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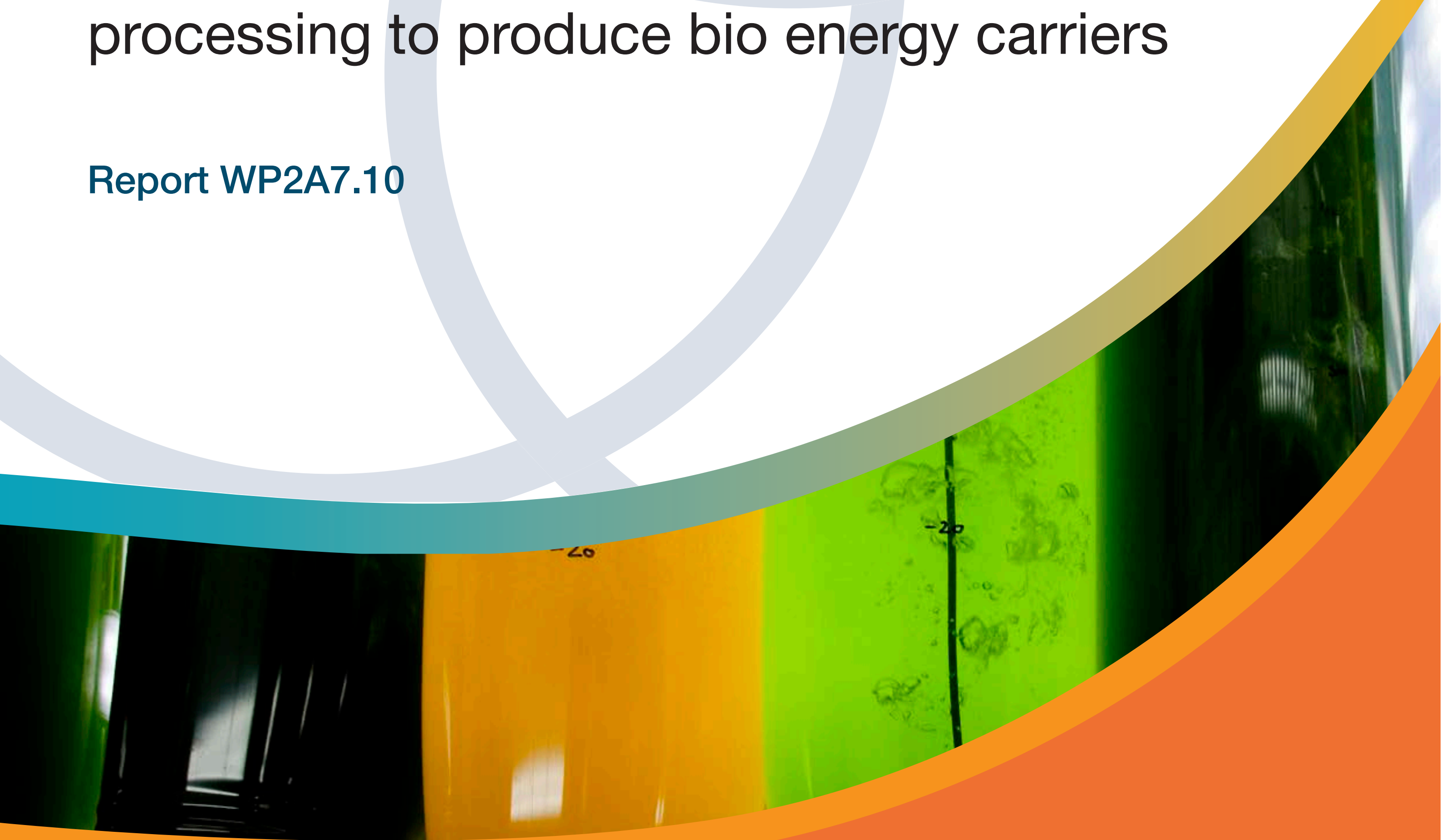
# **AlgaEconomics:** bio-economic production models of micro-algae and downstream processing to produce bio energy carriers

## Report WP2A7.10



**Energetic Algae ('EnAlgae')**

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# Report WP2A7.10 – AlgaEconomics: bio-economic production models of micro-algae and downstream processing to produce bio energy carriers

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# AlgaEconomics: bio-economic production models of micro-algae and downstream processing to produce bio energy carriers

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## Introduction

There is a worldwide interest in - macro- and micro-algae based on the promise they hold to provide the market with raw material to produce fuel, feed, food, chemicals, and materials. This has led to a large number of innovative production systems with the aim to increase production and match the products to markets. However, commercial production is growing slowly mainly because the production costs of algae are considered to be high. There is consensus that a learning curve is needed to decrease production costs to be able to profit from the promise that algae hold.

This report describes results of work carried out within the EnAlgae project to describe production costs and identify the variables that have most effect in determining future cost prices so that R&D can be focussed on these issues. This has been done by making use of pilots within the EnAlgae consortium and by describing the process in Excel models that have been spread among and discussed with stakeholders active in the field of commercial algae production. The expectation is that this transparency and interaction will lead to an increase of the learning curve to make algae production cheaper and thus supplying more markets.

Regarding micro-algae production, we have focussed on an open pond production system, a tubular, and a flat panel system. These systems are described in bio-economic models that combine the estimation of algae production of dry matter including water and energy use with an economic assessment that provides a detailed cost price analysis. These models are described in chapter 1. This chapter also includes comparisons between the different cultivation systems. Further refinery steps to produce the energy carriers biodiesel, ethanol, and methane are also described in models and are presented in chapter 2.

The development of these models does not stop with this report but will be an on-going process as more and more knowledge will become available on important parameters and characteristics. Readers are encouraged to give feedback on this report and provide us with better insights. As new versions of the models will be produced, they will be available to anybody who feels they might contribute to improvements.

# AlgaEconomics: bio-economic production models of micro-algae and downstream processing to produce bio energy carriers

## 1 Micro Algae Production

Economic aspects concerning micro-algae production systems were studied on the basis of an economic model in Excel named ‘AlgaEconomics’. The model was constructed with the contribution of EnAlgae project partners and with help from experts from private companies and public research institutes in the Netherlands, Belgium, Germany and the UK. The economic model for algae production contains an algae production model that estimates the amount of algae produced as well as a variety of parameters that are important for the economics. By constructing an algae production model in which the growth parameters are freely adjustable, the economics become applicable to any specific situation. The ‘AlgaEconomics’ model is freely available to all those interested in micro-algae production and use.

The general part of the model, being independent of the production system chosen, is discussed in section 1.1. Three micro-algae production systems are distinguished: open pond, tubular photobioreactor and flat panel photobioreactor. The system specific aspects are discussed in the sections 2.2, 2.3 and 2.4. For each system a sensitivity analysis of several important parameters is included. In section 1.5 the production and economic results are compared.

### 1.1 ‘AlgaEconomics’ an economic model of micro-algae production

#### 1.1.1 Construction of the bio-economic models

The Excel model contains several worksheets. Algae production calculations (including harvest) are carried out in a worksheet called "algae production". This worksheet provides a detailed insight in algae production results. Parameters used in the calculations are listed in a worksheet called “parameters”. The parameters can be adjusted freely and the impact of the changes is instantly seen in the “algae production” worksheet. Similar to the algae production, economic calculations are made in the "economics" worksheets with data and parameters obtained from the “parameters” worksheet and the “capital goods” worksheet listing all the investment costs required including percentages for depreciation, maintenance and interest. The capital goods costs are based on production systems with a Photo Bio Reactor (PBR) unit size corresponding to a 1,000 $m^2$ surface area of land. The “parameters” worksheet allows the user to change the number of units.

The models assume that production in all PBRs (open pond, flat panel or tubes) is sunlight driven and that the PBRs are exposed to ambient temperatures and precipitation. The “climatic data” worksheet lists monthly climatic data from different locations. For the models, the following weather data are imperative: average minimum, maximum and average temperature, average wind speed, average Relative Humidity and cumulative global radiation. In addition, the heat use calculation of the PBRs requires an indication of the saturation vapour pressure (provided in the worksheet “sat vap pressure”) and average monthly day length for which a location based calculator is provided (worksheet “daylength calc”).

The calculations also assume that minerals area available at optimum levels and that supply always corresponds to the algae demand.

Output of the algae production calculations can be found in the "process data" and "input-output" worksheets.

The figure below illustrates the algae production model structure. The model calculates algae biomass production from the available global radiation using a photosynthetic efficiency that is also provided in the “parameters” worksheet. Based on this parameter, the global radiation is transferred into an amount of glucose produced using the energy content of glucose. Next, the sugar is converted into algae biomass taking into account the biomass composition (which can be adjusted in the “parameters” worksheet). This conversion procedure builds on the knowledge that different components like carbohydrates, lipids and protein have distinctive demand for glucose input (Vertregt & Penning de Vries, 1987).

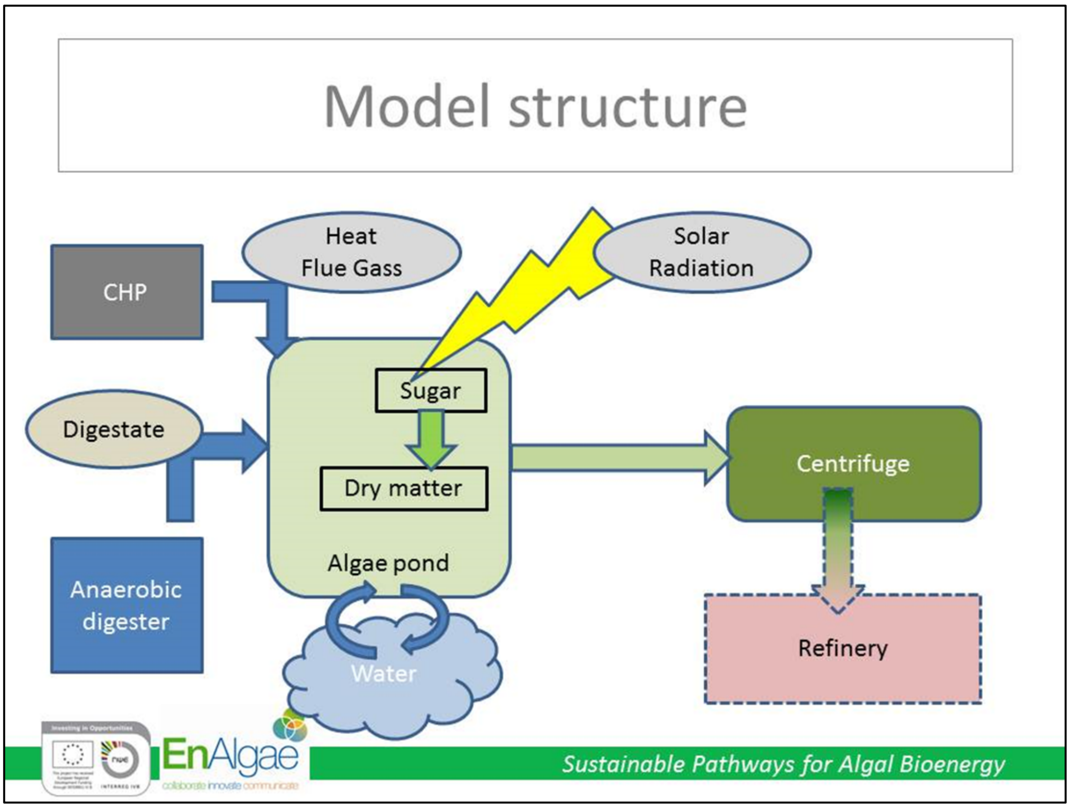


***Figure 1:*** *Model structure open pond micro-algae.*

The amount of algae biomass produced is then harvested by centrifugation. Input from the digester and the combined heat and power plant such as flue gas, heat and digestate (minerals) are optional parameters in the model that influence the algae biomass production economics. The harvest calculation procedure assumes that all the biomass potentially produced in a month will be harvested. The volume of water that subsequently needs to be processed in that particular month is calculated based on the biomass production and the biomass density in the PBR (for which a parameter value is available in the “parameters” worksheet).

Apart from the biomass harvested, the model also calculates the energy consumption, the use of water, the need for fertilisers (N and P) and the amount of labour which are used for the economics of the algae production process. The energy use assumes a 24/7 operation of the PBR and the associated cost of pumping, mixing, heating, air sparging, harvesting and climate control. The capacity of this equipment is adjustable in the “parameters” worksheet. The water use is based on evaporation (open pond) or cooling (flat panel or tubes) and waste water resulting from the harvest procedure assuming that a certain percentage of water returns to the pond (a number that can be adjusted in the “parameters” worksheet).

### 1.1.2 Detailed description of the algae biomass production calculations

In this paragraph the calculation procedure designed to estimate the production of algae is described. The glucose production (GP in kg per month) is based on the monthly global radiation (RAD in J. $cm^{-2}$. $month^{-1}$), the energy content of glucose ($E_{gluc}$ in $kJ.g^{-1}$), and a conversion factor for the photosynthesis efficiency ($CF_{LG}$), representing the percentage of global radiation absorbed by the algae cells and available for glucose production. The calculation accounts for growth inhibiting factors like sub-optimal temperatures ($T_f$), the availability or absence of $CO_2$ from flue gas ($CO_{2f}$), and the use of digestate instead of mineral fertilisers ($Dig_f$).

$$GP = RAD * PBRsurf * \frac{1}{100{,}000} * \frac{1{,}000}{Egluc} * CF_{LG} * T_f * CO_{2f} * Dig_f \quad [1],$$

where $PBR_{surf}$ equals the surface area occupied by the PBRs, 100,000 is a factor to convert J to GK and $cm^2$ to $m^2$. The factor 1,000 is used to convert $E_{gluc}$ to $GJ.kg^{-1}$.

The factors for use of flue gas and digestate are adjustable parameters. It is assumed that without $CO_2$ production is reduced because $CO_2$ becomes limiting for growth. When using digestate there is a negative influence on growth because of light reduction. The temperature factor is calculated as an exponential limitation caused by suboptimal temperature and is based on an algal growth model by James and Boriah (2010).

$$T_f = e^{-K(T-T_{opt})^2} \quad [2]$$

In this equation the difference between the actual temperature (T) and a species specific optimal temperature ($T_{opt}$) is used to calculate a production limitation. The empirical constant (K) and the optimal temperature can be adjusted in the worksheet “temperature-growth”. Equation [2] is shown in Figure 2.

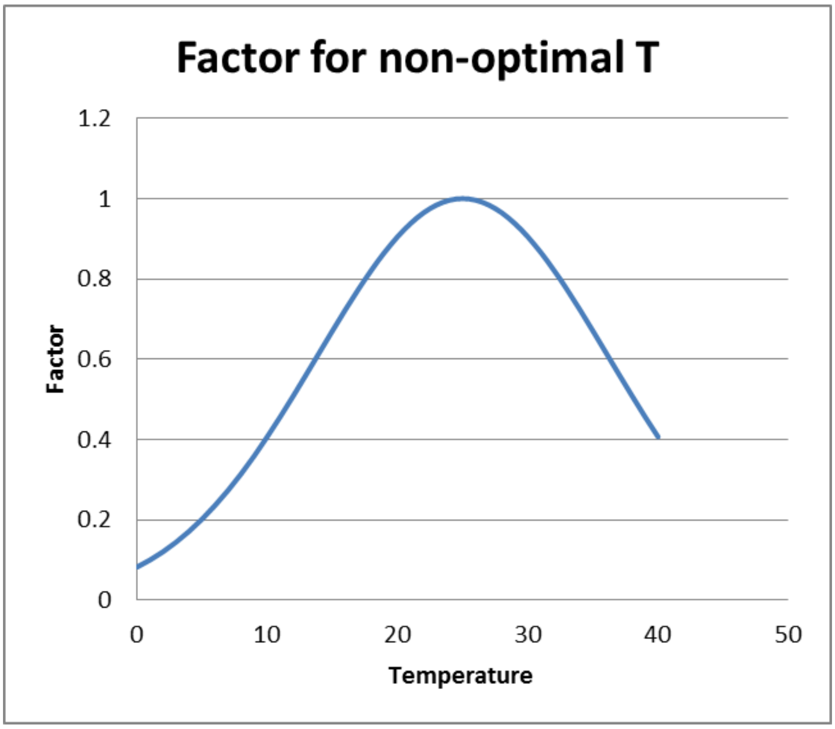


***Figure 2:** The influence of temperature on glucose production (based on James & Boriah, 2010).*

Next step in the calculation of algae production is the conversion of glucose into dry matter or potential biomass production ($Mx_{pot}$ in kg). The associated calculation is based on a publication by Vertregt & Penning de Vries (1987) who established a relation between dry matter production and the composition of that dry matter in a large variety of terrestrial plants. We assume that algae could be following the same relationship. The conversion parameters describing the need for glucose in kg per kg of cellular component ($GF_x$) are given in the parameters section.

$$Mx_{pot} = \frac{GP}{(GF_{carbohydrates*}*\%_{Carbohydrates})+(GF_{lipids}*\%_{lipids})+(GF_{proteins}*\%_{proteins})+(GF_{ash}*\%_{ash})} \quad [3]$$

In this equation GP is taken from equation [1].

