recirculated. This avoids catchment of deep waters and reduces the amount of seawater used. The water is also passed through the heat exchangers (using identical specifications as abovementioned for area and cost).

The climatic conditions in Curaçao (high relative humidity and air temperature) make unfeasible the use of cooling towers to keep culture temperature below 30 °C. Therefore this option is not studied in Curaçao.

In Eq. 2, *temperature cooling water$_{in}$* is the temperature of the cooling water leaving the cooling tower and entering the heat exchanger instead. It is 2.8 °C above the wet bulb temperature for the time and location. 2.8°C is the approach of the cooling tower (difference in temperature between the cooled-water temperature and the entering-air wet bulb temperature) [28].

Similarly, when using towers, *temperature cooling water$_{out}$* in Eq. 2 is the temperature of the cooling water leaving the heat exchanger and entering the cooling tower. It is also calculated considering an efficiency of heat exchange of 75% [25].

If cooling towers are involved, cooling involves CAPEX (from cost of cooling tower, heat exchangers, pipes and additional equipment) and OPEX (energy consumption and cost of water, including chemical treatment and replacement of lost water).

The technology used in cooling towers for 1 and 100 hectares facility changes with scale: mechanical draft towers and hyperbolic natural draft towers are used respectively. Mass flow of cooling water for 1 hectare facility is within the range that allows the calculation of cost of the tower according to [27] for all the cases (between 4,542 and 341 $m^3 \cdot h^{-1}$). Consequently this method is used for a tower made of redwood (a cost effective material and in abundant supply) and a factor "$f$" calculated in each particular case depending on the temperature difference of water entering and leaving the cooling tower. Non-installed price was obtained reducing the value a 20% according to the same source [27]. Mass flow of cooling water and temperature range for 100 hectares facility is within the scale of the cooling tower under study in [29], where 5,735,294 € is the total investment cost for a cooling tower with a flow of cooling water of 70,000 $m^3 \cdot h^{-1}$. Based on this relation and using the mentioned tower as basic unit size, the cost of cooling towers is scaled up or down using the following exponential law [7]:

$$CostB = CostA \times \left(\frac{SizeB}{SizeA}\right)^{0.85} \qquad \textit{(Eq. 6)}$$

When using cooling towers, energy consumption for temperature control (equipment) and cost of water (chemical treatment and make-up of losses) are 0.40 $kWh \cdot m^{-3}$ and 0.004 $€ \cdot m^{-3}$ of cooling water respectively [28].

Length of piping used to transport cooling water from the sea to the systems and back is assumed to be 600 m. If cooling towers are not used, water uptake requires 4,000 m extra of pipes to reach the required depth (400 m in case of The Netherlands). A pipe of precast concrete with a diameter of 2 m is required, at the cost indicated in Supplementary Table 3.

Cleaning:

Closed reactors operating for long times tend to accumulate an algae film in the inner surface, restricting sunlight supply. Open systems can accumulate external material as consequence of winds and animals. Besides, culture contamination with undesired species or pathogens occurs, especially in open systems. Therefore, cleaning is a necessary task to perform in a microalgae production facility. In the projections, three cleanings per year take place, similar way to operation at AlgaePARC.

Chemical cleaning is used for closed reactors, where systems are filled with 3% of a cleaning solution composed of 35% $H_2O_2$ and 7.5% glycerin (Supplementary Table 4). Small plastic granulates [14] are added in addition to tubular systems at a concentration of 0.5 $Kg \cdot m^{-3}$ (Supplementary Table 4), which are recovered afterwards and can be reused for 3 years. These granulates are also used in the broth during cultivation in tubular systems to prevent biofilm formation [14].

Open ponds are cleaned using compact road sweepers (Supplementary Table 4), which vacuum clean the bottom of the pond. They are rented at a price of 409 $€ \cdot day^{-1}$ and cover 10.8 hectare per day (based on general information available on websites from rental companies).

Inoculum production:

10% of area in the production facility is dedicated to inoculum production to supply biomass, free of contamination, to the systems. This area is considered identical to the rest of facility in terms of costs (OPEX and CAPEX). However, since this biomass is not continuously produced and is not harvested, but transferred to the continuous systems, this area is assumed as non-productive. Consequently, average photosynthetic efficiencies or productivities for the whole facility (Supplementary Table 5) are reduced with this percentage.

Power requirement for liquid circulation in tubular systems and raceway pond:

The power consumption to maintain the flow in tubular systems is calculated according to standard hydrodynamic principles. In tubular systems, the shaft power is calculated using 1,027 $kg \cdot m^{-3}$ as seawater density and a pump efficiency of 75% [26]. Energy dissipated due to major losses from friction and minor losses from bends is calculated and added to the total energy of the system [30,31]. The Darcy Weisbach equation for energy loss in turbulent regime is used for friction, as well as the Swamee-Jain Equation for friction factor. The roughness of tubes, affecting friction, is $3 \cdot 10^{-7}$ m and each elbow leads to an equivalent length of 30 (information from providers). Other minor head losses are neglected.

Energy required to mix the open ponds is estimated according to [30,32]. The total electrical efficiency of paddle wheels in open ponds is assumed to be 15% [33]. Values of 3.2, 1.8 and 0.3 are used for bend loss coefficients, drag coefficients for paddles and slippage factors respectively. Manning's friction coefficient of 0.012 $s \cdot m^{-1/3}$ is selected for the lining (thermosealed PVC). Consequently energy needed to overcome head losses in bends, curves and friction are considered in the open system, on the other hand those head losses from carbonation sump, rising bubbles (carbonation) and effect caused by winds are neglected.

Optimization – Future scenarios:

Different parameters are changed, from the original case for 100 hectares facility, to values expected to be feasible in the future. Future projections are done for flat panel systems located in south of Spain considering the following assumptions (all of these assumptions need confirmation):

- An increase of the average photosynthetic efficiency by a factor 2.22, resulting in an efficiency of 6%. The 6% photosynthetic efficiency in flat panel has already been obtained at lab-scale [34] and is still below the theoretical maximum of 8 to 10% [35].
- Maximum temperature in the culture is kept at 45°C. Algae able to grow at 50°C have already been identified [36].
- Light path of the flat panel is reduced to 0.01 m.
- Flue gas is used as source of carbon dioxide. A price of 29 $€ \cdot ton^{-1}$ is considered (Supplementary Table 4), which accounts for all upstream operations to concentrate $CO_2$ from the flue gas stream and make it usable in the process [37]. Transportation cost for this gas is also included, based on a recent study [38]. This cost is calculated conservatively, using 180 km as supply distance; the largest from [38]. Therefore, transportation in pipelines involves an energy of 13.68 kWh/ton of $CO_2$ [38]. Energy cost for the studied location (Supplementary Table 1) will then bring the transportation cost for flue gas.
- Aeration in is lowered during the night from 0.32 volume of air per liquid volume and time (vvm) to 0.05 vvm; a flow already used outdoors in a previous study [39]. Similarly, the flow is reduced during the day to 0.22 vvm, value used by [40].

- 310 operational days and only one cleaning performed per year.
- Number of employees is reduced to one person per 10 Ha (1 plant manager, 1 supervisor and 8 operators for 100 hectares); this number has already been mentioned in a previous study [7].
- The fraction of the facility used to prepare inoculum is reduced from 10 to 5% of the total area.
- Wastewater treatment is avoided. Pollutants, such as nitrogen and phosphorus, are below discharge limits in the wasted medium and can be discharged after harvesting.
- Harvesting is performed by microfiltration and subsequent centrifugation (Supplementary Figure 1). This combination has shown to be more cost effective than a single-step centrifugation [41]. Both methods are conventional in industry, showing high reliability and robustness when changes in the culture appear compared to other methods such as coagulation-flocculation-decantation or dissolved air flotation. By using membrane filtration, the biomass is concentrated about 15 times. The concentrated fraction (retentate) is further processed by the more expensive and energy-intensive centrifugation, giving a slurry with a final concentration of 15% dry solid biomass [41]. 32 $L \cdot m^{-2} \cdot h^{-1}$ has been a proven flux in microalgae cultures that combines the advantages of reasonable filtration with low fouling [41] and was therefore selected.
- Polyethylene plastic films in flat panels can be used for 2 years.
- 30% of nutrients added to the culture are reclaimed, after biomass is refined.

- **BIOREFINERY:**

<u>Biomass composition</u>

*Nannochloropsis sp.* is adopted as standard strain. The biomass composition at the end of the cultivation carried out under no nitrogen limitation has been retrieved from several literature contribution and shown in Supplementary Table 6 [42–51]. This biomass composition is typical from non-genetically modified Nannochloropsis sp.; this study does not consider GMO. Lipids are classified in polar lipids as glyco- (GL) and phospho-lipids (PL) and non-polar lipids as triacylglycerides (TAG), waxes and sterols. The composition of the saponifiable fraction (GL, PL, and TAG) is also provided in fatty acid percentage as saturated (SFA), monounsaturated (MUFA) and polyunsaturated (PUFA) fatty acids. Proteins are divided in water soluble and non-water soluble fractions. Carbohydrates are classified as mono- and polysaccharides. Ash content and pigments are taken into account. Amino-sugars and nucleic acids are not considered, as well as the pigment composition of the microalgal biomass. Table 6 provides some details about the most important physicochemical properties. A detailed list of any potential market application of each component is provided in Supplementary Table 11 and later discussed.