The dry matter composition can vary based on the species present. In open pond, this could well be a mixture of species. Therefore, the models provide the opportunity to include three different species and the relative share of each of these species in the population. Dry matter composition of a series of micro-algae species can be found in Becker (2007).

In algae cultivation, culture crashes can occasionally occur and PBRs need regular cleaning. The model accounts for such events and provides associated parameters to be set in the “parameters” worksheet such as the number of crashes/cleaning, the amount of labour involved, and the resulting production downtime. In the model it is assumed that after a crash or a cleaning procedure, the biomass and the culture medium in the pond are waste. Because crashes are unpredictable the number of crashes is converted into production and culturing downtime factors which can be used to divide the effect of crashes over the entire production period. The production downtime factor ($CPF_{crash}$) is used to convert biomass production potential ($Mx_{pot}$ from equation [3]) into biomass production (kg) realised ($Mx_{rel}$ ; equation [4]).

$$CPFcrash = \frac{365-Nc*DTc}{365} \quad [4],$$

in which Nc equals the number of crashes and/or cleaning events and DTc equals the downtime in days per crash and/or event. Now Mxrel can be calculated as follows:

$$Mx_{rel} = Mx_{pot} * CPF_{crash} \quad [5]$$

The maximum algae density (Cx in $kg/m^3$) is a parameter in the “parameters” worksheet which depends on the type of PBR. It should be noted that in winter periods in temperate regions, the production can be very low resulting in a maximum density lower than the parameter value associated with the PBR type. The volume of algae culture to be processed by the centrifuges ($V_{prod}$ in $m^3$ per month) can now be calculated as follows:

$$V_{prod} = Mx_{rel}/C_x \quad [6]$$

By dividing $V_{prod}$ by the number of days per month (DAYS), the volume to be processed per day can be estimated ($V_{day}$ in $m^3$ per day).

$$Vday = \frac{Vprod}{DAYS} \quad [7]$$

From the difference between the biomass production potential and the realised biomass production the loss of biomass and amount of waste water due to crashes evidently follows.

Temperature and water management of the different PBRs varies fundamentally between open ponds and closed PBRs. Cooling is of importance with closed PBRs while evaporation is relevant only to open pond systems. For all three PBRs the temperature that is required in the culture can be adjusted in the "parameters" worksheet of the model. Additionally, for the closed PBRs a second temperature set point is provided in the model to account for necessary cooling of the PBR as soon as the radiation energy would be larger than the potential heat loss in the trajectory between the maximum allowed culture temperature and the outside air temperature. Equations are provided in the paragraphs in which the PBRs are described.

### 1.1.3 Detailed description of harvesting calculations

In equation [6] the monthly volume of algae culture to be processed by the centrifuges is calculated. The centrifuge will produce a pellet and a supernatant. In our models, the separation process is governed by the following parameters:

- The biomass recovery efficiency (BIOrec) representing the amount of algae in the supernatant relative to the amount present in the inflow.
- The dry matter percentage of the pellet (Pdm in kg per $m^3$).
- The actual algae concentration (kg per $m^3$) in the inflow ($C_x$)

The amount of pellet or algae paste (Vharvest in $m^3$ per month) can now be calculated as:

$$Vharvest = \frac{Vprod*BIOrec*Cx}{Pdm} \qquad [8]$$

$C_x$ has a typical value of 150 kg of dry matter per $m^3$, a value experimentally established from the algae pilot in Lelystad.

Vprod is derived from equation [6]. Vday (equation [7]) is used to determine the required centrifuge capacity based on the month with the highest daily inflow to the centrifuge and a 10 % extra capacity to deal with variations that will inevitably occur. This centrifuge capacity is input for the capital goods and thus investment calculation. In addition Vday is used to calculate the energy consumption and amount of labour for harvesting.

In the model, the supernatant is a return flow to the PBR to save on water use. However, it might be advisable to return only part of the supernatant and thus allow for fresh water intake in the PBR. In the model a parameter is introduced to govern the percentage of supernatant returning to the PBR. The higher the percentage, the lower the total water use of the algae production and vice versa.

Flue gas consumption is estimated from the amount of algae biomass produced based on an estimated carbon content of dry matter (0.54 g/g). Accounting for an estimated $CO_2$ efficiency (expressed as a percentage of $CO_2$ supply) and the $CO_2$ content of the flue gas, the amount of flue gas consumed is calculated.

The model spreadsheets allow for water use to be calculated. This is based on the following:

- Water loss by crashes or cleaning.
- Waste water from the centrifuge (water not returning to the PBR).
- Water output in the final product (algae paste containing 15% dry matter)
- For an open pond PBR the water evaporation and rainfall are accounted for while in a closed PBR cooling requirement is met by sprinkling water on the PBR. This is explained in the sections specific to these PBRs.

Labour requirement is based on daily operations and on extra labour need in case of cleaning or PBR crash. The model distinguishes between low quality (LQ) and high quality (HQ) labour.

It is assumed that for cleaning and crash related restarts, only LQ labour is required and that for daily operations a mixture of LQ and HQ labour is needed. The amount of labour required is made dependent on the amount of PBRs installed and decreases per unit of PBR when the number of PBRs increases (efficiency gain). For daily operations the procedure for LQ labour calculations is explained as follows. First, it is assumed that cleaning a PBR takes around 16 hours LQ labour per event. This is based on our own observations. Next, the amount of LQ labour (h per crash) is calculated with [9].

$$LQcrash = PBRn * 16 * SCf,n \qquad [9]$$

in which $LQ_{crash}$ equals the LQ labour need, $PBR_n$ equals the number of PBRs and $SC_{f,n}$ represents a scaling factor depending on the number of PBR units installed. This factor is based on the capital investment cost development with increasing number of PBR (assuming that the cost efficiency will rise with increasing capacity). $SC_f$ is calculated as follows:

$$SCf,n = \frac{Cn}{PBRn*CIC1} \qquad [10]$$

where CIC1 equals the capital cost of installing 1 PBR unit and $CIC_n$ equals the capital investment cost of n units. $CIC_n$ is calculated based on Sinnott et al (2005):

$$CICn = CIC1 * PBRn^{p} \qquad [11]$$

where p is a parameter that is set at 0.6 in our calculations. This is debatable as there is no indication of the value of p in constructing algae PBRs.

The amount of nitrogen and phosphorus needed for algae production is calculated from the amount of produced biomass and the elemental composition of the biomass and the loss of nutrients with wastewater. For N 0.16 g are assumed per g of protein (typical value used to convert nitrogen content to crude protein content) while 0.0027 g P per g of dry matter is taken from the ECO Phyllis2 database using the entry of "microalgae" (https://www.ecn.nl/phyllis2/Browse/Standard/ECN-Phyllis#algae).

When digestate use is selected the amount of digestate needed is calculate from the least available component, neglecting the possible effect of over addition of other components.

### 1.1.4 Detailed description of economics

The "Economics" worksheet contains the results of the economic calculations, based on the principles and price assumptions from the "parameter" worksheet in the excel models:

**return**
The total return (TR) is the yearly amount (YA) of algae biomass multiplied by the selling price (SP) of dry matter

TR = YA * SP [12]

**costs**
The total cost (TC) is the sum of the variable costs (VC), capital good costs (CC) and costs for land use (LC).

Variable costs (VC) are yearly positive or negative costs for water, electricity, $CO_2$, labour, digestate, fertilisers and wastewater. In the parameter section the assumed prices and their sources for in- and output are set. These prices can be changed, when working with the model.

***Table 1:*** *Price assumptions per unit for in- and output.*

| In- or output | Price | Unit | Source |
|---|---|---|---|
| **Algae biomass** | 35 | €/kg | feed additive price realised in 2013 |
| **Water** | 0.878 | €/$m^3$ | Dutch price level 2014 |
| **Electricity** | 0.107 | €/kWh | Dutch price level 2014 |
| **$CO_2$ credits** | 0 | €/tonne | |
| **Heat recovery** | 0 | €/kWh | |
| **Labour LQ** | 11.53 | €/hr | KWIN (2012)[1] |
| **Labour HQ** | 25.57 | €/hr | KWIN (2012) |
| **Digestate use** | -20 | €/$m^3$ | average price of animal manure in the Netherlands in 2014 |
| **Fertiliser (N)** | 1.08 | €/kg N | KWIN (2012) |
| **Fertiliser (P)** | 2.22 | €/kg P | KWIN (2012) |
| **Wastewater** | 0.1 | €/$m^3$ | Dutch levy for waste water discharge |
| **Land** | 1,041 | €/ha | KWIN (2012) |

[1]KWIN is the Dutch database for economic parameters in arable and field vegetable crop production

The purchase prices for capital goods are listed in the worksheet "capital goods". Capital goods costs (CC) are yearly costs for depreciation, maintenance, interest and insurance.

Depreciation is dependent on the estimated lifetime for each item and based on a 10 % resale value. The percentage maintenance costs is also estimated per item. The interest for capital goods is 5.5 % which is 0.55*5.5 % =3 % per year at a resale value of 10 %. Insurance costs are assumed to be 0.5 %. The 0.55 factor is the average of the purchase price and the residual value after the life span of the installation divided by the purchase price.

Land use costs (LC) are yearly costs for land use (LU) based on a lease construction. In the calculations for the land use is assumed that 1,200 $m^2$ land is needed for a PBR production unit of 1,000 $m^2$, namely 1,000 $m^2$ for the unit itself and 200 $m^2$ extra land for the associated equipment and infrastructure. This extra land claim is likely to decrease per unit PBR when scaling up. For this reason, the models make LC dependent on the number of PBR units :

$$LU = PBRn * 200 * SCf,n \qquad [13]$$

Total cost (TC) can now be calculated as follows:

$$TC = VC + CC + LC \qquad [14]$$

**results**

The total financial result (RES) is the total return (TR) minus the total costs (TC).

$$RES = TR – TC \qquad [15]$$

**cost price**

The cost price (CP) is total costs (TC) divided by the annual amount of dry algae biomass produced (YA). YA is based on the monthly $V_{harvest}$ and the dry matter percentage of the centrifuge pellet (equation [8]).

$$CP = TC / YA \qquad [16]$$

**Return On Investment**

Total investment capital goods (CIC) is the total of the purchase prices for the capital goods.

Return On Investment (ROI) is the total return (TR) minus variable costs (VC), minus maintenance and interest costs for capital goods (CC) and minus land costs (LC) divided by the total investment capital goods (IC), (depreciation and interest costs for capital goods are excluded from this calculation).

$$ROI = (TR – VC - CC^{1} – LC) / IC \qquad [17]$$

**pay-back time**

The pay-back time (PB) in years is the total investment capital goods (IC) divided by the total return (TR) minus variable costs (VC), minus maintenance and interest for capital goods (CC) and minus land costs (LC), (depreciation and interest costs for capital goods are excluded from this calculation).

$$PB = IC / (TR – VC – CC^{1} – LC) \qquad [18]$$

[1] Depreciation and interest costs for capital goods are excluded from this calculation

### 1.1.5 Changing scenarios for algae production

The model provides several easy accessible options that influence the production level and the financial return realised. The most important of these options are:

- Scaling up the production capacity
    - Scaling up is an important facility of the model. It can be assumed that certain costs will decrease when the production site would be scaled up to increase the production volume. There is not much experience and data available to indicate to what extent the relative costs would go down in scaling up. Therefore, we have used a calculation procedure that originates from the chemical industry as published by Sinnott et al. (2005) and which was already used in equation [11]. The scaling parameter n is set at 0.6 but this can be adapted by the user into other values that are more likely applicable. This up scaling equation is used for the calculation of capital good costs, labour requirement and extra land requirement for the associated equipment and infrastructure. The basis for scale level is the number of PBRs. We have assumed that PBRs need a surface of 1,000 $m^2$ and scaling up will simply require more units to be installed. The model provides scaling up until 1,000 units and therefore 100 ha of algae production.

- Varying the location of the production facility
    - The standard location for the model is Lelystad, the Netherlands, but in the parameters section a list of options is provided. Changing to a different location changes the climate data set thereby changes the algae production based on global radiation, temperature and rainfall. The location also changes the day length used for the calculation of heat energy consumption (see sections 1.2-1.4). Note however that not all listed climate data sets are currently complete which could result in incorrect calculations for these locations.

- Seasonal operation
    - In the parameters section an option is given to stop the operation of the algae facility for part of the year. This was introduced bearing in mind cold winter periods when algae production might become limited while variable costs still occur. When the installation is not in operation it is assumed that there are zero operational costs, obviously there is also no production in that period.

- Use of flue gas for $CO_2$ addition
    - Flue gas addition is considered optional for the open pond model. It is assumed that without artificial $CO_2$ addition the algae growth is reduced or limited. In this simple model the reduction of growth is estimated by a simple factor on the conversion of light energy to biomass. When the option of flue gas addition is switched off, the capital costs for sparging are neglected in the economic calculation. For obvious reasons, without $CO_2$ there will be no production possible in closed PBRs as gas exchange facilities are essential in these constructions.

- Use of digestate as a nutrient source
    - Within the model it is possible to use digestate as a source of nutrients for the growth of algae. If the option is not used, the model calculates the need for artificial fertiliser as separate nitrogen and phosphorus. If digestate is used, a negative growth factor is introduced as we assume that digestate turbidity and non-ideal composition will reduce the growth. The use of digestate will reduce the cost of fertilisers and thus reduce VC. This cost reduction can vary depending on the market for animal manure and can in some countries be a negative price.

- Application of heat
    - In the model it is possible to select the use of heat to support production levels. If the option is not selected the temperature of the cultures is assumed identical to the average environmental temperature. When heating is selected, it is assumed that the algae culture in the PBR is heated to a desired temperature level that is an input parameter located in the worksheet “parameters”. This temperature target level can be chosen to vary per month. When heat use is selected, the heat consumption is calculated assuming that the PBR cools down at night to minimum ambient temperature level and is heated to the desired temperature level during daytime. When there is no heat use, the capital cost for heating equipment is ignored in the economic calculations.

- Varying the algae species
    - In the “parameters” worksheet there is an option to select a maximum of three different algae species and their relative occurrence. By changing the algae species and their relative occurrence within the culture, the composition of the algae biomass harvested changes accordingly. Also, the production level is influenced depending on equation [3].

- Varying the prices of inputs
    - In the parameter section the assumed prices and their sources for in- and output are set. These prices can be changed to analyse the effects on the economics.

### 1.1.6 Extensive parameter adaptation

The parameter values listed in section 1.1.5 are the most important and influential parameters, but the model offers a large variety of parameters to be adapted for specific circumstances. The parameters can all be found in the "parameters" worksheet and can be changed freely. The use of excel to build the model makes the model transparent so that anybody can assess the potential influence of a parameter of his or her choice.

## 1.2 Open pond AlgaEconomics

### 1.2.1 The open pond model

The open pond system described in the model is based on a pilot built in Lelystad, the Netherlands, at ACRRES, the Application Centre for Renewable Resources (ACRRES is a Wageningen-UR initiative) (Figure 3).

***Figure 3:*** *Midwinter 2013 at the outdoor open pond of the ACRRES micro-algal pilot in Lelystad (Netherlands).*

The open pond business economic model is based largely on the data and experience collected by running this pilot. In paragraph 1.1 the model is described. Yet, an open pond PBR deviates substantially from closed PBRs especially where direct interactions between the atmosphere and the algae culture are concerned. This relates to heat and water transfer. In this paragraph the heat and water relations in the open pond model are explained.

Evaporation of water from the pond and rainfall into the pond determine the amount of water needed to support a constant water table. The amount of rainfall is taken from climatic data. Evaporation on the other hand is calculated based on equations for evaporation from an inactive outdoor swimming pool (Smith et al, 1994). The evaporation rate ($M_{evap}$ in kg/m$^2$/h) is calculated (equation [18]) from wind velocity ($v_{wind}$), saturation vapour pressure at water temperature ($P_w$), and the saturation vapour pressure at air dew point ($P_a$) and the latent heat of water at the pond temperature ($DH_v$), multiplied by the surface area of the ponds ($A_{pond}$). Because the algae ponds are mixed and sparged, which surely influences the evaporation rate, the calculated data should be considered indicative.

$$M_{evap} = A_{pond} * \frac{(30.6+32.1*v_{wind})*(P_w-P_a)}{DH_v} \qquad [19]$$

in which 30.6 and 32.1 are constants as presented by Smith *et al*. (1994). Pa (kPa) is calculated as:

$$Pa = 0.6108 * e^{\frac{17.27*Tair}{237.3+Tair}} \quad [20]$$

in which $T_{air}$ is the air dew point temperature. The same equation can be used to calculate Pw while substituting $T_{air}$ by $T_w$ representing the average water temperature. The latent heat of water at pond temperature ($DH_v$ in kJ per kg) is calculated as follows:

$$DHv = 2500.8 - 2.36 * Tw + 0.0016 * Tw^2 - 0.00006 * Tw^3 \quad [21]$$

which equation can be found in Rogers & Yau (1989). The total amount of monthly evaporation ($M_{evap,\ total}$) is calculated by multiplying the evaporation rate ($M_{evap}$) by the number of daylight hours per month and the area ($m^2$) involved.

The model calculates the amount of heat required to keep the pond's temperature at the target temperature. Heating is only applied during daytime so the pond cools down during night. It is assumed that the pond's temperature at dawn equals the ambient air temperature. In the calculation of heating required, the heat sources involved are: 1) heating of the pond's water body at dawn, 2) evaporation heat, 3) long wave radiation heat loss to the atmosphere , 4) heat input from radiation, 5) air radiation loss. Neglected are convection and conduction, as these sources are minor relative to the others (see: Béchet et al, 2011).

1) Heating of the pond's water body at dawn. This assumes that the culture cools down during the night to the ambient air temperature ($T_{min}$) and is heated up in the morning to the targeted temperature ($T_{set}$). The heat needed to do so ($H_{dawn}$) is calculated by multiplying the temperature difference with the specific heat capacity of water ($C_w$), the volume of water ($V_w$), and the number of days per month (Days).

$$H_{dawn} = (T_{set} - T_{min}) * C_w * V_w * Days \quad [22]$$

2) Evaporation heat ($H_{evap}$) is calculated from the amount of evaporation ($M_{evap,\ total}$) by multiplying with the latent heat of evaporation ($DH_v$).

$$H_{evap} = M_{evap,total} * DH_v \quad [23]$$

3) Long wave radiation heat loss or pond radiation ($H_{prad}$) is calculated according stefan-boltzmann's fourth power law using emissivity and water ($e_w$), boltzmanns constant **$\sigma$** the temperature of air ($T_{air}$), the ponds surface area ($A_{pond}$), the number of days (Days) and the average length of the day ($Day_{time}$). The latter is used because the air radiation is only accounted during daytime, at night the ponds are allowed to cool down to the minimum temperature which is compensated for by the heat use in the morning.

$$H_{Prad} = -\varepsilon_w * \sigma * T_{air} * A_{pond} * Days * Day_{Time} \quad [24]$$

4) Heat input from solar radiation ($H_{sol}$ in kWh) is taken from the climatic data as global radiation ($G_{rad}$), multiplied by the surface area ($A_{pond}$) and adjusted to the correct units.

$$H_{sol} = G_{rad} * A_{pond} \quad [25]$$

5) air radiation loss or air radiation ($H_{Arad}$ in kWh) is calculated according stefan-boltzmann's fourth power law using emissivity of air and water (respectively $e_a$ and $e_w$ ), boltzmanns constant s the temperature of air ($T_{air}$), the ponds surface area ($A_{pond}$), the number of days (Days) and the average length of the day ($Day_{time}$).

$$H_{Arad} = \varepsilon_a * \varepsilon_w * \sigma * T_{air} * A_{pond} * Days * Day_{Time} \quad [26]$$

### 1.2.2 Description of open pond parameters

In this paragraph a short description of the most important parameters is given.

The open pond economic model comes with a basic set of parameters that are based on a crossover between the open pond algae production pilot in Lelystad and a standard raceway pond (paddle wheel driven). This crossover was made because the pond configuration in Lelystad is atypical in several aspects (maximum depth of the Lelystad pond which is 70 cm, the mixing of the pond is realised with a propeller and the feed-in of flue gas with a blower) and generalising the data makes the open pond model more applicable to other situations. The pond size of 1,000 $m^2$ was selected as estimation of most probable commercial size, and data from the 250 $m^2$ algae pilot ponds have been recalculated to that size. The depth of the model ponds is the standard 30 cm instead of the 80 cm that is used for the pilot ponds. With such a basic lay-out the algae production is estimated at 1.5% conversion of light energy to glucose bound energy, which is within the range of conversion efficiencies used in literature for modelling open ponds (Norsker, 2011 and Slegers, 2013) and results in (for Dutch climate conditions) an overall production of ~20 tonnes of dry biomass per hectare per year, similar to production levels reported for large scale open pond pilots within the Netherlands (Verberkt, 2012 and Uijterlinde, 2010).

Harvesting from the pond is included in the model as direct centrifugation to a relatively high density algae paste (150 kg dm/$m^3$). The parameters for operation and costs used in the model are taken from information found on the Evodos website and brochures, more specifically for the Evodos type 25. A recovery efficiency of 95% is estimated and to avoid the loss of large volumes of water a recycling stream of 90% is introduced.

Algal biomass composition is partly derived from Phyllis2, a database for biomass and waste developed by Energy research Centre of the Netherlands (https://www.ecn.nl/phyllis2) and on a review paper describing microalgal composition (Becker, 2007).

Investment costs of the total pilot built in Lelystad include two open ponds, a greenhouse (containing one of these open ponds and a separate space for processing), blowers, pumps, coalescers, centrifuges and pipes. The costs that could be allocated to the outdoor open pond (including a building for processing) were estimated and extrapolated to a 1,000 $m^2$ open pond. The resulting figures are the basis for the capital goods list. The total estimated investment costs for an open micro-algae pond of 1,000 $m^2$ are presented in the Table 2. These costs can be questionable, but the model allows user to change price levels according to their own information.

***Table 2:*** *Estimated investment costs for an open micro-algae pond of 1,000 $m^2$.*

| Equipment | Cost |
|---|---:|
| Reactor construction | € 5,000 |
| Mixing equipment | € 9,000 |
| Heating equipment | € 27,000 |
| Sparging equipment | € 27,000 |
| Process control | € 15,000 |
| Infrastructure open pond | € 75,000 |
| Centrifuge | € 95,000 |
| Infrastructure harvesting | € 17,000 |
| **Total** | **€ 270,000** |

### 1.2.3 Economic results basic scenario

**Assumptions**
The basic scenario uses following settings:

| | |
|---|---|
| Location: | Lelystad (the Netherlands) |
| Number of ponds: | 1 * 1,000 $m^2$ |
| Flue gas supply: | yes |
| Heat supply: | yes |
| Temperature setting: | 25 °C (year round) |
| Digestate use: | no |
| Operational: | 12 months |
| Algae species: | 100% *Chlorella vulgaris* |
| Culture crashes: | 4 times per year |

**Algae biomass production**
The yearly algal biomass production calculated by the model in this set-up is 1,538 kg of dry matter, equalling 15 tonnes dry matter per hectare pond surface.

**Economics**
With a selling price of € 35 per kg of dry matter, the Return On Investment is 10% and the pay-back time is 10 years. However, the assumed selling price for the algae biomass is based on the price in a specific high value niche market on feed additives (horses, pets, cows) in the Netherlands. This is a very high price level for algae products and at the same time a very limited market. The EnAlgae project is mainly focused on the algae energy market. In the energy market or other bulk markets this price level is unrealistic and would not support algae production.

**Cost price of algal biomass**
According to the model, the cost price for algae biomass (15% dry matter) produced in an open algae pond of 1,000 $m^2$ is € 35.92 per kg of dry matter. This high cost price mainly consists of capital good costs (72%). But even if we exclude the costs for capital goods, the cost price is almost €10 per kg of dry matter. Labour (16% of the cost price), electricity (8 %) and water (3 %) are important cost factors as well (see Table 3). Moreover, the use of heat and $CO_2$ is free in this case. The cost price could be reduced if $CO_2$ credits could be taken into account, or subsidies on heat use from a CHP (in the Netherlands, anaerobic digestion receives more subsidy if the CHP-heat is used beneficially). Of course, the question is whether this financial advantage should be allocated to the digester or to the algae pond.

***Table 3:*** *Algae cost price structure for an open pond of 1,000 $m^2$.*

| Component | Cost price | Share |
|---|---:|---:|
| Capital goods | € 26.00 | 72.4 % |
| Labour | € 5.63 | 15.7 % |
| Electricity | € 3.02 | 8.4 % |
| Water | € 0.96 | 2.7 % |
| Fertilisers | € 0.13 | 0.4 % |
| Wastewater | € 0.09 | 0.3 % |
| Land | € 0.08 | 0.2 % |
| **Totals** | **€ 35.92** | **100.0 %** |

### 1.2.4 Cost price effects other scenarios

The model is very suitable to study a range of scenarios by changing the settings in the parameter section. In this paragraph the effects of upscaling, production stops during winter, heating, and geographical location will be discussed.

**Scaling up**

In Figure 4 the results are shown of scale effects on cost price if the production capacity increases from one open pond of 1,000 $m^2$ to 10, 100 or 1,000 ponds. The contributions of different cost price factors are shown. From the graph it is clear that the cost price dramatically decreases to € 6.27 per kg dry matter. This still is too high for bulk markets but could make high value niche markets more profitable. The figure also makes clear that the electricity costs make up a greater part of the cost price as upscaling continues showing that the electricity requirement of algae production needs to decrease considerably in order to make bulk markets accessible for micro-algae.

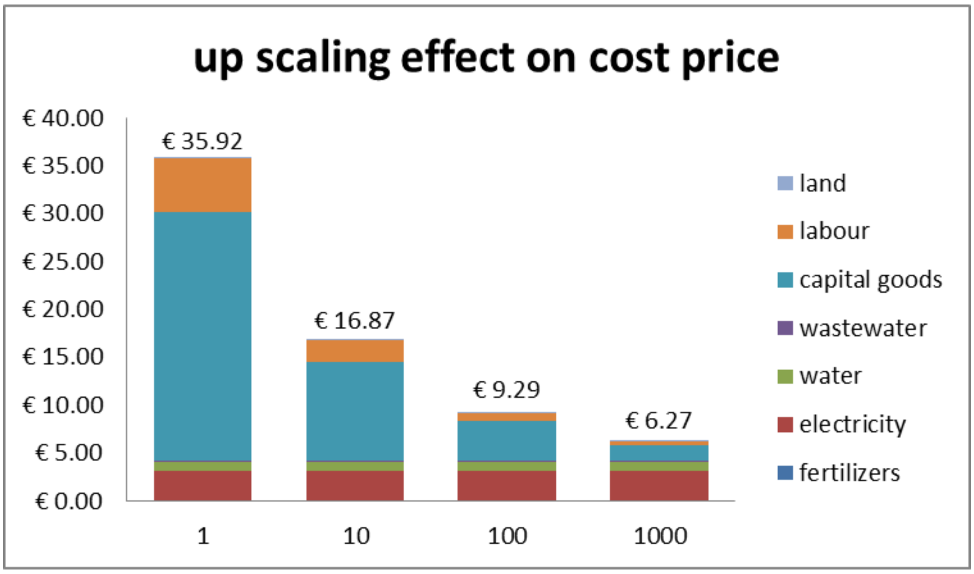


***Figure 4:** Upscaling effect on cost price in €/kg of dry matter per number of open algae ponds (1,000 $m^2$).*

**Winter production stop**

Because of the low daylight intensity during winter in the Netherlands, the relative share of the biomass production during the winter months is small, see Figure 5. Only 5 % of the total production is realised in the months November, December, and January when global radiation levels are low.

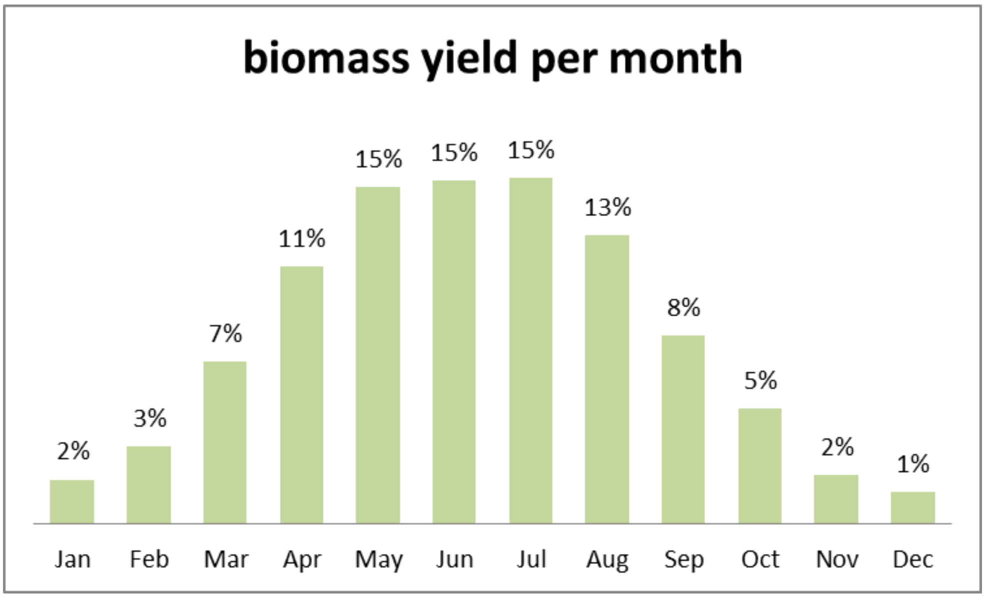


***Figure 5:*** *Biomass yield per month in the Netherlands for an open algae pond (1,000 $m^2$).*

A production stop during these months will save energy and labour costs, but the question is what it does on the cost price. According to the model (Figure 6), the cost price for algae biomass only decreases with a stop in December and the effect is limited. However, thorough cleaning the ponds in this month could prevent crashes later on in the season and could thus increase profitability.