**Supplementary Table 6: Biomass composition used in the study**

| Component | Percentage | | | | Main assumed physicochemical properties |
|---|---|---|---|---|---|
| Lipids | 20% | SFA* | MUFA* | PUFA* | |
| Glyco-, Phospho-lipids | 12% | 35% | 30% | 35% | Membrane lipids, water insoluble, reference molecule: Phosphatidylglycerol. Volatility and density estimated by group contribution method [52]. |
| Triacylglycerides | 2% | | | | Intracellular lipids, water insoluble, reference molecule: Triolein. Volatility, and density estimated by group contribution method [52]. |
| Waxes | 3% | | | | Membrane and intracellular lipids, water insoluble, reference molecule: squalene. Volatility and density estimated by group contribution methods [52]. |
| Sterols | 3% | | | | Intracellular lipids, water insoluble, reference molecule: cholesterols. Volatility and density estimated by group contribution methods [52]. |
| Proteins | 50% | | | | |
| Water soluble | 20% | | | | Cytosolic proteins, water soluble, reference molecule: Rubisco, negligible volatility. |
| Non-water soluble | 30% | | | | Structural proteins, mainly present in cell debris, water insoluble, negligible volatility. No specific reference compound, since they do not participate in thermodynamic equilibrium. Molecular weight assumed larger than 1,000 kDa. |
| 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% | | | | Poorly water soluble, negligible volatility, reference component: lutein. |
| Ashes | 7% | | | | Water soluble, negligible volatility, no reference component. |

* As percentage of glycolipids, phospholipids and triacylglycerides

Process Design

The biomass addressed in the downstream process is a slurry with 15% dry weight concentration, as mentioned in the previous section. Biomass output for each case is produced in the cultivation system for the particular projection.

A microalgal biorefinery process is designed to fractionate biomass into the main components: proteins, lipids, carbohydrates and pigments. Furthermore, in the most complete scenario studies saponifiable lipids are converted into fatty acid methyl esters (FAME) and fractionated afterwards as saturated, mono-unsaturated and poly-unsaturated. On the other hand proteins are fractionated into water soluble and non-water soluble fractions. The analysis of the biorefinery process chain is performed in a benchmark flow-sheet: it involves conventional unit operations adopted in biotechnology and/or chemical engineering fields for cell disruption, extraction, fractionation and purification.

According to this scenario, the most general biorefinery is designed in SuperPro Designer® v9.0 and visualized in Supplementary Figure 2. The process consists of four sections:

I. Cell disruption
II. Extraction and fractionation of water soluble compounds such as water soluble proteins, mono- and poly-saccharides
III. Extraction and fractionation of hydrophobic compounds such as TAG, GL, PL, waxes, sterols and pigments
IV. Exploitation of the residual cell debris consisting mainly of non-water soluble proteins.

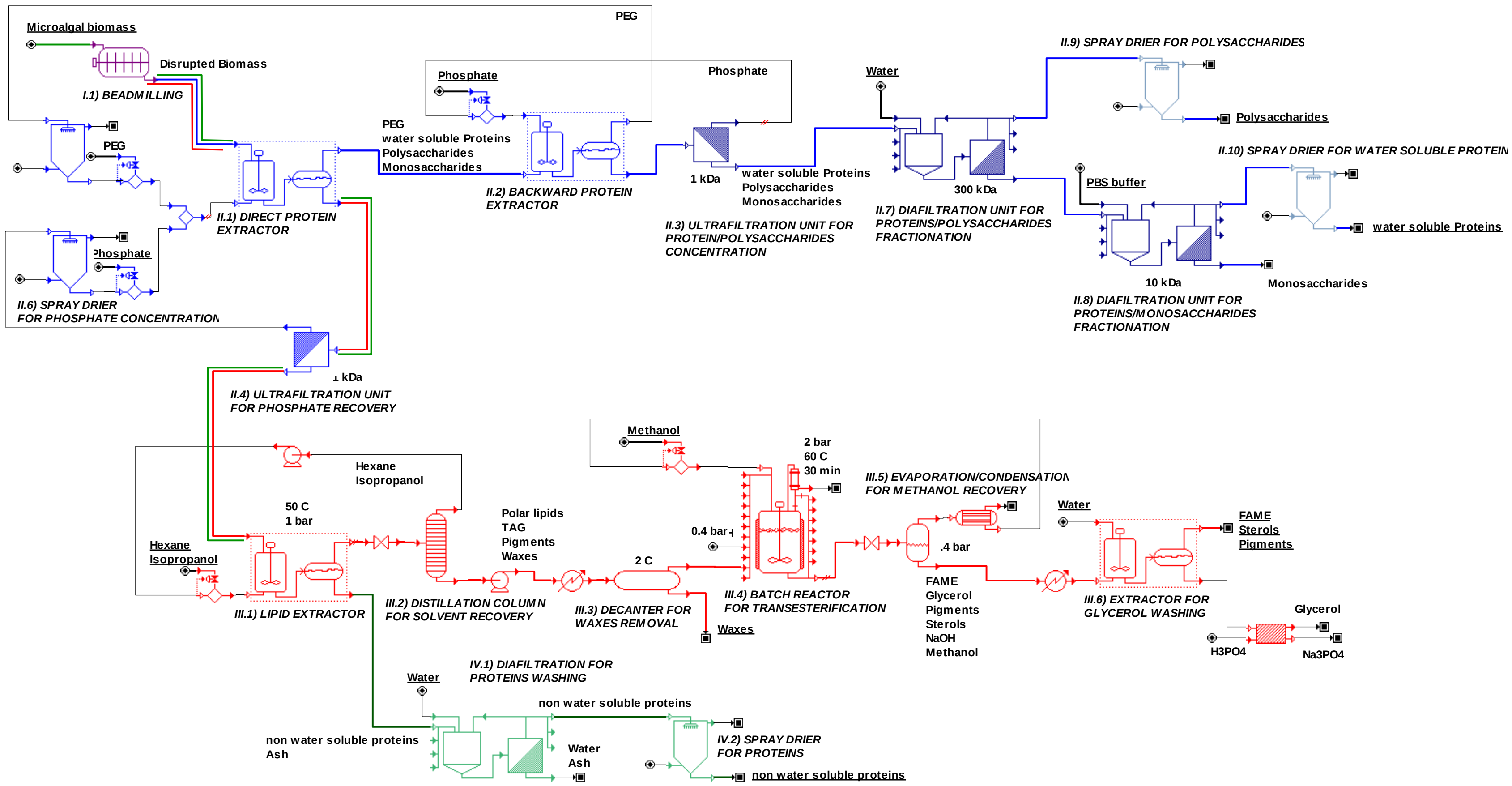
Operating conditions for all the unit operations were set according to realistic values for industrial process.

Section I – Cell disruption

Cell disruption is a required pretreatment to improve the extraction efficiency of intracellular compounds. Mechanical cell disruption represents the industrial benchmark, achieving high yields with high energy consumption. Bead-milling and high pressure homogenization represent mechanical cell disruption technologies using solid- and liquid-shear forces, respectively [53,54]. Since the biomass dry weight achieved during harvesting in the microalgal facility is quite high (15% DW) bead-milling was selected for cell disruption (I.1), being a much more suitable technology in case of highly viscous streams [55]. During cell disruption temperature is kept at 25 °C and all the consumed energy is assumed to be dissipated as heat. The cell disruption efficiency is set to 95% due to constraints in time and energy to achieve a complete cell disruption. At a higher amount of cell disruption, individual cells would have a similar probability of being disrupted and as the process continues a higher amount of energy would be consumed by the disruption of cell fragments rather than for disruption of intact cells [53]. The cell disruption has been assumed as a simple reaction process in which intact biomass becomes a blend of separated components according to the table 6.

Section II – Extraction and fractionation of water soluble components

To extract water soluble proteins and carbohydrates from the biomass, a direct aqueous extraction (II.1) followed by a back aqueous extraction stage (II.2) are done. This aqueous two phase system (ATPS) is performed by adding polyethylene glycol (PEG4000) and potassium phosphate at 26% and 15% weight fraction respectively. With these operating conditions the system separates during the direct extraction in a light aqueous phase mainly containing PEG400 and a heavy aqueous phase being almost exclusively potassium phosphate [56–61]. Equilibrium conditions of the ATPS found by [62] are adopted. According to [56] the partition coefficient of the protein Rubisco (Ribulose-1,5-bisphosphate carboxylase/oxygenase) in ATPS can be more than 10 at pH 7. This value is assumed as partitioning coefficient for all the water soluble proteins.




**Supplementary Figure 2 – Flowsheet of the complete biorefinery –Section I – Purple; Section II – Blue; Section III - Red; Section IV - Green**

**Blue line: path of water soluble components - Red line: path of lipid components; Green line: path of non-water soluble components**

During the direct extraction of soluble proteins and other hydrophilic compounds (mono- and polysaccharides) are mainly extracted by PEG400 and therefore allocated in the top phase. Lipids (TAG, GL, PL, waxes and sterols) and pigments are mainly segregated from the two aqueous phases in the bottom phase together with non-water soluble proteins and ash. Due to the sufficient density difference between the two aqueous phases, a mixer-settler configuration is adopted for both the direct and back extraction unit with fixed residence times in the mixer and settler of 5 and 30 minutes, respectively. The number of theoretical stages in both cases is selected in order to have a degree of extraction of water soluble protein always higher than 95%.

In the direct ATPS extraction (II.1) a 1:1 ratio is selected for the volumetric partitioning of water in the two phases. After the direct ATPS extraction, a back extraction (II.2) is designed to transfer the proteins, mono- and poly-saccharides to the salt-rich phase. To have favorable operating conditions a seven times higher concentrated salt-rich phase in comparison with the first extraction step is used. The overall amount of energy for mixing is set to 0.5 $kW \cdot m^3$ for both extraction steps, assumed to be totally dissipated into heat [63]. Cooling water is used to maintain the temperature at 25 °C.

Since the ATPS are characterized by large quantities of chemicals (PEG at 26%w and phosphate at 15%w) it is crucial to efficiently recycle them. The phosphate rich phase coming from the direct extraction is concentrated by means of ultrafiltration (UF) units (II.4) with 1 kDa membranes. In UF II.3 the feed is concentrated at least five times in order to recover more than 80% of the phosphate from the permeate. A further concentration carried out in spray driers (II.5) after the phosphate is recycled to the direct extraction (II.1). PEG rich phase from the back extraction is also recycled to the direct extraction phase prior to a further concentration in another spray drier (II.5).