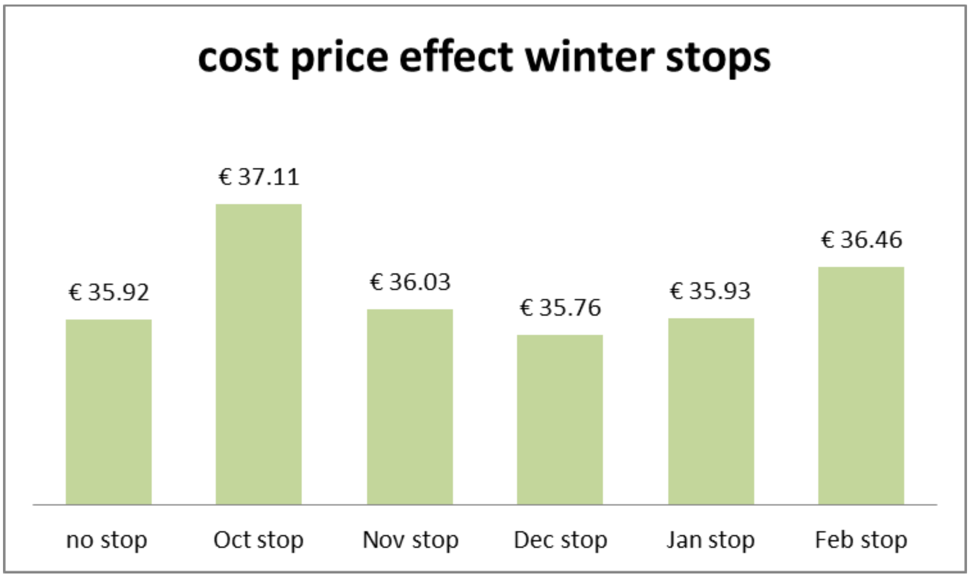


***Figure 6:*** *Cost price effect production stops during the winter for an open algae pond (1,000 $m^2$).*

**Heating**

The average temperature in the Netherlands is 10 °C, while the optimum temperature for algae biomass production is 25 °C. If the ponds are not heated, the yield decreases from 1,538 kg to 910 kg (see Figure 7 and Figure 8) and the cost price would increase from € 35.92 to € 52.33. In this pilot case heat is freely available from the CHP and the costs associated with heating are for electricity and heating equipment. As shown by the cost prices, heating is economically feasible.

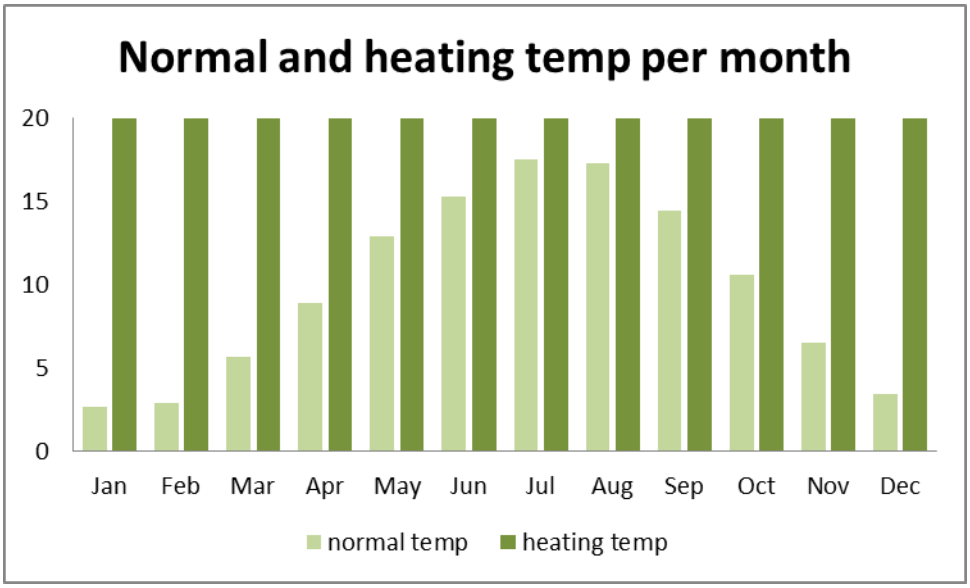


***Figure 7:*** *Normal and heating temperature for algae biomass production per month for an open algae pond (1,000 $m^2$).*

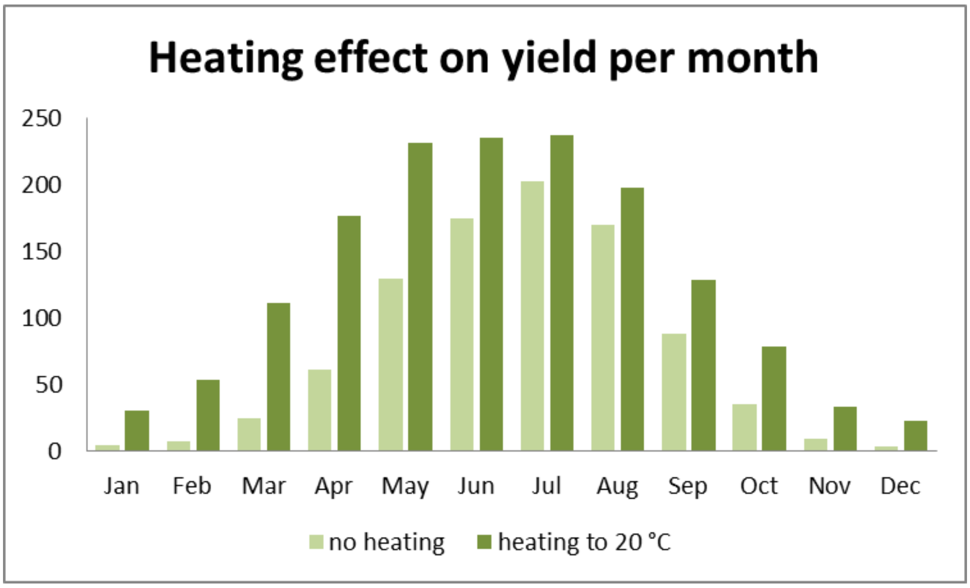


***Figure 8:*** *Heating effect on algae biomass yield per month for an open algae pond (1,000 $m^2$).*

## Geographical location

In North-West Europe the average global radiation per surface area and temperature are low, especially around winter. In other parts of the world much higher yields are possible based on more favourable light and temperature conditions. When the model is fed with climatic data from Southampton (United Kingdom), Seville (Spain) or Willemstad (Curacao) much higher yields and lower cost prices are possible, see Figure 9 to Figure 11.

For Willemstad no heating system is needed and the cost price is €17.78 per kg algae at a production capacity level of one pond of 1,000 $m^2$.

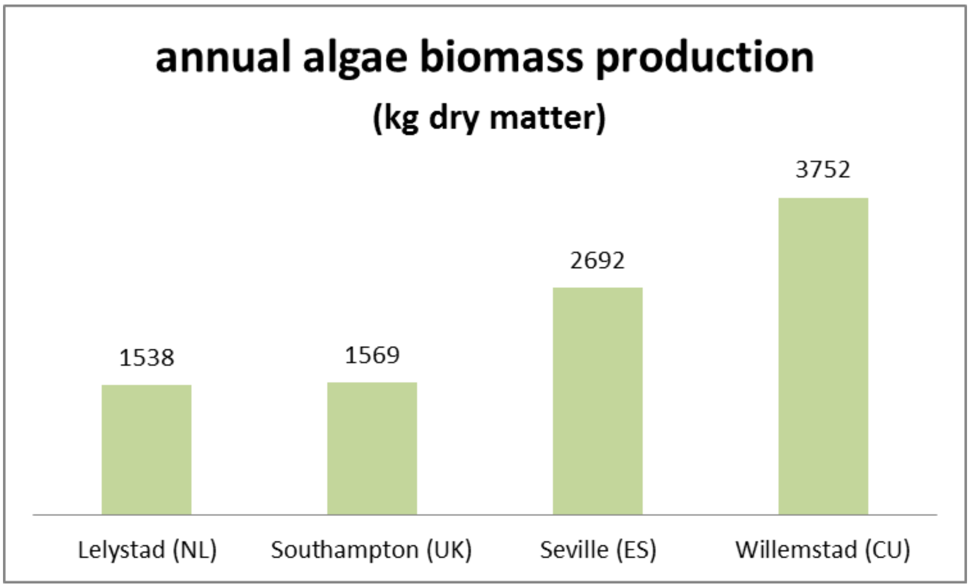


***Figure 9:*** *Annual algal biomass production at locations in Europe and South America for an open algae pond (1,000 $m^2$).*

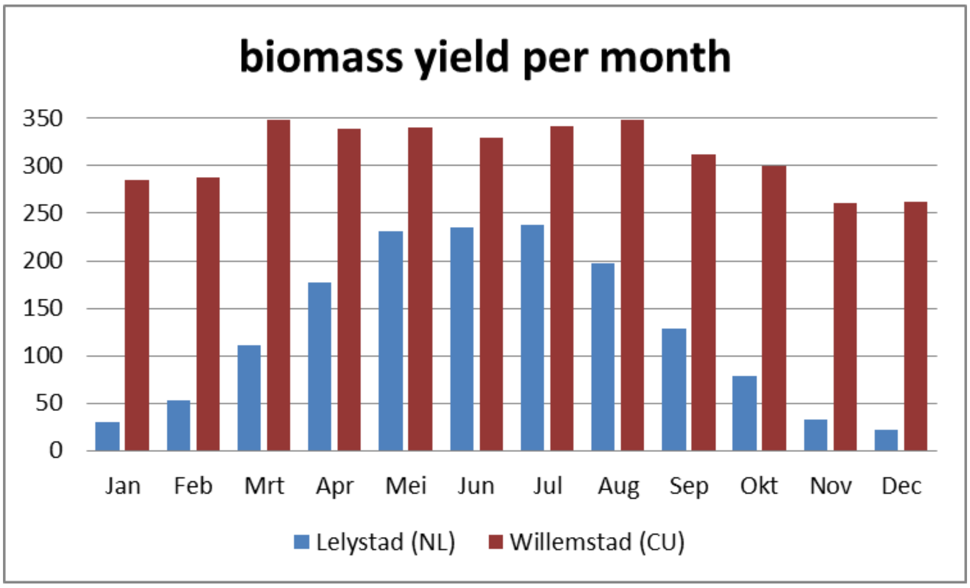


***Figure 10:*** *Algal biomass production per month in Lelystad (the Netherlands) and Willemstad (Curacao) for an open algae pond (1,000 $m^2$).*

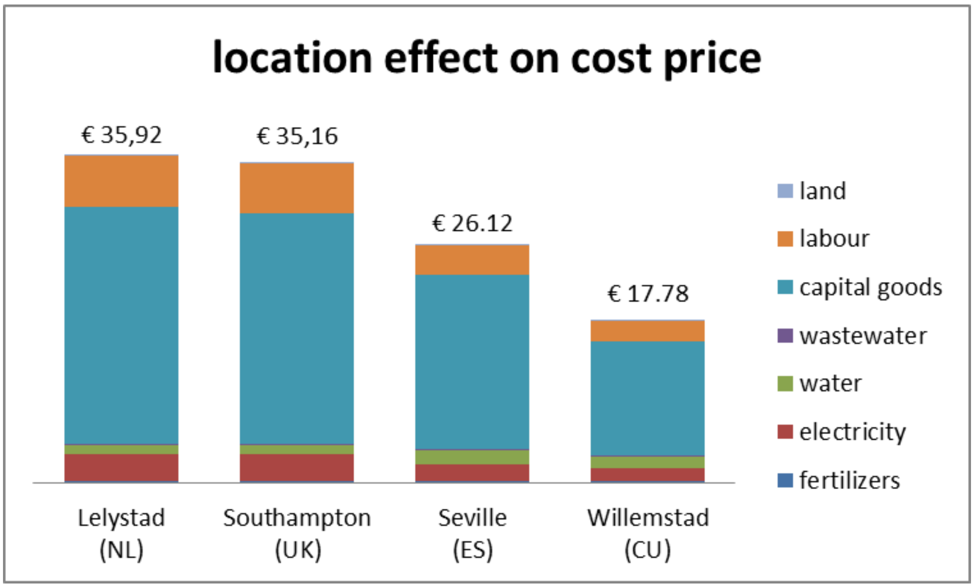


***Figure 11:*** *Cost price using climatic data from locations in Europe and South America for an open algae pond (1,000 m$^2$).*

## 1.3 Tubular photobioreactor AlgaEconomics

### 1.3.1 The tubular fence photobioreactor model

The basic EnAlgae model described in chapter 1.1 was adapted to represent a tubular photobioreactor of the fence type. The typical tubular photobioreactor fence consists of vertically stacked transparent tubes through which algae culture flows at a steady pace while collecting sunlight. During this photosynthesis period the culture produces oxygen which collects in the liquid and is removed and replaced by carbon dioxide in a degassing column, see Figure 12.

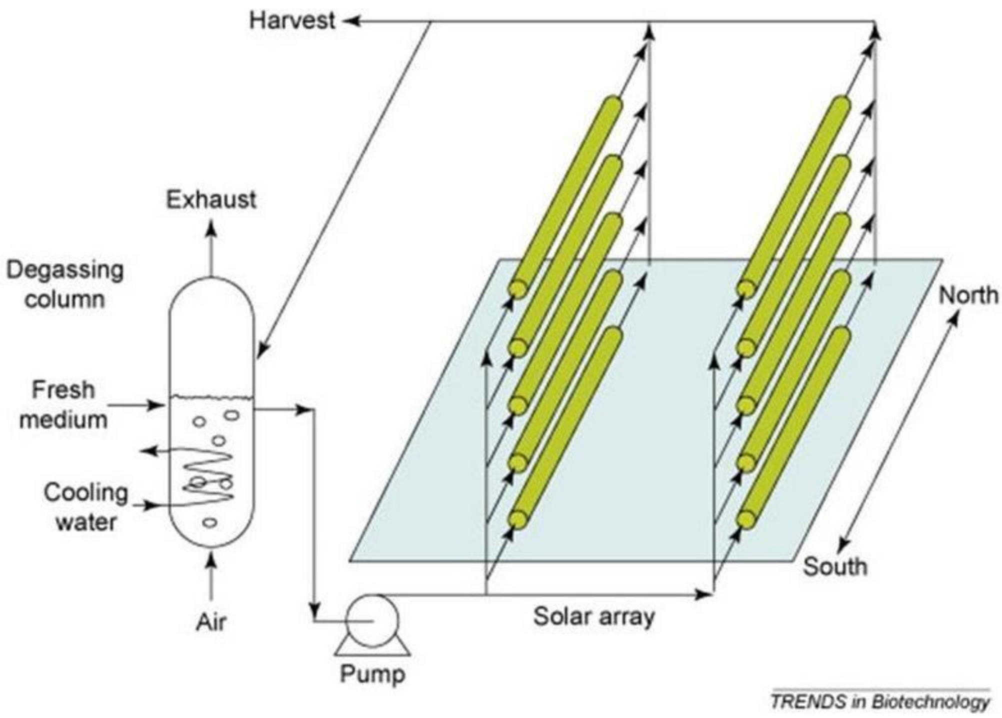


***Figure 12:*** *Example of a tubular fence bioreactor set-up. Picture from Christi (2008).*

In the EnAlgae model assumptions are made about the dimensions and layout of the photo-bioreactor. From these assumptions, details are calculated or estimated to get a better idea of the facility. These details are not derived from an optimal design and are not meant to be used for a reactor design. They do not influence the resulting algae production which is calculated from the floor surface covered by the PBR and a conversion efficiency of 3% from sunlight energy to energy captured in sugar and from there on to algae biomass (see Chapter 1.1).

The model for the tubular photobioreactor was made using the open pond business economic model as a starting point. Both model structure and calculations were kept identical as much as possible which means that the description in chapter 1.1 is mostly accurate for the tubular PBR version of the model. Exception to this is the calculation of heat use and the need for cooling. To accommodate the calculations an extra parameter is introduced; instead of a single temperature setpoint there is now a temperature setpoint range with a lower end “Tsetmin” and a higher end “Tsetmax”. Heat use is calculated as the energy needed to reach and keep the temperature at least at the lower end of the range while overheating of the algae culture above the maximum end of the range necessitates cooling. Heat use is still optional, as was the case for the open pond model, while the need for cooling is not optional, because overheating of the algae culture leads to culture crashes.

In the tubular model, calculation of heat use is based on 1) heating of the water body at dawn and 2) heat radiation loss.

1) Heating of the water body at dawn. This assumes that the culture cools down at night to the average minimum temperature ($T_{min}$) and is heated up in the morning to the bottom end of the temperature range ($T_{setmin}$). The heat needed to do so is calculated by multiplying the temperature difference with the specific heat capacity of water ($C_w$), the volume of water ($V_w$), and the number of days per month (Days).

$$H_{dawn} = (T_{setmin} - T_{min}) * C_w * V_w * Days \quad [26]$$

2) Heat radiation (during heat use) is calculated by multiplying the temperature difference between the algae culture ($T_{setmin}$) and the maximum ambient temperature ($T_{max}$) with the heat transfer rate of the PBR ($H_{pbr}$), the number of hours that heating is needed, and the number of days per month (Days). The heat radiation is set to zero if the environmental temperature is above the culture temperature. To assess the number of hours that heating is needed, (direct) sunshine hours are subtracted from the daylight hours based on the assumption that there is no need for heating when there is direct sunlight on the PBR.

$$H_{rad} = (T_{setmin} - T_{max}) * H_{pbr} * (h_{day} - h_{sun}) * Days \quad [27]$$

The heat transfer rate ($MJ.h^{-1}.K^{-1}$) is calculated as:

$$Hpbr = 10 * \frac{Rvolsur*PBRn*Usize}{0.5*Tdiam} * \frac{3600}{1000000} \quad [28]$$

where:

- $R_{volsur}$ equals the ratio of volume to surface of the tubes being 0.045 $m^3m^{-2}$,
- $PBR_n$ equals the number of PBR units,
- $U_{size}$ equals the ground surface coverage of a PBR unit being 1,000 $m^2$,
- $T_{diam}$ equals the diameter of the tube being 0.06 m
- the constant 10 equals the heat transfer coefficient of plastic material ($W\ m^{-2}\ K^{-1}$)

Calculations on which the need for cooling of the photobioreactor is based were added to the "heat use" worksheet. This calculation assumes a need for cooling whenever the energy from direct sunlight is higher than the heat loss to the environment at the upper end of the setpoint temperature range. Heat loss is calculated by multiplying the temperature difference between the algae culture ($T_{setmax}$)and the temperature of the environment ($T_{max}$) with the heat transfer coefficient of the PBR ($H_{pbr}$). This is subtracted from the energy input from direct sunlight ($H_{sun}$ in MJ per h and per unit PBR) which then results in the required cooling capacity in MJ per hour if incoming direct sunlight is larger than the heat loss by radiation. Finally, the cooling capacity needed per month is calculated by multiplying the number of sunshine hours and days with the required cooling capacity.

$$H_{cooling} = ((RADi * PBRsurf * 0.01 - (T_{setmax} - T_{max}) * H_{pbr})) * H_{sun} * Days * \frac{1}{1000} \quad [29],$$

where:
$H_{cooling}$ = the cooling capacity in GJ per month,
$RAD_i$ = the global radiation in $J.\ cm^{-2}.h^{-1}$,
$PBR_{surf}$ = the area occupied by the PBRs in $m^2$,
0.01 = a factor to convert the multiplication ($RADi * PBRsurf$) to $MJ.h^{-1}$,
$T_{setmax}$ = the maximum temperature above which cooling is required.

If the direct sunlight is not measured, an indication can be derived from 65 % of total radiation on sunny days being direct sunlight radiation (an estimate found with multiyear averages from climatic measurements in Lelystad, the Netherlands) while neglecting the amount of sunlight at overcast skies relative to that of sunny days (around 20% of the amount of sunlight on overcast days relative to that on sunny days).

### 1.3.2 Description of tubular fence photobioreactor parameters

The tubular PBR model assumes a modular PBR that can be scaled to the desired size, each module being a large tubular PBR with a volume of 45 $m^3$ covering an area of 1,000 $m^2$ (data based on Norsker *et al*., 2011). An area of 10 m by 100 m houses 16 fences of 100 meters long and 10 tubes high, with these fences spaced 63 cm apart. With these dimensions the diameter of the tubes is 6 cm. The module of 16 fences is operated as one reactor with one degassing column and one centrifugal pump.

Algae production of the PBR is estimated at 3 % conversion of light energy to glucose bound energy (Norsker et al, 2011) yielding an overall production of ~40 tonnes of dry biomass per hectare per year which is similar to estimations found in literature (Slegers, 2013 and Norsker, 2011). The density of the algae culture is estimated at 1.7 $kg/m^3$ dry weight biomass and harvesting is performed by direct centrifugation to a relatively high density algae paste (150 kg $dm/m^3$). The parameters for operation and costs used in the model are taken from information found on the Evodos website and brochures, more specifically for the Evodos type 25. A recovery efficiency of 95 % is estimated and to avoid the loss of large volumes of water a recycling stream of 90% is introduced.

The total estimated investment costs for construction of a tubular photo bioreactor of 1,000 $m^2$ are currently based on a rough estimation of various components. The investment costs are comparable to numbers mentioned in literature (Hemming *et al*., 2014).

***Table 4:*** *Estimated investment costs for tubular PBR of 1,000 $m^2$.*

| Equipment | Cost |
|---|---|
| Reactor construction | € 11,000 |
| Circulation pump | € 10,000 |
| Heating & cooling equipment | € 55,000 |
| Degassing equipment | € 45,000 |
| Process control | € 40,000 |
| Infrastructure PBR | € 75,000 |
| Centrifuge | € 33,000 |
| Infrastructure harvesting | € 17,000 |
| **Total** | **€ 286,000** |

The upscaling equation that was applied for the open pond is also used for the tubular PBR to calculate costs for capital goods, labour, and extra land requirement (equation [11]). Cost- and economic calculation methods were the same as for the open pond.

### 1.3.3 Economic results basic scenario

**Assumptions**

| | |
|---|---|
| Location: | Lelystad (the Netherlands) |
| Number of PBR units: | 1 * 1,000 $m^2$ |
| Flue gas supply: | yes |
| Heat supply: | yes |
| Cooling: | yes |
| Temperature setting: | min. 20 °C, max. 30 °C (year round) |
| Digestate use: | no |
| Operational: | 12 months |
| Algae species: | 100% *Chlorella vulgaris* |
| Reactor cleaning: | 4 times per year |

Other principles and price assumptions are discussed in the paragraphs before.

**Algae biomass production**

The yearly algae biomass production calculated by the tubular PBR model is 3,076 kg of dry matter, equalling 31 tonnes dry matter per ha PBR surface.

**Cost price of algae biomass**

According to the model the cost price for algae biomass (15% dry matter) produced in tubular PBR of 1,000 $m^2$ is €19.07 per kg of dry matter. This cost price mainly consists of capital good costs (66%). Electricity (18% of the cost price) and labour (15%) are important cost factors as well, see Table 5. The use of heat and $CO_2$ is free, like for the open pond.

***Table 5:** Algae cost price structure for a tubular PBR of 1,000 $m^2$.*

| Component | Cost price | Share |
|---|---|---|
| Capital goods | € 12.66 | 66.4 % |
| Electricity | € 3.39 | 17.8 % |
| Labour | € 2.81 | 14.8 % |
| water | € 0.06 | 0.3 % |
| Fertilisers | € 0.10 | 0.5 % |
| Wastewater | € 0.01 | 0.0 % |
| Land | € 0.04 | 0.2 % |
| **Total** | **€ 19.07** | **100 %** |

### 1.3.4 Cost price effects other scenarios

**Upscaling**

The upscaling effect on algae biomass cost price level in a tubular reactor is shown in Figure 13. At a 100 ha scale level (1,000 units), the price level is still high for bulk biomass applications like energy or feed. This also is the case with the open pond PBR. Electricity cost are key and should drop dramatically. More energy efficient procedures are needed. The model gives insight in the energy requirements (sparging, culture pump, harvest, and climate control).

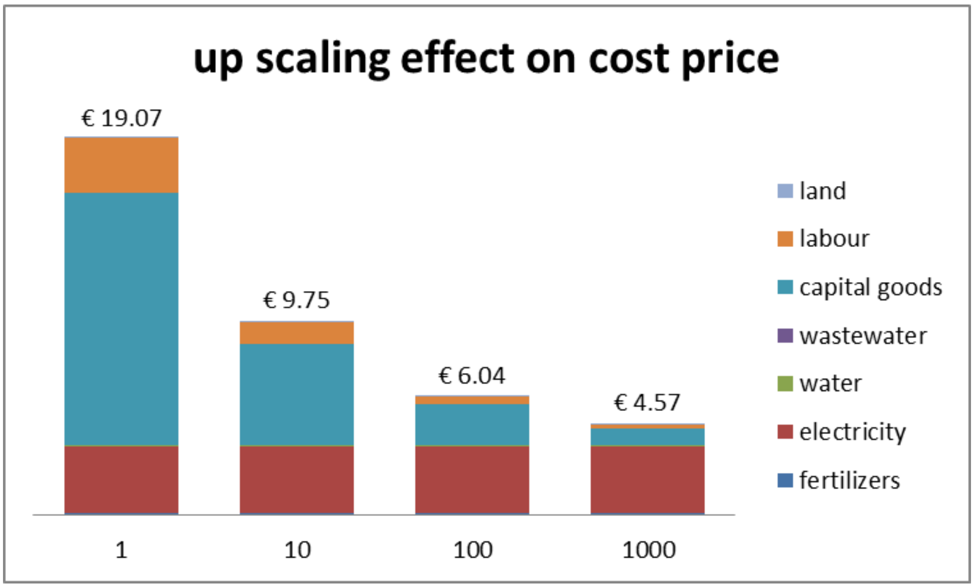


*Figure 13: Upscaling effect on cost price per number of tubular PBRs (1,000 $m^2$).*

**Winter production stop**

Because of the low daylight intensity during the winter in the Netherlands, the relative share of the biomass production during the winter months is small (Figure 14). As with the open pond PBR, the months November, December and January account for only 5% of the total production.

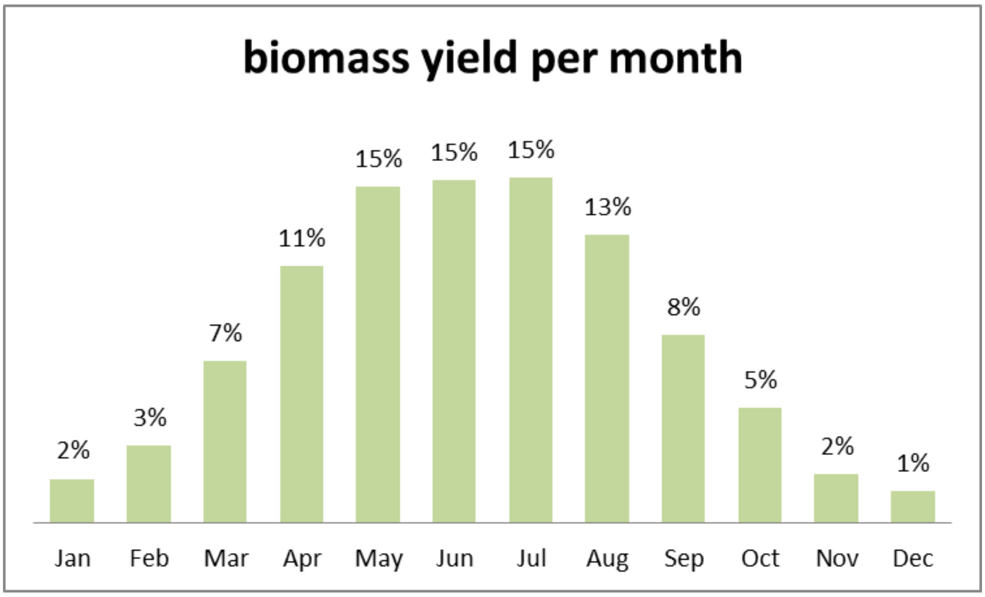


*Figure 14: Biomass yield per month in the Netherlands for a tubular PBR of 1,000 $m^2$.*

A production stop during these months could save energy and labour costs. According to the model the cost price for the algae biomass will decrease with a stop from November till February, see Figure 15.

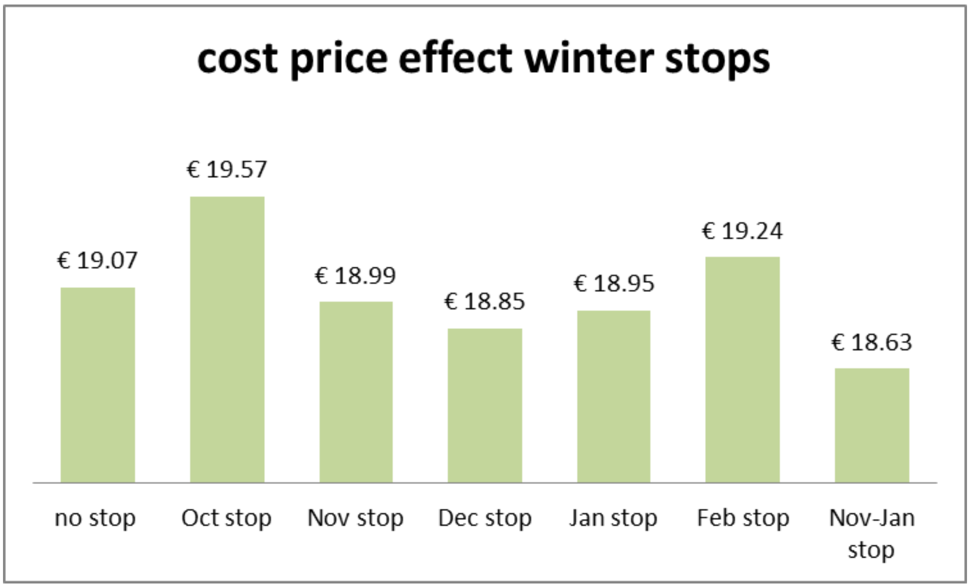


*Figure 15:* *Cost price effect production stops during the winter for a tubular PBR of 1,000 m$^2$.*

**Heating**

If the PBRs are not heated the yield would decrease from 3,076 kg to 1,820 kg (see Figure 16) and the cost price would increase from €19.07 to €31.34. In this calculation, heat is available for free (from the CHP).