Both the spray driers (II.5 and II.6) operate at a ratio of air/evaporated water equal to five, achieving a water loss larger than 99%. Operation is performed at standard condition based on [64]: the evaporation rate set to 100 $kg \cdot h^{-1} \cdot m^{-3}$, temperature kept constant at 40 °C and the absorbed power set at 0.02 $kWh \cdot kg^{-1}$ of feed.

After the back extraction the phosphate rich phase is treated in a UF unit (II.3) where the feed is concentrated at least 40 times to recover more than 95% of the phosphate from the back extraction (II.2).

Further fractionation of the carbohydrates from proteins is performed by a sequence of two diafiltration (DF) units with different membrane cut-off (II.7 and II.8). According to [65] a DF membrane characterized by a 300 kDa cut off value is used to fractionate and concentrate large polysaccharides and permeate consists of proteins and monosaccharides. . In the DF (II.7), the polysaccharides are washed out ten times with water from the protein fraction and monosaccharides. The next DF step (II.8) is performed with a 10 kDa cut off membrane which retains and concentrates proteins and permeate monosaccharides. Here, a five times washing with a phosphate buffer solution is used to remove monosaccharides from the protein fraction. In both cases the DF units are operated at standard conditions [66]: flux set to 40 $L \cdot m^{-2} \cdot h^{-1}$, power consumption to 0.2 $kW \cdot m^{-2}$, and heat dissipation to 10% of the power consumption, with a maximum operational timeframe of the membrane of 1,000 hours.

The final achieved concentration of soluble proteins and polysaccharides is within 14-20% dry weight. Two spray drying units are selected to complete the purification of both the proteins (II.10) and polysaccharides (II.9). A ratio of air/evaporated water of 5 and an evaporation rate of

100 kg·h$^{-1}$·m$^{-3}$ are assumed. The temperature is kept constant at 40 °C by heat exchanges and the absorbed power is set at 0.02 kWh·kg$^{-1}$ of feed.

Section III – Extraction and fractionation of hydrophobic components

After the direct aqueous extraction, the lipid phase present in with the salt-rich aqueous phase from the UF (II.4) is extracted.

The extraction (III.1) is carried out with a mixture of hexane/isopropanol (1:4) in a ten times higher amount of volume with respect to the water phase. Extraction was performed at 50 °C, according to [67], among different combinations of hexane, methanol and isopropanol; the extraction of lipids is optimized at this temperature.

Also in this case, a mixer-settler (III.1) configuration with similar operating conditions to the ATPS is selected and designed. Number of theoretical contact stages is set up in order to have a minimum degree of lipid extraction close to 85%. The organic solvents are then recovered by distillation (III.2) while keeping the temperature at the bottom of the distillation column below 250 °C. Temperatures above 250 °C have to be avoided to prevent lipid degradation [68,69]. The distillation unit is operated with a boiler reflux ratio of $R/R^{min} = 1.25$, under vacuum conditions (p=0.4 atm) with a condensing temperature of 60 °C at the top and a reboiler temperature at the bottom equal to 200 °C as reported in [69]. The efficiency of the stages is set at 80%. The polar lipid purity was calculated at 80% accordingly and, consequently, the organic solvents fraction is not recycled to the extraction unit.

The modified Raoult's law is adopted to calculate thermodynamic parameters using the activity coefficient according to the Wilson model. Since some binary interaction parameters are not available in the simulation databanks, these are estimated using the UNIFAC vapor–liquid equilibrium (VLE) and UNIFAC liquid–liquid equilibrium (LLE) models [52,70,71]. Palmitic acid (C16:0), linoleic acid (C18:1) and eicosapentaenoic acid (C20:5) are adopted as model components for saturated, mono-unsaturated and poly-unsaturated fatty acids for the calculation of the thermodynamics of TAG, GL, PL and FAME.

The extracted lipid fraction is subjected to a de-waxing step. A winterisation process (III.3) is adopted to separate the waxes from the other lipids: the lipids are gradually cooled to 2 °C. The cooled lipids are kept at this temperature for 5-10 hours prior to the separation of solid waxes from the liquid oils by decanting the oil-solid fat slurry [72,73].

Transesterification (III.4) of the saponifiable fraction of extracted lipids (TAG, GL and PL) into fatty acids methyl esters (FAME) is carried out under alkaline conditions at 60 ºC with 1% of NaOH and a methanol excess of 300%. Under these conditions the residence time to achieve significant conversion of triglycerides into fatty acids is about 10 min [71–73]. To assure a 99% conversion of extracted lipids the residence time was fixed to 60 min. The excessive amount of methanol is recovered by evaporation (III.5) and further condensation. Methanol recovery is set to 95%. Evaporation is carried out under vacuum conditions (0.4 bars) to avoid glycerol degradation. In this way the bottom temperature of the distillation unit is below 150 °C [69,74].

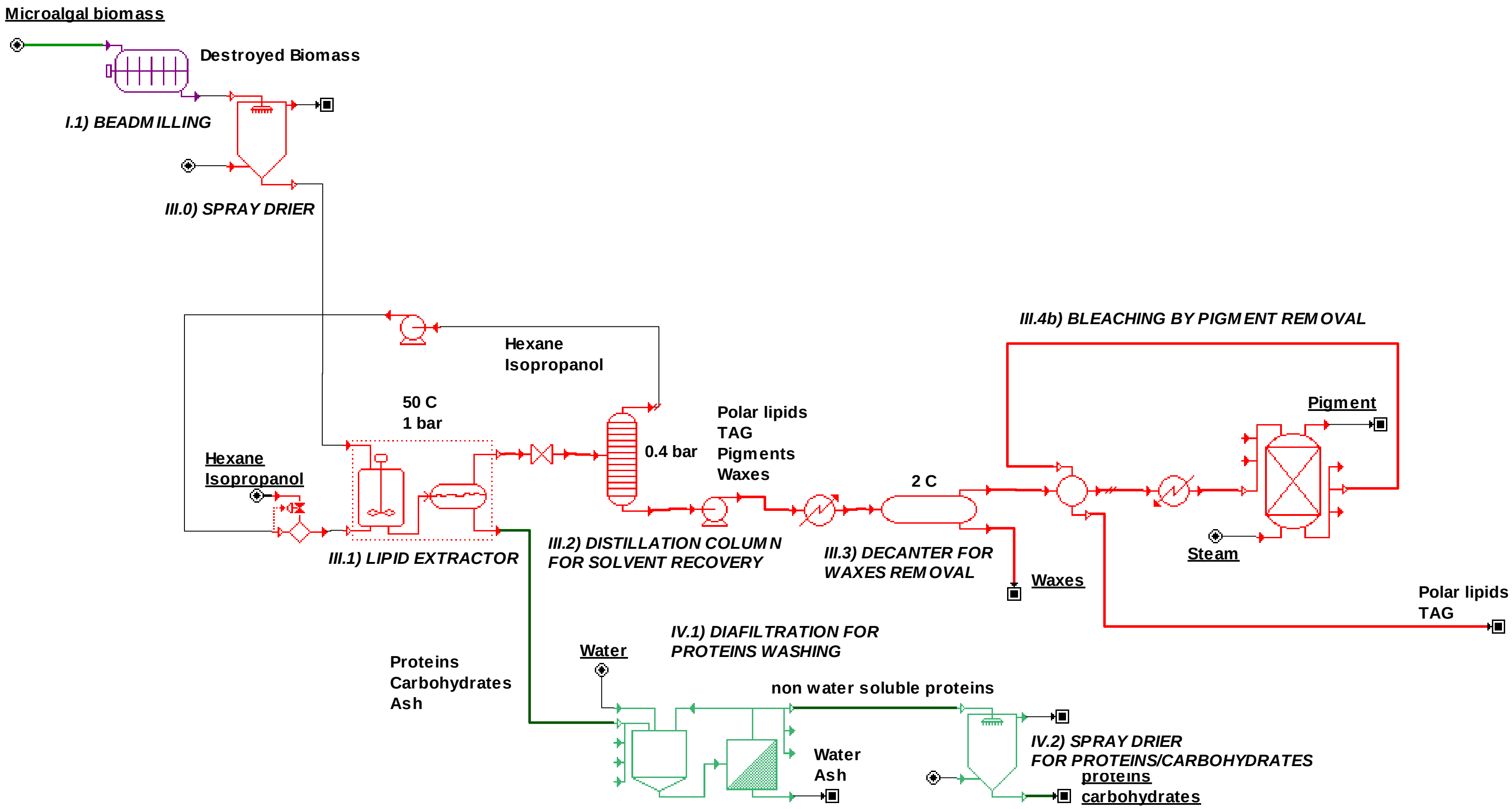
Glycerol is then washed out by an equal amount of water in a continuous extraction unit (III.6), obtaining a separate FAME, sterols and pigments fractions.

Section IV – Recovery of non-water soluble proteins

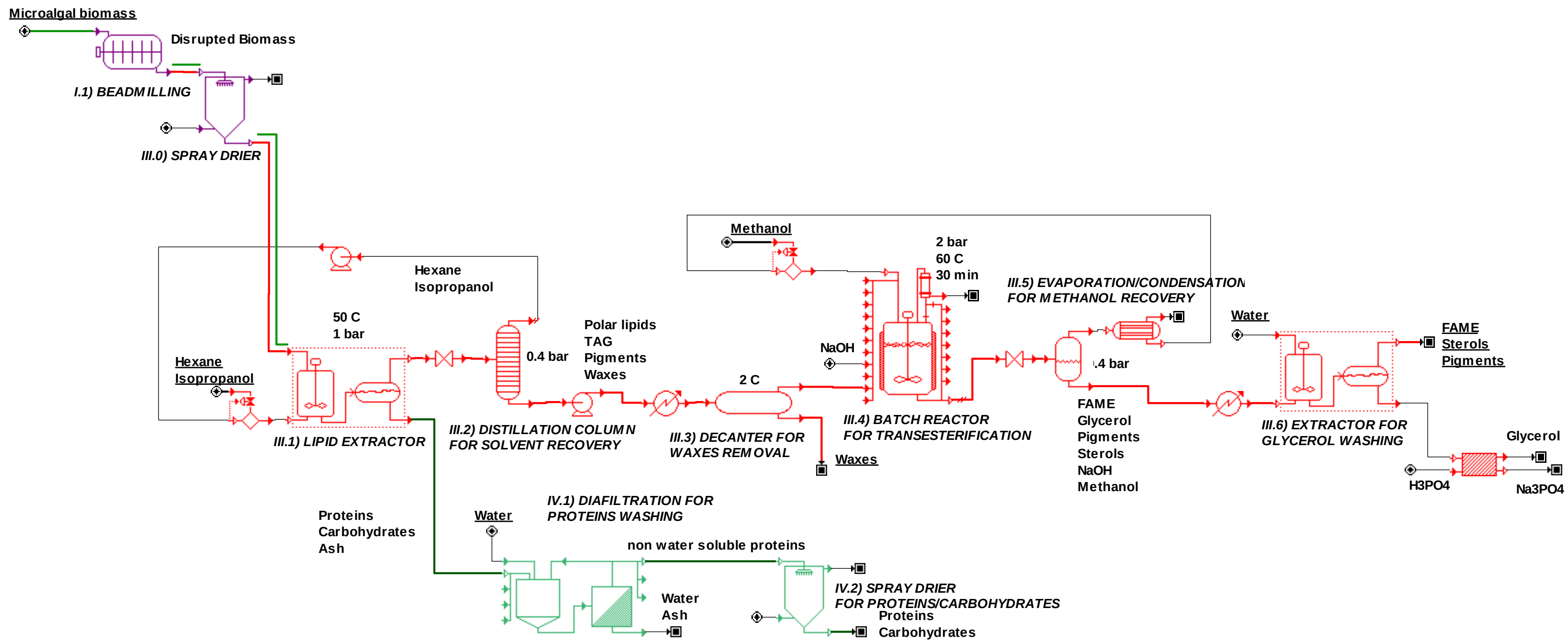
A DF (IV.1) unit is used to wash out the salts and to concentrate the non-water soluble proteins phase. The operating conditions were the same to conditions described in section II.8. The upper limit of non-water soluble protein concentration in the pre-concentration step was set to 200 $g \cdot L^{-1}$ in the DF. A fivefold washing is carried out during diafiltration with tap water in order to reduce the residual ash content. A drying step of the proteins was designed using a spray dryer (IV.2) as in II.10.

The general flowsheet (Supplementary Fig. 2) is designed to obtain the most complete level of fractionation of the separate biomass components. In particular to exploit the full potential of the biomass to address all the high value components. In case of the production of a subset of them the flow sheet can be derived by taking out the unnecessary units/sections.

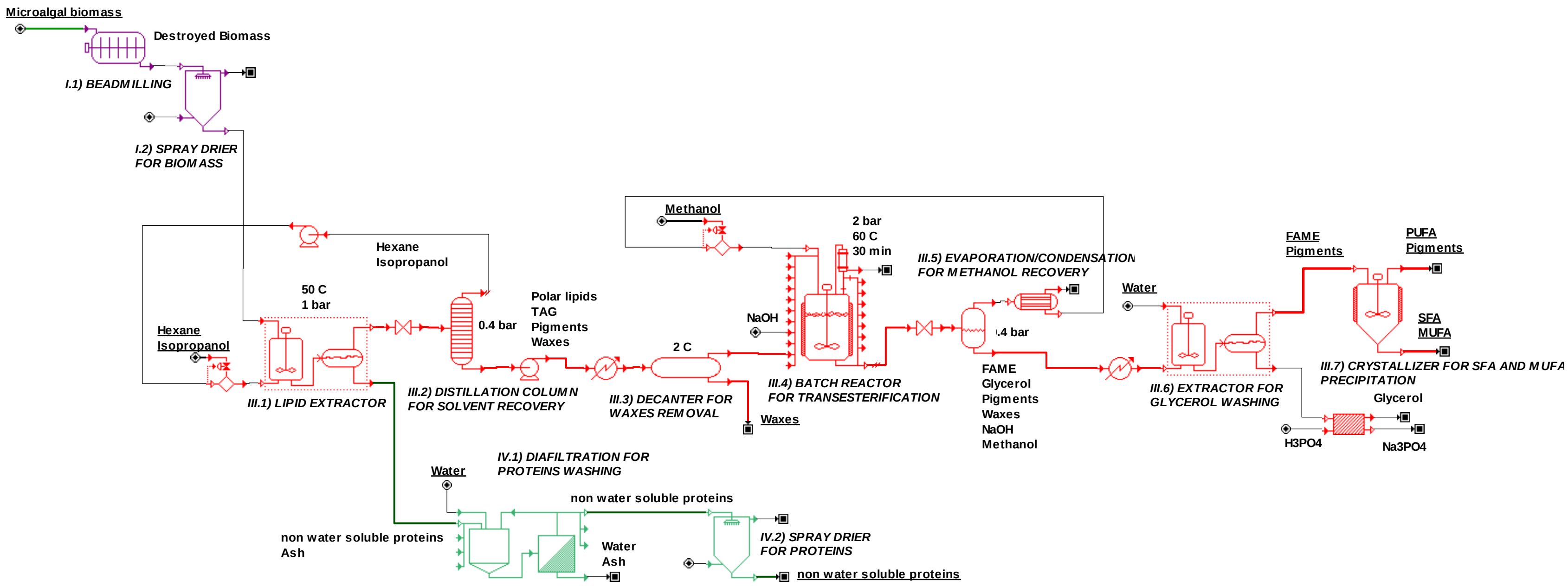
Supplementary Figure 3 shows more simple flow sheets in case of food/feed commodities (Supplementary Fig. 3.A) and biofuel production (Supplementary Fig. 3.B). In case of food/feed application, oil is not transesterified after de-waxing. A bleaching step (III.4b) is added to remove pigments by adsorption on activated clays [75,76].



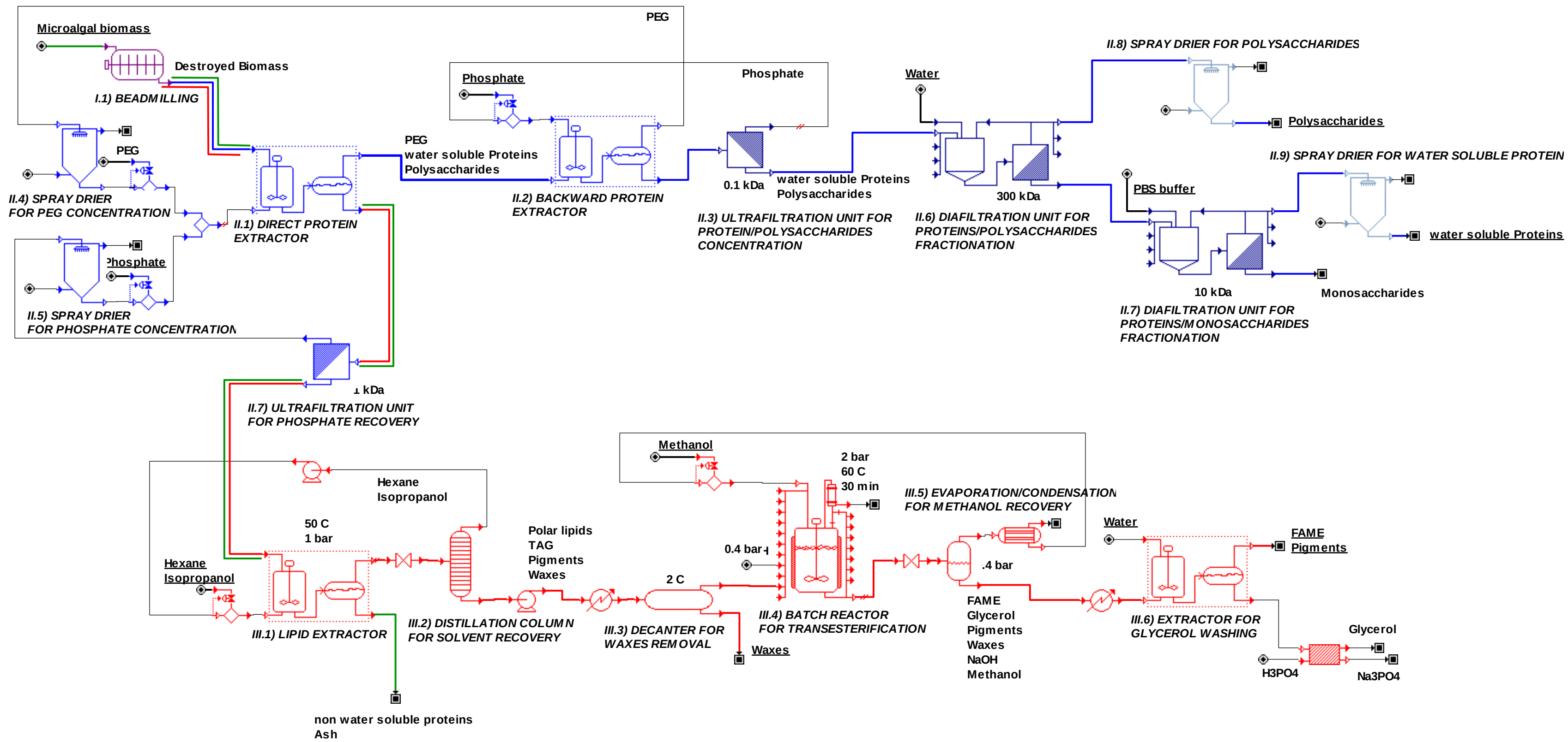

**Supplementary Figure 3.A – Flow sheet of the biorefinery for food/feed application –Red line: path of lipid components - Green line: path of hydrophilic components**