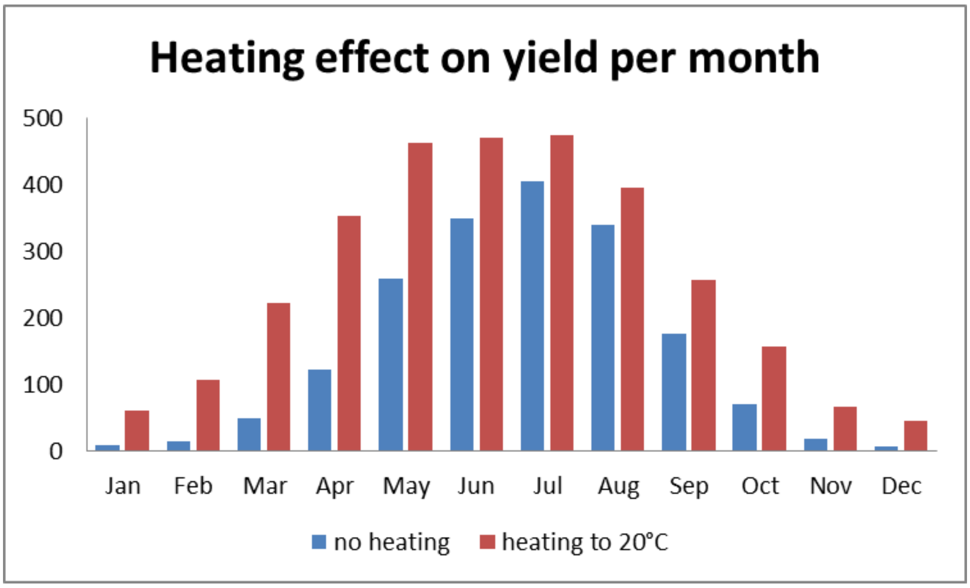


*Figure 16:* *Heating effect on algae biomass yield per month for a tubular PBR of 1,000 m$^2$.*

**Geographical location**

In North-West Europe the average daily light intensity and temperature are low. In other parts of the world much higher yields are possible based on more favourable light and temperature conditions. When the model is fed by climatic data from Southampton (United Kingdom), Seville (Spain) or Willemstad (Curacao) much higher yields and lower cost prices are possible, see Figure 17 and Figure 18. For Willemstad cost price is € 8.38 per kg for a tubular PBR of 1,000 $m^2$.

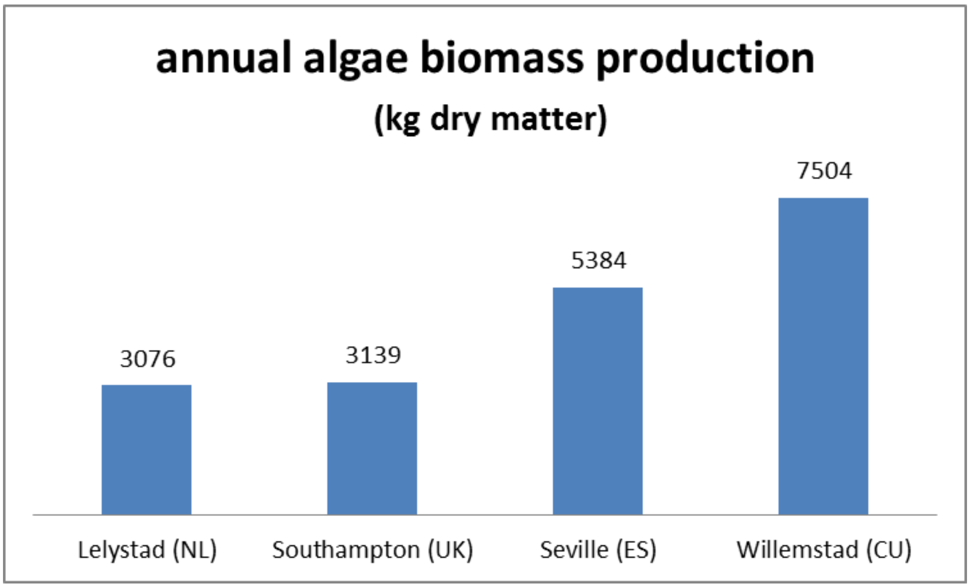


***Figure 17:*** *Annual algae biomass production at locations in Europe and South America for a tubular PBR of 1,000 $m^2$.*

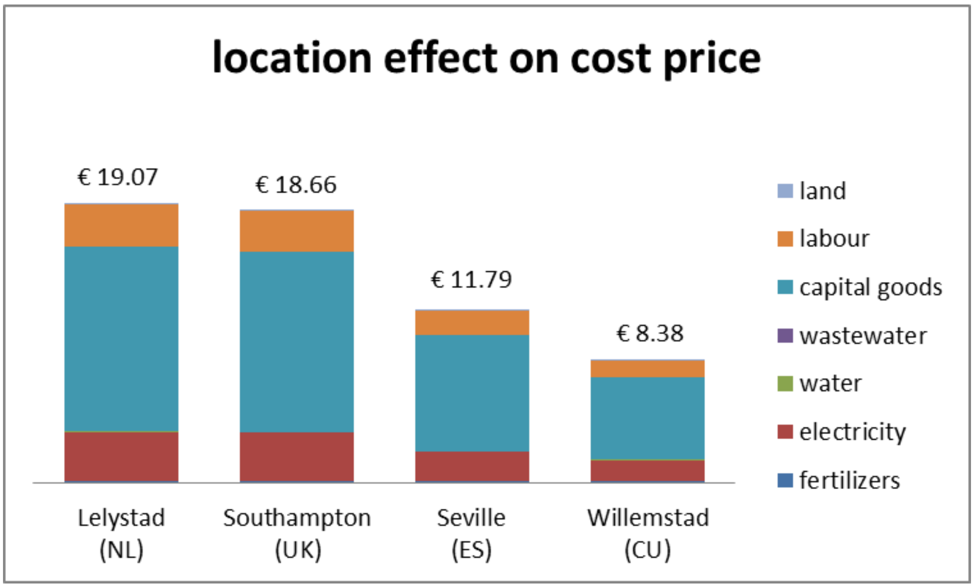


***Figure 18:*** *Cost price with climatic data from locations in Europe and South America for a tubular PBR of 1,000 $m^2$.*

## 1.4 Flat panel PBR micro-algae

### 1.4.1 The flat panel photobioreactor model

The flat panel model was adapted from the tubular photo bioreactor model, which was in turn adapted from the open pond model that is described extensively in chapter 1.1. The calculations in the flat panel model are identical to those in the tubular photobioreactor model and thus fit the description of chapter 1 except for the heat use and cooling need which are described in 1.3.

The flat panel model describes a typical airlift reactor in which $CO_2$ rich gasses are sparged from the bottom to provide both mixing and $CO_2$ supply. In comparison to the tubular reactors this flat panel airlift reactor requires less pumps for transportation of the culture, and does not require a degassing column for $O_2$ removal. The reactors do require a larger flow of air.

### 1.4.2 Description of flat panel parameters

The model assumes a modular flat panel photobioreactor that can be scaled to the desired size, each module being a PBR with a volume of 60 $m^3$ covering an area of 1,000 $m^2$. On an area of 10 by 100 meter the model assumes 13 fences of 100 meters long and 1.50 meters high, these fences being spaced 77 cm apart. The module of 13 fences is operated as one reactor. The reactor depth is set at 3 cm. Data were based on Norsker *et al.*, 2011.

Algae production of the PBR is set at 5% conversion of light energy to glucose bound energy (Norsker *et al.*, 2011) yielding an overall production of ~60 tonnes of dry biomass per hectare per year which is similar to estimations found in literature (Norsker *et al.*, 2011). The density of the algae culture is estimated at 2.1 kg/$m^3$ dry weight biomass, harvesting is performed by direct centrifugation to a relative high density algae paste (150 kg dm/$m^3$). The parameters for operation and cost given in the model are taken from information found on the Evodos website and brochures, more specifically for the Evodos type 25. A recovery efficiency of 95% is estimated and to avoid the loss of large volumes of water a recycling stream of 90% is introduced.

The total estimated investment costs for construction of a flat panel photo bioreactor of 1,000 $m^2$ are currently based on a rough estimation and require updating based on actual construction cost. However, the investment cost are in line with literature data (Norsker *et al.*, 2011).

***Table 6:*** *Estimated investment costs for a flat panel PBR of 1,000 $m^2$.*

| **Equipment** | **Cost** |
|---|---|
| Reactor construction | € 26,000 |
| Heating & cooling equipment | € 55,000 |
| Degassing equipment | € 45,000 |
| Process control | € 40,000 |
| Infrastructure PBR | € 75,000 |
| Centrifuge | € 45,000 |
| Infrastructure harvesting | € 17,000 |
| **Total** | **€ 303,000** |

**Others**
Calculations concerning labour, fertilisers, $CO_2$ and land use are the same as in the open pond model.

**Scaling up, costs and economic results**
The upscaling equation that was applied for the open pond is also used for the flat panel PBR to calculate capital good costs, labour and extra land requirement (equation [11]). Also cost- and economic calculation methods were the same as for the open pond.

### 1.4.3 Economic results basic scenario

**Assumptions**

| | |
|---|---|
| Location: | Lelystad (the Netherlands) |
| Number of PBR units: | 1 * 1,000 $m^2$ |
| Flue gas supply: | yes |
| Heat supply: | yes |
| Cooling: | yes |
| Temperature setting: | min. 20°C, max. 30°C (year round) |
| Digestate use: | no |
| Operational: | 12 months |
| Algae species: | 100 % Chlorella vulgaris |
| Reactor cleaning: | 4 times per year |

Other principles and price assumptions are discussed in the paragraphs above.

**Algae biomass production**
The yearly algae biomass production calculated by the flat panel PBR model is 5,127 kg of dry matter, equalling 51 tonnes dry matter per ha PBR surface.

**Cost price of algae biomass**
According to the model the cost price for algae biomass (harvested as an algae paste of 15% dry matter) produced in a flat panel PBR of 1,000 $m^2$ ground surface is €12.52 per kg of dry matter. This cost price mainly consists of capital good costs (55%). Electricity (31% of the cost price) and labour (14%) are important cost factors too, see Table 7. The use of heat and $CO_2$ is free, same as for the open pond.

***Table 7:** Algae cost price structure for a flat panel PBR of 1,000 $m^2$.*

| Component | Cost price | Share |
|---|---:|---:|
| Capital goods | € 6.83 | 54.5 % |
| Electricity | € 3.83 | 30.6 % |
| Labour | € 1.69 | 13.5 % |
| Water | € 0.05 | 0.4 % |
| Fertilisers | € 0.10 | 0.8 % |
| Wastewater | € 0.00 | 0.0 % |
| Land | € 0.02 | 0.2 % |
| **Total** | **€ 12.52** | **100 %** |

### 1.4.4 Cost price effects other scenarios

**Up scaling**

The upscaling effect on algae biomass cost price level in a flat panel reactor is shown in Figure 19. At a 100 ha scale level (1,000 units), the price level is still high for bulk biomass applications like energy or feed and like is the case with the open pond PBR or tubular reactor, Electricity cost are key and should drop dramatically. More energy efficient procedures are needed. The model gives insight in the energy requirements (sparging, culture pump, harvesting).

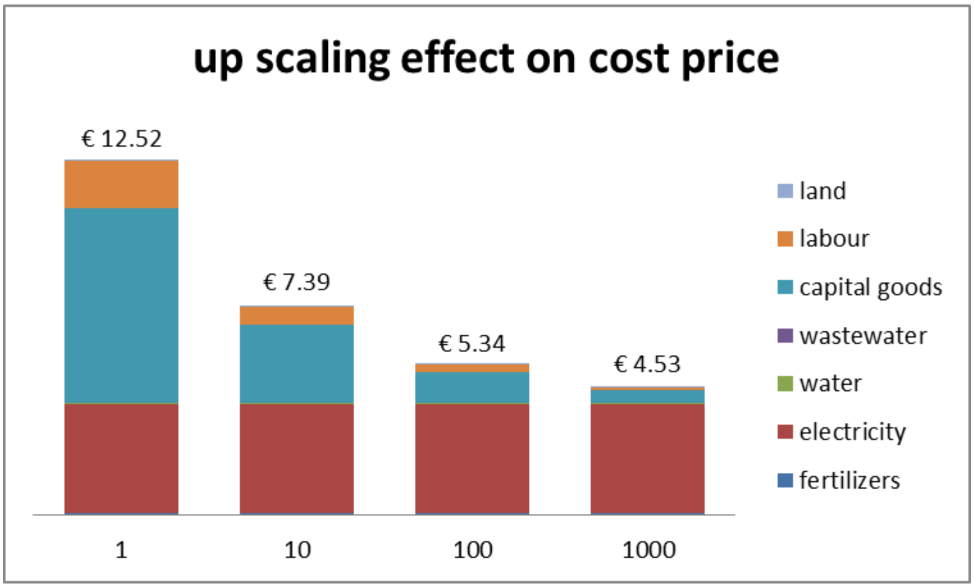


*Figure 19: Upscaling effect on cost price per number of flat panel PBRs (1,000 $m^2$).*

**Winter production stop**

Because of the low daylight intensity during the winter in the Netherlands, the relative share of the biomass production during the winter months is small, see Figure 20. During the months with the lowest level of global radiation (November –January) the total share in biomass production only amounts to 5%.

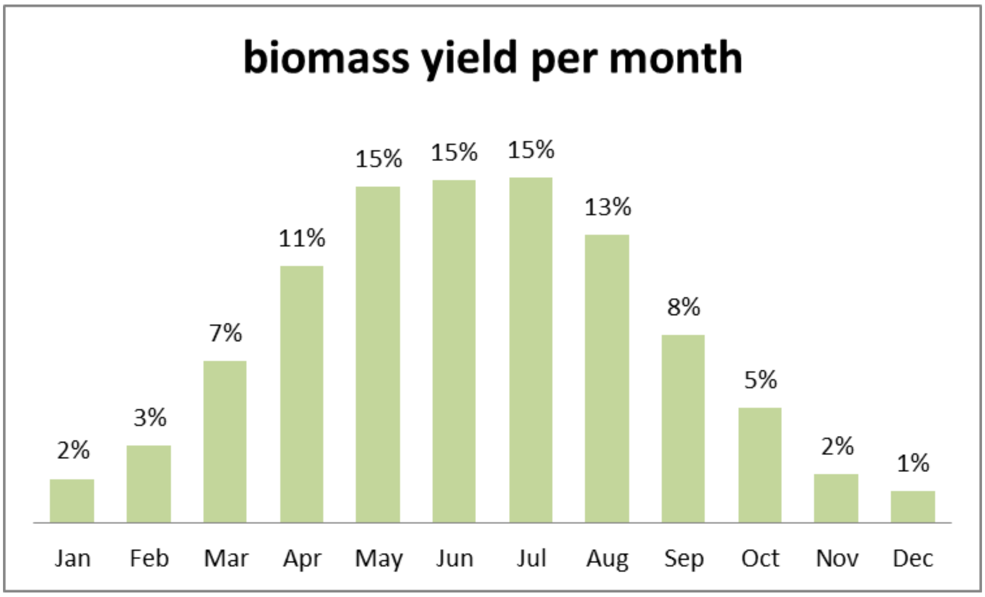


*Figure 20: Biomass yield per month in the Netherlands for a flat panel PBR of 1,000 $m^2$.*

A production stop during these months could save energy and labour costs. According to the model calculation, the cost price for the algae biomass will decrease only marginally with a stop from November to February, see Figure 21.

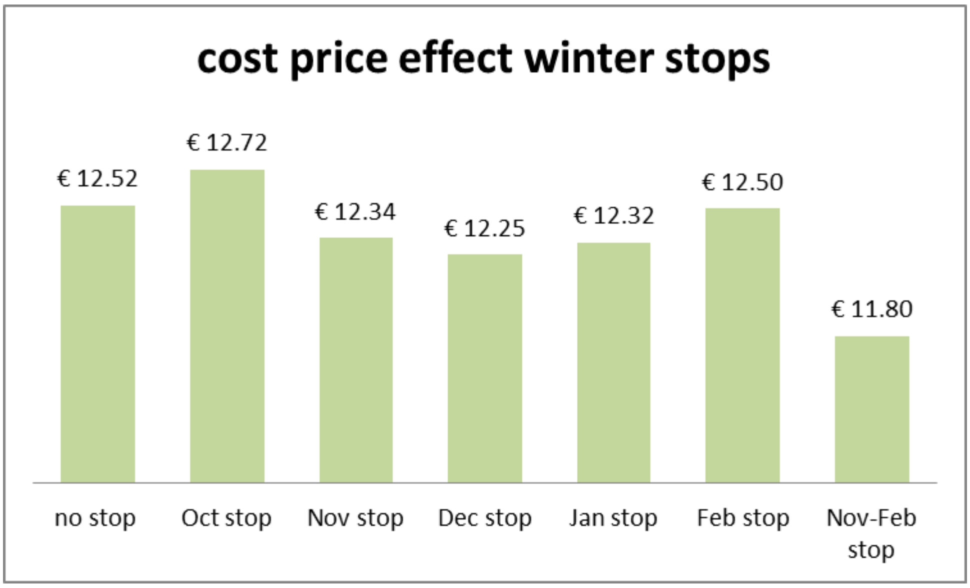


***Figure 21:*** *Cost price effect production stops during the winter for a flat panel PBR of 1,000 m$^2$.*

**Heating**

If the PBRs are not heated, the yield would decrease from 5,127 kg to 3,033 kg (see Figure 22) and the cost price per kg would increase from €12.52 to €20.69. In this calculation heat is free available (from the CHP). Figure 22 shows that the main effect of heating is in the months from February to May as in spring temperature is relatively low while global radiation is at a high level.