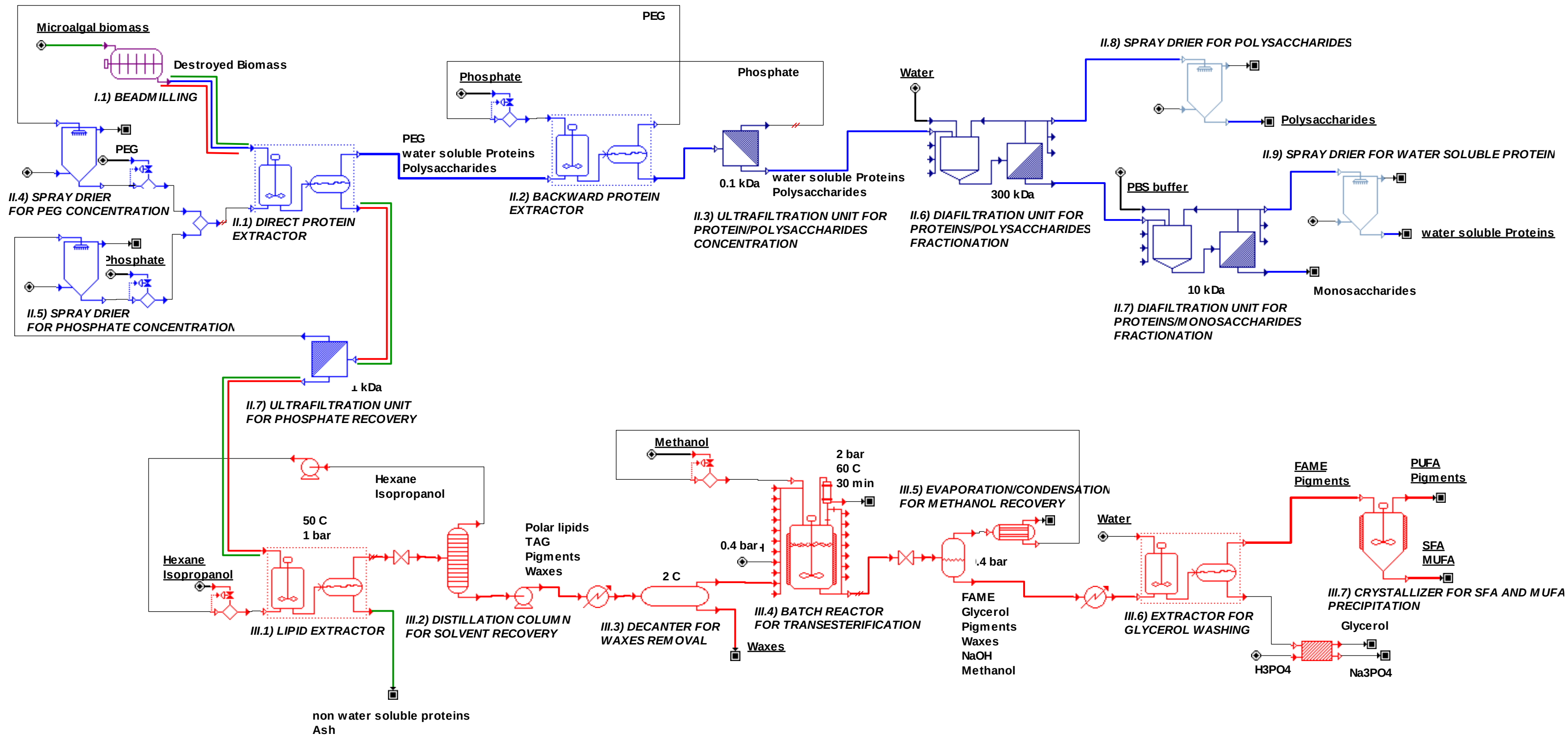

**Supplementary Figure 3.B – Flowsheet of the biorefinery for biofuel –Red line: path of lipid components - Green line: path of hydrophilic components**




**Supplementary Figure 3.C – Flowsheet of the biorefinery for Chemicals –Red line: path of lipid components - Green line: path of hydrophilic components**




**Supplementary Figure 3.D – Flowsheet of the biorefinery for Food additives – Green Blue line: path of water soluble components - Red line: path of lipid components; Green line: path of non-water soluble components**




**Supplementary Figure 3.E – Flowsheet of the biorefinery for Cosmetics/Healtcare – Green Blue line: path of water soluble components - Red line: path of lipid components; Green line: path of non-water soluble components**

Utilities:

Costs of the utilities (Supplementary Table 7) are location-dependent. In principle, costs of utilities should take into account both the CAPEX and OPEX to produce them. In this case CAPEX is assumed negligible and OPEX as linearly dependent on the amount of energy necessary to supply the utilities. Consequently the costs of both heating and cooling agents are scaled based on the cost of the energy (Supplementary Table 7).

**Supplementary Table 7: Cost and properties of utilities used in biorefinery**

| Location | | NaCl brine | Chilled water | Cooling water | Steam (low p) | Steam (high p) |
|---|---|---|---|---|---|---|
| **Energy cost** (€·ton$^{-1}$) | **Canary Islands** | 0.235 | 0.377 | 0.047 | 11.30 | 18.83 |
| | **Curaçao** | 0.592 | 0.948 | 0.118 | 28.43 | 47.38 |
| | **The Netherlands** | 0.185 | 0.296 | 0.037 | 8.89 | 14.81 |
| | **Saudi Arabia** | 0.056 | 0.09 | 0.011 | 2.68 | 4.47 |
| | **South of Spain** | 0.235 | 0.377 | 0.047 | 11.30 | 18.83 |
| | **Turkey** | 0.179 | 0.287 | 0.036 | 2.61 | 14.35 |
| **Temperature range (°C)** | | -10 - 0 | 5 - 10 | 15 - 20 | 152 | 242 |

Labor:

As the biorefinery consists mainly of continuous unit operations with high degree of automatisation the labor is initially fixed assuming a requirement of labor hour per hour of unit operation equal to 0.2 operator/0.03 supervisor/0.01 manager. According to this assumption an operator can manage up to five unit operations as reported in [9]. Supplementary Table 8 shows the labor demand for each application in terms of labor hours per year and personnel units referring to a standard workweek of 40 h.

**Supplementary Table 8: labor amount and distribution in 100 ha biorefinery**

| | | **Biofuel** | **Chemical** | **Food/feed** | **Food additives** | **Cosmetics** | **Complete biorefinery** |
|---|---|---|---|---|---|---|---|
| Operator | Labor $hr \cdot yr^{-1}$ | 28,114 | 25,432 | 18,877 | 48,171 | 49,967 | 52,628 |
| | fte | 13.5 | 12.2 | 9.1 | 23.1 | 24 | 25 |
| Supervisor | Labor $hr \cdot yr^{-1}$ | 3,960 | 3,711 | 3,040 | 7,508 | 7,893 | 8,177 |
| | fte | 1.9 | 1.8 | 1.5 | 3.6 | 3.8 | 3.9 |
| Manager | Labor $hr \cdot yr^{-1}$ | 1,354 | 1,287 | 1,048 | 2,537 | 2,611 | 2,760 |
| | fte | 0.65 | 0.6 | 0.5 | 1.2 | 1.2 | 1.3 |

Materials:

Costs of materials for the biorefinery are retrieved from ICIS [77] and from indexmundi [78]. Supplementary Table 9 gives an overview of their purchasing prices and quantities.

**Supplementary Table 9: Costs and amounts of materials in the 100 ha biorefinery**

| **Chemical** | **Cost (€$\cdot kg^{-1}$)** | **Source** | **Amount – ton$\cdot yr^{-1}$** |
|---|---|---|---|
| Hexane | 0.93 | ICIS | 5-70 |
| Isopropanol | 1.34 | ICIS | 8-100 |
| Methanol | 0.45 | ICIS | 20-360 |

| | | | |
|---|---|---|---|
| PBS buffer | 1.51 | ICIS | 30-400 |
| Phosphoric acid | 0.45 | ICIS | 0.8-10 |
| PEG400 | 1.35 | ICIS | 20-300 |
| Sodium hydroxide | 0.24 | ICIS | 0.5-6 |
| Sodium phosphate | 0.38 | ICIS | 1,440-20,000 |

Economic analysis

Simulations were conducted incorporating the costs of resources (materials, energy and utilities), and equipment in 2015. Costs of equipment and materials were mainly obtained from suppliers and when not possible from the standard database in SuperPro Designer®. The utility costs were addressed specifically for each particular case of location, biomass production and products of interest. Ranges of biomass throughputs were set according to the productivities from the cultivation.

Supplementary Table 10 shows the capacity, costs, power inputs and labor requirements for each unit operation. Since the units were sized depending on the overall throughput of the process, the values in the table are reported as a range dependent on the range of biomass output from cultivation.

**Supplementary Table 10: Details of the major equipment considered in the study for the range of biomass throughput for a 100 ha plant**

| Section | MAJOR EQUIPMENT | Section/number in the flowsheet | Capacity | k€·unit$^{-1}$ |
|---|---|---|---|---|
| I | **Beadmilling** | I.1 | 0.56-2 $m^3$ | 300-600 |
| II | **ATPS mixer-settlers** | II.1-II.2 | 0.15-9 $m^3$ (mixer)<br>1-50 $m^3$ (settler) | 120-400 |
| | **Ultrafiltration (UF)** | II.3-II.4 | 40-400 $m^2$ | 80-240 |
| | **Diafiltration (DF)** | II.7-II.8 | 18-80 $m^2$ | 50-120 |
| | **Spray driers** | II.5-II.6-II.9-II.10 | 0.15-60 $m^3$ | 90-230 |
| III | **Mixer-settler** | III.1-III.6 | 0.003-2 $m^3$ (mixer)<br>0.018-12 $m^3$ (settler) | 4.5-160 |
| | **Distillation column** | III.2 | 1-14 $m^3$ | 20-50 |
| | **Decanter+cooler** | III.3 | 0.007-0.1 $m^2$ (cooler)<br>0.56-30 $m^3$ (decanter) | 30-40 |
| | **Batch reactor** | III.4 | 25-350 $m^3$ | 380-420 |
| | **Evaporator+condenser** | III.5 | 4-60 $m^3$ (flash)<br>0.075-1 $m^2$ (condenser) | 3-4.5 |
| IV | **Diafiltration (DF)** | IV.1 | 15-75 $m^2$ | 50-120 |
| | **Spray drier** | IV.2 | 0.8-13 $m^3$ | 110-180 |

- **MARKET ANALYSIS:**

The overall turnover coming from the exploitation of the biomass changes with market application. Final products obtained from the different fractions of the biomass depend on the projected scenario [79,80]. The market analysis has been conducted looking at five different market scenarios according to the biomass value pyramid: biofuel, chemical/technical, food/feed, application in food (additives) and cosmetics and health care. In addition, a hybrid scenario has been considered aiming at maximizing the value of all the biomass components by allocating them to the most profitable application. The analysis has been carried out according to the following steps:

I. Define the market scenarios and products.

II. Identify the selling price of each product (Supplementary Table 11).

III. Allocate the biomass components to different applications according to the selected market scenario. For each scenario the most profitable application of each component of the biomass was selected in order to maximize revenues.

IV. Calculate the overall turnover from the biomass exploitation according to equation 7.

$$Turnover = \sum y_i \cdot MAX(€_{ij}) \qquad \text{(Eq. 7)}$$

Where $y_i$ is the weight fraction of the component $i$ (Supplementary Table 6), $€_{ij}$ is the value of the component i allocated to the application $j$ (Supplementary Table 11). The selling price of each component (Supplementary Table 11) has been assumed according to the market application [78,81–85].

**Supplementary Table 11: Selling price of products potentially obtained from microalgal biorefinery**

| | **Product** | **Selling price ($€\cdot Ton^{-1}$)** | **Suitable biomass component** |
|---|---|---|---|
| **Biofuel** | Biodiesel | 710 | Triacylglycerides<br>Glycolipids<br>Phospholipids |
| | Bioethanol | 370 | Carbohydrates<br>Proteins[a] |
| **Chemical/Technical** | Biopolymers | 2,000<br>2,300<br>1,400<br><br>2,000 | FAME<br>Sterols<br>water soluble and non-water soluble protein<br>Polysaccharides |
| | Biolubricants | 1,500<br>3,500<br><br>4,000 | Saturated FAME<br>Mono- and poly-unsaturated FAME<br>Wax |

| | | | |
|---|---|---|---|
| | Biopolymer additives | 2,300<br>3,700<br>15,000 | Sterols<br>Polysaccharides<br>Pigments |
| | Coatings | 4,000 | Waxes |
| | Paints | 15,000 | Pigments |
| **Food/ Feed** | Proteins | 1,100 | Water soluble protein<br>Non-water soluble proteins |
| | Lipids | 950 | Lipids |
| | Carbohydrates | 750 | Carbohydrates |
| **Food additives** | Poly-unsatured fatty acids | 5,000<br>75,000 (only EPA/DHA) | PUFA |
| | Functional protein | 3,300 | Soluble protein |
| | Pigments | 900,000 | Pigments |
| | Sterols | 45,000 | Sterols |
| **Cosmetics Health care** | Antioxidants | 30,000 | PUFA, sterols, |
| | Antioxidants | 900,000 | pigments |
| | Proteins | 3,500 | Water soluble protein |
| | Glycolipids/ Phospholipids | 6,000 | GL and PL |
| | Wax-esters | 4,000 | Waxes |
| | Sterols | 17,000 | Sterols |
| | Bioactive sulfated polysaccharides | 2,500 | Polysaccharides |
| **Complete biorefinery** | Antioxidant | 30,000 | PUFA |
| | Biolubricant | 1,500-3,500 | Saturated and<br>Mono-unsaturated FAME |
| | Food additives | 3,300 | Water soluble protein |
| | Wax-esters for cosmetics | 4,000 | Waxes |
| | Antioxidants for cosmetics | 17,000 | Sterols |
| | Biopolymer | 2,000 | Carbohydrates |
| | Food additives | 900,000 | Pigments |

[a]Conversion yield into bioethanol is 0.46

For instance, carbohydrates can be used as biopolymer, as source of biofuel for fermentation, for health and as food/feed. Instead proteins can be used as source for bioethanol, food/feed, functional proteins as food additives, for producing glues and as a source of bulk chemicals (amino acids). A hybrid and complete biorefinery scenario allocating all the biomass components to the most profitable market and application is also considered in order to maximize the overall turnover.

## - PROFITABILITY OF THE PROJECT:

An economic analysis of combined cultivation, biorefinery and market value has been performed. A production facility of 100 hectares in the current scenario for flat panels photobioreactors in south of Spain is evaluated. The economic decision tools Net Present Value (NPV) and Internal Rate of Return (IRR) are applied to evaluate the project. In addition, the Discounted Payback Period of the investment is calculated, as the years of operation required to reach the break-even point (to recoup the investment).

The construction period for the production and biorefinery facilities has been considered to be 2 years in total, where there is no production and hence no revenues. Total initial investment is distributed equally during the construction period (50% each year). During the start-up period, taking place in the first 2 years of operation (right after the construction period), biomass productivities were assumed to be 50% and 75% of the base case. Productivities are 100%, or those from the base case, the remaining 13 years. Tax rate on earnings for Spain is 23% [86]. A working capital (operating liquidity) is available for the first three months of operation as part of the initial investment and is returned at the end of the lifetime. Its amount is equal to one quarter of OPEX for such year (Supplementary Table 2), time considered to turn the net current assets and current liabilities into cash.

Net Present Value provides the overall economic value of a project, being used as an indicator to determine whether an investment will result in a net profit or a loss. NPV above 0 with appropriate risks are a good option of investment.

$$NPV\ (i, N) = \sum_{t=0}^{N} \frac{R_t}{(1+i)^t} \qquad \text{Eq. 8}$$

Where,

"$N$" total number of periods (17 periods; being 15 years of lifetime or operation plus 2 years for construction)

"$t$" is the period of the cash flow

"$i$" is the discount rate, assuming a value of 10% [87]

"$R_t$" is the net cash flow (cash inflow from selling the products minus cash outflow or total costs)

Cash inflow includes sales revenues and return of working capital at the end of the lifetime. On the other hand, cash outflow comprises investment (including working capital), operating costs and tax rate on earnings. Cash flows do not include interest payments; the discounting procedure already simulates interest payments.

On the other hand, Internal Rate of Return measures and compares the profitability on investments. It is the discount rate (previous "$i$") for which the NPV of the venture equals zero. If the discount rate of 10% considered for NPV is lower than IRR then NPV is positive and venture is profitable.

For the break-even analysis, point where revenues equals the costs, ratio CAPEX and OPEX has not been modified with respect to the base case.

## - ENERGY EFFICIENCY RATIO:

Energy Efficiency Ratio is an indicator to express the relationship between energy invested (input) and energy obtained (output). For the scenario of biofuels this was calculated as the ratio between the energy produced as biofuels and the total electricity needed for the production (energy involved in biomass production and biorefinery). Values above 1 designate net positive energy produced; i.e. more energy generated than invested. Energy obtained from the biomass as biofuels was 7.9 $MJ \cdot kg^{-1}$. Embodied energy (the sum of the total energy necessary for an entire product life-cycle) has not been considered in this calculation.

- **ANNEX:**

**Supplementary Table 12: Set of projected results for microalgae production in 100 hectares (cultivation and harvesting). *RW*: raceway pond; *HT*: horizontal tubular photobioreactor; *VT*: vertically stacked horizontal tubular photobioreactor; *FP*: flat panels photobioreactor. *CU*: Curaçao; *TN*: The Netherlands; *SP*: Spain; *CI*: Canary Islands; *TU*: Turkey; *SA*: Saudi Arabia**