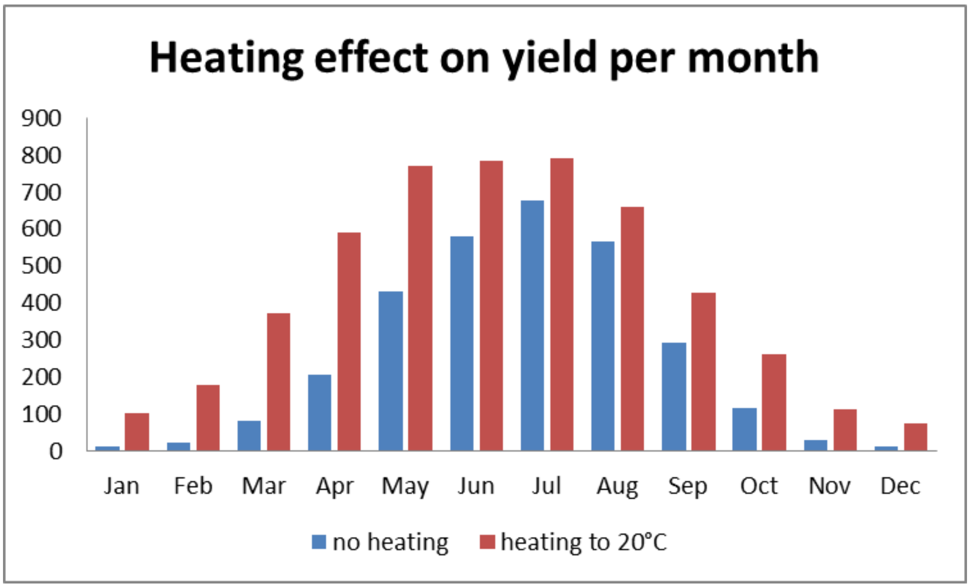


***Figure 22:*** *Heating effect on algae biomass yield per month for a flat panel PBR of 1,000 m$^2$.*

**Geographical location**

In North-West Europe the average daily light intensity and temperature are low. In some other parts of the world much higher yields are possible based on more favourable light and temperature conditions. When the model is fed by climatic data from Southampton (United Kingdom), Seville (Spain) or Willemstad (Curacao) much higher yields and lower cost prices are possible, see **Error! Reference source not found.** and Figure . For Willemstad the cost price is €5.44 per kg for a flat panel PBR of 1,000 $m^2$.

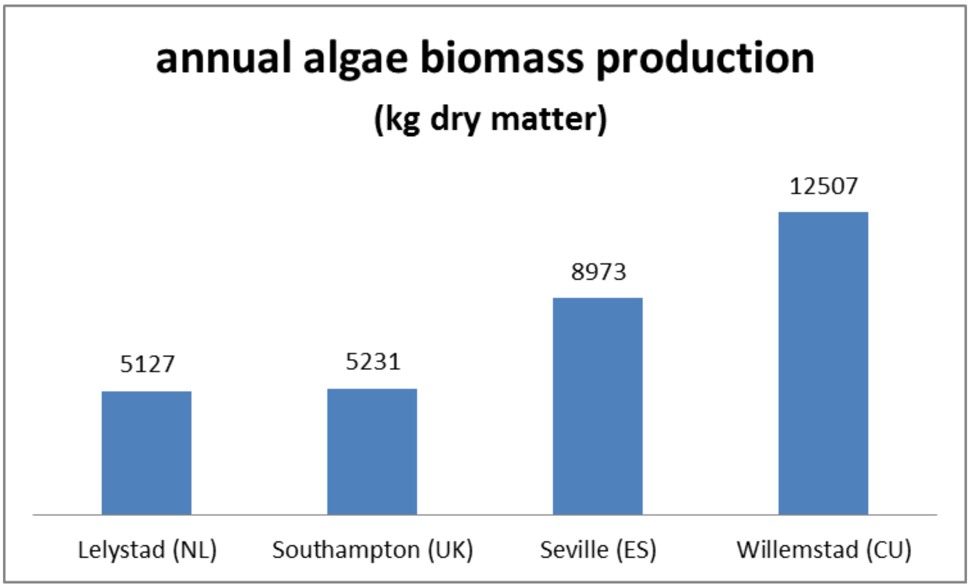


***Figure 23:*** *Annual algae biomass production at locations in Europe and South America for a flat panel PBR of 1,000 $m^2$.*

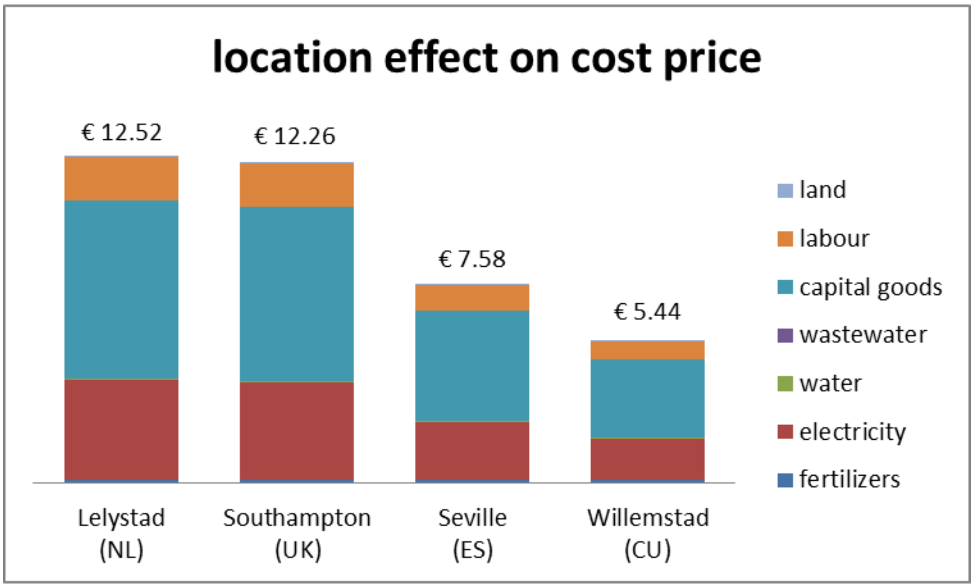


***Figure 24:*** *Cost price with climatic data from locations in Europe and South America for a flat panel PBR of 1,000 $m^2$.*

## 1.5 Comparison micro-algae production systems

In this chapter we compare the three PBR systems on both biomass production and cost price. Cost price comparisons are given for the until size area (1,000 $m^2$) as well as for a large scale (100 ha). Also, cost prices are calculated given a specific production capacity of 1,500 tonnes per annum.

### 1.5.1 Basic assumptions

The basic assumptions for each algae PBR system are listed in Table 8 and investment costs are summarised in Table 9.

***Table 8:*** *Basic parameters for each algae production system.*

| | **open pond** | **tubular PBR** | **flat panel PBR** |
|---|---|---|---|
| location | Lelystad | Lelystad | Lelystad |
| ground surface | 1,000 $m^2$ | 1,000 $m^2$ | 1,000 $m^2$ |
| unit volume | 300 $m^3$ | 45 $m^3$ | 57.7 $m^3$ |
| flue gas supply | yes | yes | yes |
| heating | yes | yes | yes |
| cooling | no | yes | yes |
| temperature setting | 20 °C | 20-30 °C | 20-30 °C |
| digestate use | no | no | no |
| operational | 12 months | 12 months | 12 months |
| algae species | 100 % Chlorella vulgaris | 100 % Chlorella vulgaris | 100 % Chlorella vulgaris |
| culture crashes | 4 | 0 | 0 |
| reactor cleaning | 0 | 4 times per yr | 4 times per yr |
| photosynthetic efficiency (PE) on daylight | 1.5 % | 3 % | 5 % |
| biomass concentration in the pond | 0.3 g/L | 1.7 g/L | 2.1 g/L |
| dry matter concentration of algae after centrifuge | 150 g/L | 150 g/L | 150 g/L |
| biomass recovery of algae | 95 % | 95 % | 95 % |
| media recovery after centrifuge | 90 % | 90 % | 90 % |
| | | | |
| **energy use:** | **open pond** | **tubular PBR** | **flat panel PBR** |
| air sparging | 7.3 kWh per month per $m^3$ of water | 28.8 kWh per month per $m^3$ of water | 96 kWh per month per $m^3$ of water |
| circulation | 3.7 kWh per month per $m^3$ of water | 176 kWh per month per $m^3$ of water | 200 kWh per month per $m^3$ of water |
| heating pump | 0.7 kWh per month per $m^3$ of water | 0.7 kWh per month per $m^3$ of water | 0.7 kWh per month per $m^3$ of water |
| cooling water pump | n.a. | 0.4 kWh/$m^3$ cooling water | 0.4 kWh/$m^3$ cooling water |
| centrifuge | 1.25 kWh/$m^3$ harvested | 1.25 kWh/$m^3$ harvested | 1.25 kWh/$m^3$ harvested |
| | | | |
| **labour:** | **open pond** | **tubular PBR** | **flat panel PBR** |
| low qualified | 2 hours/unit, per day | 2 hours/unit, per day | 2 hours/unit, per day |
| high qualified | 0.1 hours/unit, per day | 0.1 hours/unit, per day | 0.1 hours/unit, per day |
| in case of crashes low qualified | 16 hours/crash, unit | 16 hours/crash, unit | 16 hours/crash, unit |

Table 9 shows that open ponds need larger centrifuges as the algae density in the culture is much lower leading to larger amounts of water to be processed. This largely compensates for the lower construction costs.

***Table 9:*** *Investment cost per PBR en per unit of 1,000 $m^2$.*

| investment costs per unit: | open pond | tubular PBR | flat panel PBR |
|---|---|---|---|
| **raceway/PBR:** | | | |
| Reactor construction | € 5,000 | € 11,000 | € 26,000 |
| Mixing equipment | € 9,000 | € 10,000 | € 0 |
| Heating/cooling equipment | € 27,000 | € 55,000 | € 55,000 |
| degassing equipment | € 27,000 | € 45,000 | € 45,000 |
| Process control | € 15,000 | € 40,000 | € 40,000 |
| Infrastructure | € 75,000 | € 75,000 | € 75,000 |
| **subtotal** | **€ 158,000** | **€ 236,000** | **€ 241,000** |
| | | | |
| **harvesting:** | | | |
| Centrifuge | € 95,000 | € 33,000 | € 45,000 |
| Infrastructure | € 17,000 | € 17,000 | € 17,000 |
| **subtotal** | **€ 112,000** | **€ 50,000** | **€ 62,000** |
| | | | |
| **Total** | **€ 270,000** | **€ 286,000** | **€ 303,000** |

### 1.5.2 Biomass production

Biomass production per area in a tubular PBR is twice and in a flat panel PBR more than three times as much as in an open pond, see Figure 25. Differences in production between production systems can be attributed fully to differences in Photosynthetic Efficiency (1.5%, 3% and 5% respectively).

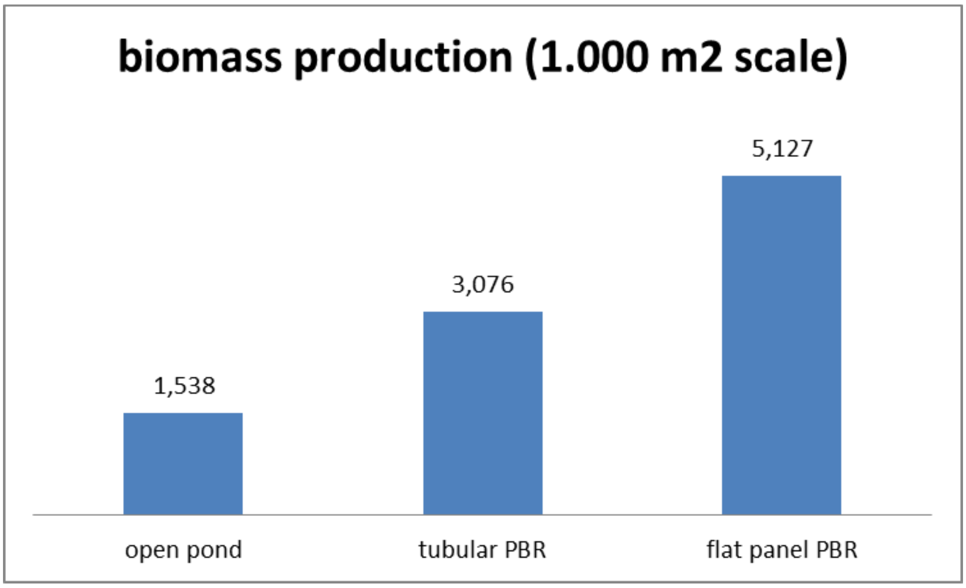


***Figure 25:*** *Biomass production (kg dry matter per year) for the three PBRs (1,000 $m^2$ scale).*

### 1.5.3 Cost price basic unit

The cost price for algae biomass produced in closed systems is much lower than in open ponds, mainly because of the higher yield levels in closed production systems (Figure 26 and 10). The results show that the cost price of closed PBRs show a higher share of electricity costs and higher electricity cost per unit production as such. The capital investment cost of closed PBRs is slightly higher than in open ponds but the contribution to the cost price drops substantially from open to closed photobioreactors with flat panels having the smallest investment cost per kg produced. However, in all cases investment cost have the highest share, but – as shown earlier in this report – scaling up will decrease the share of investment cost in total cost.