| **SCENARIO** | | **RESULTS FOR MICROALGAE CULTIVATION** | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| System | Location | Biomass cost (€/kg) | Biomass Capacity (Ton/Year) | Initial Investment (M€) | Total cost (M€/Year) | CAPEX (M€/Year) | OPEX (M€/Year) | CAPEX (€/kg) | OPEX (€/kg) | Biomass concentration in culture (g/L) | Energy efficiency ratio |
| RW | TN | 11.0 | 1296 | 47.7 | 14.2 | 3.4 | 10.8 | 2.6 | 8.4 | 0.15 | 0.73 |
| RW | CI | 5.0 | 2838 | 51.2 | 14.1 | 3.6 | 10.5 | 1.3 | 3.7 | 0.30 | 1.44 |
| RW | TU | 4.7 | 2672 | 49.9 | 12.5 | 3.5 | 9.0 | 1.3 | 3.4 | 0.28 | 1.35 |
| RW | SP | 5.2 | 2708 | 49.9 | 14.0 | 3.5 | 10.4 | 1.3 | 3.9 | 0.28 | 1.37 |
| RW | SA | 4.0 | 3049 | 51.2 | 12.2 | 3.6 | 8.6 | 1.2 | 2.8 | 0.32 | 1.55 |
| HT | TN | 8.9 | 1621 | 47.8 | 14.4 | 3.4 | 11.0 | 2.1 | 6.8 | 1.01 | 0.80 |
| HT | CI | 4.8 | 3548 | 75.3 | 16.9 | 5.2 | 11.7 | 1.5 | 3.3 | 1.98 | 1.15 |
| HT | TU | 4.9 | 3340 | 86.8 | 16.3 | 6.0 | 10.3 | 1.8 | 3.1 | 1.87 | 1.00 |
| HT | SP | 5.2 | 3385 | 84.1 | 17.7 | 5.8 | 11.9 | 1.7 | 3.5 | 1.89 | 1.05 |
| HT | SA | 5.4 | 3811 | 136.6 | 20.5 | 9.4 | 11.1 | 2.5 | 2.9 | 2.13 | 0.92 |
| VT | TN | 8.3 | 2593 | 75.3 | 21.5 | 5.2 | 16.2 | 2.0 | 6.3 | 0.76 | 0.77 |
| VT | CI | 4.6 | 5676 | 112.6 | 26.0 | 7.8 | 18.3 | 1.4 | 3.2 | 1.49 | 1.13 |
| VT | TU | 4.8 | 5344 | 130.3 | 25.6 | 9.0 | 16.7 | 1.7 | 3.1 | 1.40 | 0.98 |
| VT | SP | 5.0 | 5417 | 123.6 | 27.0 | 8.5 | 18.5 | 1.6 | 3.4 | 1.42 | 1.03 |
| VT | SA | 5.1 | 6097 | 199.7 | 31.0 | 13.7 | 17.3 | 2.2 | 2.8 | 1.60 | 0.93 |
| FP | TN | 6.0 | 2917 | 62.6 | 17.6 | 4.4 | 13.2 | 1.5 | 4.5 | 1.08 | 0.71 |
| FP | CI | 3.2 | 6386 | 82.0 | 20.2 | 5.7 | 14.5 | 0.9 | 2.3 | 2.13 | 1.38 |
| FP | TU | 3.1 | 6012 | 89.3 | 18.7 | 6.2 | 12.5 | 1.0 | 2.1 | 2.00 | 1.30 |
| FP | SP | 3.4 | 6094 | 86.9 | 20.5 | 6.0 | 14.4 | 1.0 | 2.4 | 2.03 | 1.32 |
| FP | SA | 3.2 | 6859 | 134.4 | 22.1 | 9.2 | 12.9 | 1.3 | 1.9 | 2.29 | 1.36 |
| FP | SP (Fut.) | 0.5 | 14771 | 44.4 | 7.9 | 3.1 | 4.8 | 0.2 | 0.3 | 9.18 | 13.23 |
| RW | CU | 5.2 | 3089 | 53.3 | 16.0 | 3.7 | 12.3 | 1.2 | 4.0 | 0.32 | 1.57 |

| | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| HT | CU | 5.5 | 3861 | 83.9 | 21.3 | 5.8 | 15.5 | 1.5 | 4.0 | 2.16 | 1.18 |
| VT | CU | 5.4 | 6178 | 127.3 | 33.4 | 8.8 | 24.7 | 1.4 | 4.0 | 1.62 | 1.17 |
| FP | CU | 3.8 | 6951 | 90.7 | 26.3 | 6.3 | 20.1 | 0.9 | 2.9 | 2.32 | 1.47 |

**Supplementary Table 13: Summary of projected results for microalgae production in 100 hectares (cultivation and harvesting).** ***RW*****: raceway pond;** ***HT*****: horizontal tubular photobioreactor;** ***VT*****: vertically stacked horizontal tubular photobioreactor;** ***FP*****: flat panels photobioreactor.** ***CU*****: Curaçao;** ***TN*****: The Netherlands;** ***SP*****: Spain;** ***CI*****: Canary Islands;** ***TU*****: Turkey;** ***SA*****: Saudi Arabia**

| **SCENARIO** | | **COST BREAKDOWN FOR MICROALGAE CULTIVATION (as % of cost)** | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| System | Location | Major equipment | Additional capital cost | Raw materials | Consumables | Utilities | Energy | Labor | Wastewater treatment | Others |
| RW | TN | 6.9 | 16.8 | 3.9 | 2.9 | 0.0 | 7.5 | 18.7 | 26.4 | 16.9 |
| RW | CI | 7.5 | 18.0 | 8.5 | 2.9 | 0.0 | 10.6 | 9.9 | 29.5 | 13.2 |
| RW | TU | 8.2 | 19.9 | 9.0 | 3.3 | 0.0 | 9.2 | 5.5 | 33.3 | 11.5 |
| RW | SP | 7.4 | 17.8 | 8.2 | 2.9 | 0.0 | 10.8 | 10.0 | 29.8 | 13.1 |
| RW | SA | 8.7 | 20.9 | 10.6 | 3.4 | 0.0 | 3.0 | 6.6 | 34.2 | 12.7 |
| HT | TN | 4.4 | 19.0 | 15.6 | 12.7 | 0.0 | 8.4 | 18.5 | 4.9 | 16.4 |
| HT | CI | 5.9 | 25.0 | 18.1 | 10.8 | 0.9 | 13.9 | 8.2 | 4.6 | 12.6 |
| HT | TU | 7.1 | 29.8 | 18.3 | 11.2 | 1.2 | 11.9 | 4.2 | 4.8 | 11.5 |
| HT | SP | 6.3 | 26.6 | 17.0 | 10.3 | 1.0 | 13.9 | 7.9 | 4.4 | 12.6 |
| HT | SA | 8.9 | 36.9 | 15.5 | 8.9 | 5.4 | 3.7 | 3.9 | 3.8 | 13.0 |
| VT | TN | 4.7 | 19.7 | 19.3 | 13.6 | 0.0 | 9.5 | 12.4 | 6.9 | 13.9 |
| VT | CI | 5.8 | 24.1 | 21.0 | 11.2 | 0.2 | 14.8 | 5.4 | 6.3 | 11.2 |
| VT | TU | 6.8 | 28.2 | 20.8 | 11.4 | 0.7 | 12.4 | 2.7 | 6.4 | 10.7 |
| VT | SP | 6.1 | 25.4 | 19.8 | 10.8 | 0.5 | 14.9 | 5.2 | 6.1 | 11.3 |
| VT | SA | 8.6 | 35.5 | 18.3 | 9.4 | 4.2 | 3.9 | 2.6 | 5.3 | 12.3 |
| FP | TN | 4.7 | 20.2 | 19.6 | 4.1 | 0.0 | 14.0 | 15.2 | 6.7 | 15.5 |
| FP | CI | 5.4 | 22.7 | 24.2 | 3.6 | 0.9 | 17.5 | 6.9 | 6.4 | 12.4 |
| FP | TU | 6.3 | 26.7 | 25.3 | 3.9 | 1.1 | 14.4 | 3.7 | 7.0 | 11.6 |
| FP | SP | 5.6 | 23.8 | 23.3 | 3.5 | 0.8 | 17.2 | 6.8 | 6.4 | 12.4 |

| | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| FP | SA | 8.1 | 33.7 | 23.0 | 3.3 | 5.0 | 4.2 | 3.6 | 5.9 | 13.3 |
| FP | SP (Fut.) | 7.5 | 32.3 | 24.9 | 5.5 | 0.1 | 10.2 | 6.0 | 0.0 | 13.6 |
| RW | CU | 6.9 | 16.5 | 8.1 | 2.5 | 0.0 | 23.7 | 5.8 | 26.0 | 10.5 |
| HT | CU | 5.2 | 22.0 | 15.0 | 8.5 | 2.0 | 29.5 | 4.4 | 3.6 | 9.6 |
| VT | CU | 5.1 | 21.1 | 17.0 | 8.7 | 1.2 | 30.4 | 2.8 | 4.9 | 8.8 |
| FP | CU | 4.6 | 19.2 | 19.5 | 2.8 | 1.6 | 34.6 | 3.5 | 4.9 | 9.3 |

**Supplementary Table 14: Set of projected results for biorefinery of microalgae. Results are based in 100 hectares of culture in flat panel systems. *B*: Biofuels; *Ch*: Chemicals; *f*: Food-Feed; *F*: Food additives; *Co*: Cosmetics-Healthcare; *C*: Complete biorefinery (exploitation maximizing revenues in different markets). *CU*: Curaçao; *TN*: The Netherlands; *SP*: Spain; *CI*: Canary Islands; *TU*: Turkey; *SA*: Saudi Arabia**

| **SCENARIO** | | **RESULTS FOR BIOREFINERY** | | | | | | |
|---|---|---|---|---|---|---|---|---|
| Market | Location | Biomass cost (€/kg) | Initial investment (M€) | Total cost (M€/Year) | CAPEX (M€/Year) | OPEX (M€/Year) | CAPEX (€/kg) | OPEX (€/kg) |
| B | CU | 1.5 | 12.8 | 10.1 | 1.1 | 9.0 | 0.2 | 1.3 |
| Ch | CU | 1.8 | 19.3 | 12.4 | 1.9 | 10.5 | 0.3 | 1.5 |
| f | CU | 1.4 | 15.7 | 9.9 | 1.4 | 8.5 | 0.2 | 1.2 |
| F | CU | 3.6 | 32.0 | 24.8 | 2.6 | 22.2 | 0.4 | 3.2 |
| Co | CU | 3.6 | 32.8 | 25.2 | 2.6 | 22.6 | 0.4 | 3.3 |
| C | CU | 3.7 | 34.7 | 25.7 | 2.8 | 22.9 | 0.4 | 3.3 |
| B | TN | 1.8 | 11.4 | 5.1 | 1.0 | 4.1 | 0.3 | 1.4 |
| Ch | TN | 1.8 | 14.0 | 5.3 | 1.4 | 3.9 | 0.5 | 1.3 |
| f | TN | 1.4 | 11.3 | 4.2 | 1.0 | 3.2 | 0.3 | 1.1 |
| F | TN | 4.1 | 22.9 | 11.9 | 1.8 | 10.1 | 0.6 | 3.5 |
| Co | TN | 4.2 | 23.4 | 12.1 | 1.8 | 10.3 | 0.6 | 3.5 |
| C | TN | 4.3 | 24.9 | 12.5 | 2.0 | 10.6 | 0.7 | 3.6 |
| B | SP | 0.9 | 12.6 | 5.6 | 1.1 | 4.5 | 0.2 | 0.7 |
| Ch | SP | 1.1 | 17.8 | 6.6 | 1.7 | 4.9 | 0.3 | 0.8 |
| f | SP | 0.9 | 14.5 | 5.2 | 1.3 | 4.0 | 0.2 | 0.7 |

| | | | | | | | | |
|---|---|---|---|---|---|---|---|---|
| F | SP | 2.9 | 30.2 | 17.7 | 2.4 | 15.3 | 0.4 | 2.5 |
| Co | SP | 2.9 | 30.9 | 17.9 | 2.4 | 15.5 | 0.4 | 2.5 |
| C | SP | 3.0 | 32.7 | 18.3 | 2.6 | 15.7 | 0.4 | 2.6 |
| B | CI | 0.9 | 12.6 | 5.7 | 1.1 | 4.6 | 0.2 | 0.7 |
| Ch | CI | 1.1 | 18.0 | 6.8 | 1.8 | 5.0 | 0.3 | 0.8 |
| f | CI | 0.8 | 14.6 | 5.3 | 1.3 | 4.1 | 0.2 | 0.6 |
| F | CI | 2.9 | 30.7 | 18.3 | 2.4 | 15.8 | 0.4 | 2.5 |
| Co | CI | 2.9 | 31.3 | 18.6 | 2.5 | 16.1 | 0.4 | 2.5 |
| C | CI | 3.0 | 33.2 | 18.9 | 2.7 | 16.3 | 0.4 | 2.5 |
| B | TU | 0.7 | 12.5 | 4.3 | 1.1 | 3.2 | 0.2 | 0.5 |
| Ch | TU | 0.9 | 17.7 | 5.4 | 1.7 | 3.6 | 0.3 | 0.6 |
| f | TU | 0.7 | 14.4 | 4.2 | 1.3 | 2.9 | 0.2 | 0.5 |
| F | TU | 2.6 | 30.1 | 15.8 | 2.4 | 13.4 | 0.4 | 2.2 |
| Co | TU | 2.7 | 30.8 | 16.0 | 2.4 | 13.6 | 0.4 | 2.3 |
| C | TU | 2.7 | 32.6 | 16.3 | 2.6 | 13.7 | 0.4 | 2.3 |
| B | SA | 0.4 | 12.8 | 3.0 | 1.1 | 1.9 | 0.2 | 0.3 |
| Ch | SA | 0.6 | 18.3 | 3.9 | 1.8 | 2.1 | 0.3 | 0.3 |
| f | SA | 0.4 | 14.9 | 3.0 | 1.3 | 1.7 | 0.2 | 0.2 |
| F | SA | 2.3 | 31.9 | 15.7 | 2.5 | 13.1 | 0.4 | 1.9 |
| Co | SA | 2.3 | 32.6 | 15.9 | 2.6 | 13.4 | 0.4 | 1.9 |
| C | SA | 2.4 | 34.3 | 16.3 | 2.7 | 13.6 | 0.4 | 2.0 |

**Supplementary Table 15: Summary of projected results for biorefinery of microalgae. Results are based in 100 hectares of culture in flat panel systems. *B*: Biofuels; *Ch*: Chemicals; *f*: Food-Feed; *F*: Food additives; *Co*: Cosmetics-Healthcare; *C*: Complete biorefinery (exploitation maximizing revenues in different markets). *CU*: Curaçao; *TN*: The Netherlands; *SP*: Spain; *CI*: Canary Islands; *TU*: Turkey; *SA*: Saudi Arabia**

| **SCENARIO** | | **COST BREAKDOWN FOR BIOREFINERY (as % of cost)** | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| Market | Location | Major equipment | Additional capital cost | Energy | Labor | Raw materials | Utilities | Wastewater treatment | Consumables | Others |
| B | CU | 1.8 | 9.2 | 32.2 | 6.2 | 1.4 | 35.8 | 0.2 | 3.5 | 9.7 |
| Ch | CU | 2.1 | 10.4 | 32.5 | 5.0 | 1.1 | 36.0 | 0.2 | 3.5 | 9.3 |
| f | CU | 2.3 | 11.6 | 32.8 | 3.8 | 0.8 | 36.1 | 0.2 | 3.6 | 8.9 |
| F | CU | 1.9 | 8.5 | 20.2 | 3.9 | 12.9 | 15.6 | 0.0 | 31.1 | 6.1 |
| Co | CU | 1.9 | 8.6 | 20.0 | 4.0 | 12.5 | 15.8 | 0.0 | 31.0 | 6.2 |
| C | CU | 2.0 | 8.8 | 19.8 | 4.1 | 12.0 | 16.1 | 0.0 | 30.9 | 6.2 |
| B | TN | 3.2 | 16.2 | 18.2 | 30.8 | 1.2 | 9.8 | 0.1 | 2.9 | 17.7 |
| Ch | TN | 2.7 | 13.9 | 25.5 | 17.3 | 1.0 | 23.0 | 0.2 | 3.2 | 13.3 |
| f | TN | 2.3 | 11.6 | 32.8 | 3.8 | 0.8 | 36.1 | 0.2 | 3.6 | 8.9 |
| F | TN | 2.8 | 12.5 | 9.7 | 20.0 | 11.3 | 4.5 | 0.0 | 27.1 | 12.2 |
| Co | TN | 2.8 | 12.7 | 9.5 | 20.4 | 10.8 | 4.5 | 0.0 | 26.8 | 12.4 |
| C | TN | 2.9 | 12.9 | 9.4 | 20.8 | 10.3 | 4.6 | 0.0 | 26.6 | 12.6 |
| B | SP | 3.2 | 16.3 | 22.6 | 14.7 | 2.3 | 22.6 | 0.3 | 5.5 | 12.4 |
| Ch | SP | 3.6 | 18.3 | 23.4 | 12.1 | 1.8 | 23.3 | 0.3 | 5.7 | 11.5 |
| f | SP | 4.0 | 20.2 | 24.2 | 9.5 | 1.4 | 24.0 | 0.3 | 5.9 | 10.6 |
| F | SP | 2.5 | 11.2 | 10.6 | 7.1 | 15.9 | 7.7 | 0.0 | 38.2 | 6.8 |
| Co | SP | 2.5 | 11.4 | 10.5 | 7.3 | 15.3 | 7.8 | 0.0 | 38.1 | 6.9 |
| C | SP | 2.6 | 11.7 | 10.4 | 7.5 | 14.8 | 8.0 | 0.1 | 38.0 | 7.0 |
| B | CI | 3.2 | 16.2 | 22.4 | 14.5 | 2.3 | 23.2 | 0.3 | 5.7 | 12.3 |
| Ch | CI | 3.6 | 18.1 | 23.1 | 11.9 | 1.9 | 23.9 | 0.3 | 5.9 | 11.5 |
| f | CI | 3.9 | 20.0 | 23.9 | 9.3 | 1.4 | 24.5 | 0.3 | 6.1 | 10.6 |
| F | CI | 2.7 | 12.1 | 8.8 | 3.8 | 17.8 | 6.5 | 0.0 | 42.8 | 5.4 |
| Co | CI | 2.6 | 11.8 | 9.6 | 5.5 | 16.4 | 7.3 | 0.0 | 40.7 | 6.1 |

| | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| C | CI | 2.5 | 11.5 | 10.3 | 7.2 | 15.0 | 8.0 | 0.1 | 38.5 | 6.9 |
| B | TU | 4.2 | 21.1 | 22.3 | 9.6 | 2.9 | 22.0 | 0.3 | 7.0 | 10.5 |
| Ch | TU | 4.5 | 23.1 | 22.6 | 7.8 | 2.3 | 22.2 | 0.4 | 7.1 | 9.9 |
| f | TU | 4.9 | 25.0 | 23.0 | 6.0 | 1.7 | 22.5 | 0.4 | 7.2 | 9.3 |
| F | TU | 2.8 | 12.5 | 9.0 | 4.0 | 17.6 | 6.5 | 0.0 | 42.2 | 5.5 |
| Co | TU | 2.8 | 12.8 | 9.0 | 4.1 | 17.0 | 6.6 | 0.0 | 42.2 | 5.6 |
| C | TU | 2.9 | 13.1 | 8.9 | 4.3 | 16.4 | 6.7 | 0.1 | 42.1 | 5.6 |
| B | SA | 6.1 | 30.7 | 10.1 | 13.5 | 4.7 | 11.1 | 0.6 | 11.5 | 11.8 |
| Ch | SA | 6.6 | 33.6 | 10.2 | 11.5 | 3.7 | 11.1 | 0.6 | 11.5 | 11.2 |
| f | SA | 7.2 | 36.4 | 10.3 | 9.5 | 2.7 | 11.2 | 0.6 | 11.6 | 10.6 |
| F | SA | 2.9 | 13.3 | 3.0 | 4.6 | 20.2 | 2.3 | 0.0 | 48.4 | 5.3 |
| Co | SA | 3.0 | 13.5 | 3.0 | 4.7 | 19.8 | 2.3 | 0.0 | 48.3 | 5.4 |
| C | SA | 3.0 | 13.8 | 2.9 | 4.8 | 19.4 | 2.3 | 0.1 | 48.1 | 5.4 |

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