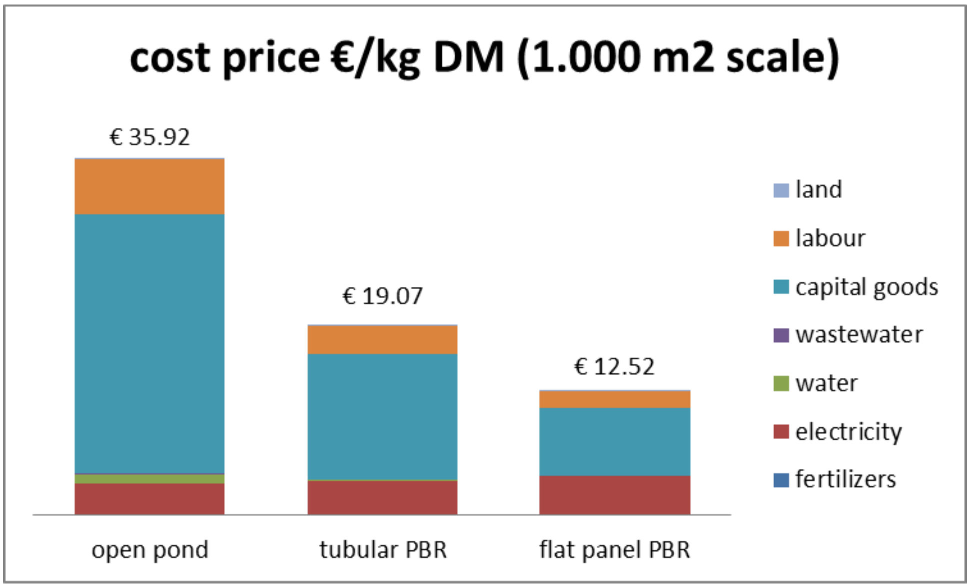


***Figure 26:*** *Algae cost price for the three algae production systems (1,000 m$^2$ scale).*

***Table 10:*** *Algae cost price structure for the three algae production systems (1,000 m$^2$ scale).*

| | open pond | | tubular PBR | | flat panel PBR | |
|---|---|---|---|---|---|---|
| number of units (1,000 m$^2$) | 1 | | 1 | | 1 | |
| biomass production (kg dm) | 1,538 | | 3,076 | | 5,127 | |
| | | | | | | |
| **cost price structure:** | | | | | | |
| capital goods | € 26.00 | 72 % | € 12.66 | 66 % | € 6.83 | 55 % |
| labour | € 5.63 | 16 % | € 2.81 | 15 % | € 1.69 | 13 % |
| electricity | € 3.02 | 8 % | € 3.39 | 18 % | € 3.83 | 31 % |
| water | € 0.96 | 3 % | € 0.06 | 0 % | € 0.05 | 0 % |
| fertilisers | € 0.13 | 0% | € 0.10 | 1 % | € 0.10 | 1 % |
| wastewater | € 0.09 | 0 % | € 0.01 | 0 % | € 0.00 | 0 % |
| land | € 0.08 | 0 % | € 0.04 | 0 % | € 0.02 | 0 % |
| **total cost price** | **€ 35.92** | **100 %** | **€ 19.07** | **100 %** | **€ 12.52** | **100 %** |

### 1.5.4 Cost price 100 ha

The models allow for scaling up to be estimated based on equation [11]. This approach is an extrapolation of what is known, but economic data from larger scale operations are missing. It can be expected from equation [11] will produce decreasing cost when scaling up algae production facilities. This applies to fixed cost and variable cost like labour and land use. Costs for electricity, water, waste water and fertilisers are insensitive to the scale level as they are directed related to the production of a unit of PBR. Electricity costs define the major part of the costs for each system on a large scale (Figure 27). For the tubular and flat panel PBR electricity and capital good costs make up more than 90% of the cost price. For the open pond, costs for water are still substantial besides electricity and capital goods. The high water use in open ponds is related to evaporation, a phenomenon not occurring in closed PBRs, while water for cooling only relates to closed PBRs but at a much smaller volume.

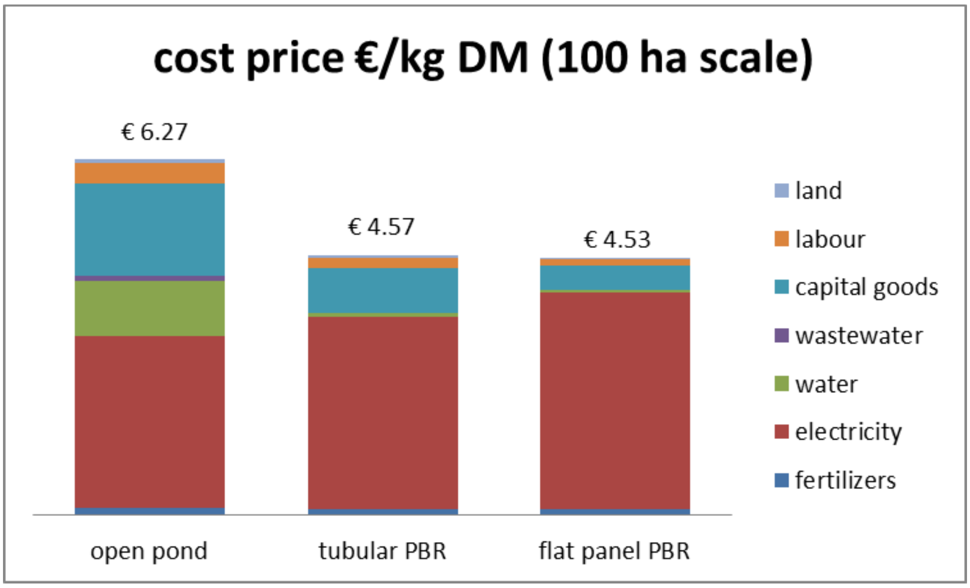


***Figure 27:*** *Algae cost price for the three algae production systems (100 ha scale).*

***Table 11:*** *Algae cost price structure for the three algae production systems (100 ha scale).*

| | **open pond** | | **tubular PBR** | | **flat panel PBR** | |
|---|---|---|---|---|---|---|
| number of units (1,000 $m^2$) | 1,000 | | 1,000 | | 1,000 | |
| biomass production (tonne dm) | 1,538 | | 3,076 | | 5,127 | |
| | | | | | | |
| **cost price structure:** | | | | | | |
| electricity | € 3.02 | 48 % | € 3.39 | 74 % | € 3.82 | 84 % |
| capital goods | € 1.64 | 26 % | € 0.80 | 17 % | € 0.43 | 10 % |
| water | € 0.96 | 15 % | € 0.06 | 1 % | € 0.05 | 1 % |
| labour | € 0.35 | 6 % | € 0.18 | 4 % | € 0.11 | 2 % |
| fertilisers | € 0.13 | 2 % | € 0.10 | 2 % | € 0.10 | 2 % |
| wastewater | € 0.09 | 1 % | € 0.01 | 0 % | € 0.00 | 0 % |
| land | € 0.07 | 1 % | € 0.03 | 1 % | € 0.02 | 0 % |
| **total cost price** | **€ 6.27** | **100 %** | **€ 4.57** | **100 %** | **€ 4.53** | **100 %** |

### 1.5.5 Cost price high biomass production volumes

The cost price calculations in Table 11 were based on a given area of production resulting in different production volumes based on photosynthetic efficiency. It's questionable whether the focus on area is relevant given the relative low costs for land use in the final cost price. It may be more interesting to compare cost prices for the different PBRs at a certain planned production volume. In the end, production is based on the volume of the market and not on the amount of land available. We have chosen a production target of 1,500 tonnes of algae dry matter, Our model calculations have shown that in order to produce this 1,000 open ponds are needed, 500 tubular PBRs or 300 flat panel PBRs. The question now is what effect this will have on the cost price based on the PBR used. The results (Figure 28 and Table 12) show that the cost price between flat panels and tubular panels is comparable although the flat panel systems profits less from economy of scale because less units are required. However, the total investment cost for flat panel algae production is lower and this can be attractive from a risk point of view.

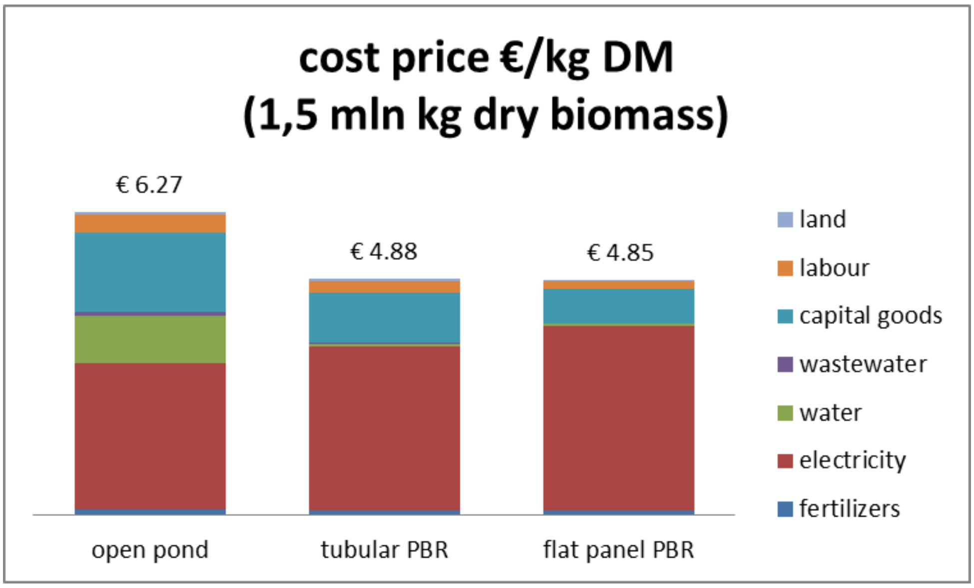


***Figure 28:*** *Algae cost price for the three algae production systems (1,500 tonnes kg production volume).*

***Table 12:*** *Algae cost price structure for the three algae production systems (1,500 tonnes dry matter per year production volume).*

| | **open pond** | | **tubular PBR** | | **flat panel PBR** | |
|---|---|---|---|---|---|---|
| number of units (1,000 $m^2$) | 1,000 | | 500 | | 300 | |
| biomass production (tonne dm) | 1,538 | | 1,538 | | 1,538 | |
| **cost price structure:** | | | | | | |
| electricity | € 3.02 | 48 % | € 3.39 | 70 % | € 3.82 | 79 % |
| capital goods | € 1.64 | 26 % | € 1.05 | 22 % | € 0.70 | 14 % |
| water | € 0.96 | 15 % | € 0.06 | 1 % | € 0.05 | 1 % |
| labour | € 0.35 | 6 % | € 0.23 | 5 % | € 0.16 | 3 % |
| fertilisers | € 0.13 | 2 % | € 0.10 | 2 % | € 0.10 | 2 % |
| wastewater | € 0.09 | 1 % | € 0.01 | 0 % | € 0.00 | 0 % |
| land | € 0.07 | 1 % | € 0.03 | 1 % | € 0.02 | 0 % |
| | **€ 6.27** | **100 %** | **€ 4.88** | **100 %** | **€ 4.85** | **100 %** |

# 2 Algal refinery for the energy-market

To produce energy products from algae further process steps are required. To calculate the economics of these processes, separate Excel models for biodiesel, ethanol and methane are developed within the EnAlgae project. The downstream process, the model and the economic results are described in the next paragraphs.

The biodiesel and ethanol models are based on 10 tonnes algae dry matter, but the models allow for scaling up. The methane production is based on the process of anaerobic digestion. Because the minimum algae volume to produce methane in a digester is about 1,500 tonnes algae (see chapter 2.3) the comparison of the cost price for the energy carriers will be based on this large algae biomass volume. All models assume a selling price of algae, functioning as a raw material input price in the processing models.

## 2.1 Diesel

### 2.1.1 Downstream processing of algae to diesel

The process involves the recovery of intracellular lipid, and the subsequent conversion to the fatty acid methyl ester (FAME) via a transesterification reaction to make biodiesel. Harvested algae must be dried and milled prior to supercritical carbon dioxide ($sCO_2$) extraction to separate lipid from the cells. The cells are milled to a fine powder to improve packing within the $sCO_2$ reactor, and to enhance the extractability of the neutral lipid. Under conditions of high pressure and elevated temperature, $CO_2$ acts as solvent to remove neutral lipids, such as triacylglycerides. A pressure drop is used to collect various fractions from the biomass. In this example, only the lipid fraction is considered. Other by-products would be species dependent, e.g. pigments or carotenoids.

Following $sCO_2$ extraction, the remaining biomass can be used for other purposes, for example the protein fraction could be considered as a revenue-generating side stream. The lipid is collected and refined prior to transesterification using methanol and a catalyst. Further purification is required to reduce the contaminants present in the FAME mixture to acceptable levels for EN14105 biodiesel. Glycerol, which is a by-product of the transesterification reaction, can be purified and sold. Alternatively, the crude material can be used as an additive in anaerobic digestion to improve yield.

The process alternative to $sCO_2$ extraction is solvent extraction, either using a single solvent such as hexane, or employing a mixture of polar and non-polar solvents (e.g. hexane and IPA). The $sCO_2$ route is preferable as the residual biomass is not contaminated with organic solvents and can be used as a supplement for the feed industry.

### 2.1.2 Explanation of the model

**Steps involved in DSP for biodiesel**

1. *Drying*
   All assumptions made for drying are the same as for the ethanol process. Dry biomass is preferable for $sCO_2$ extraction, as the presence of water can create a boundary layer, making it more difficult to extract lipids. The model is generic and does not differentiate between saltwater and freshwater species – a washing or salt removal step may be required for marine species of microalgae prior to drying. At present, physical methods of drying algae include drum drying and spray drying. The model is agnostic to technology, and considers the energetic burden of

removing water. Input to the drying process step is an algae paste of 15 % dry matter content. The output is a dried product (BMdry):

$$BMdry = BMwet * \frac{DM\%wet}{DM\%dry} \quad [30],$$

where $BM_{wet}$ equals the biomass weight of the algae paste, $DM_{\%wet}$ equalling the dry matter percentage of the algae paste (15 %) and $DM_{\%dry}$ the dry matter percentage of the dried product (80 %).

The energy required for drying ($E_{dry}$ in kWh) is calculated as follows:

$$Edry = \{(100\% - DM\%wet) * BMwet\} - (100\% - DM\%dry) * BMdry\} * 2.26/3.6 \quad [31],$$

where 2.26 stands for the energy required to evaporate water at 100 $^0$C in MJ per kg of water and 3.6 for the conversion of MJ to kWh.

2. *Milling*
   Breaking open cell walls will enhance extraction of intracellular contents such as neutral lipids. In this step, a ball mill is considered for mechanical cell breakage. The capacity of the ball mill is based on 12.5 tonnes dried algae paste (10 tonnes of 80 % w/w dry matter) and 8,000 operational hours. Electricity required for ball milling is 1.87 kWh/kg dry biomass (Balasundaram et al, 2012), and the assumed cell disruption efficiency is 95 %.
   The percentage of cells disrupted is dependent on species and the composition of the cell wall. The energy use for this process step ($E_{mill}$) can be calculated like:

$$Emill = 1.87 * BMdry * \ DM\%dry \quad [32]$$

The ball milling is assumed to have a certain efficiency for which a parameter is introduced ($PERC_{mill}$) having a value of 95 %. The low value labour for ball milling is set at 400 hours per year and the high value labour at 12 hours per year.

Alternatives for cell disruption include wet milling, enzyme lysis, sonication and homogenisation, which can be performed with wet material. The presence of water in a $sCO_2$ reactor reduces extraction efficiency. While this may be overcome with the use of co-solvents to $CO_2$, this is not considered here.

3. *Supercritical $CO_2$ extraction*
   In the supercritical $CO_2$ process step $CO_2$ is passed through a sub-cooler to become liquid so it can be pumped towards the vessel containing algal biomass. The temperature and pressure are raised to enable the $CO_2$ to become supercritical. In this instance, conditions are estimated to be 500 bar, and 50 °C. The base case for the model (10 tonnes of dry algal biomass) assumes a vessel size of 10 L, the packing density of the material determines the throughput of algal biomass. The $CO_2$ extraction is a batch process. The model assumes 8 kg per batch is packed into the reactor (BATCH). The energy requirement ($E_{scrit}$) is calculated as follows:

$$Escrit = 0.8 * BMdry \quad [33],$$

where 0.8 is a the energy consumption (kWh $kg^{-1}$) for supercritical $CO_2$ extraction as given by Pellerin (undated). The amount of $CO_2$ required (CO2) can be calculated like:

$$CO2 = Nbatch * (100\% - CO2rec) * FLOW * Fh \quad [34]$$

and

$$Nbatch = \frac{BMdry}{BATCH} \quad [35],$$

where $N_{batch}$ is the number of batches per year, CO2 is the recovery rate of $CO_2$ from the process (value of 80 %[2] assumed[3]), FLOW being the flowrate of $CO_2$ in the system and $F_h$ equalling the flow time (h) matching the biomass processing time in the reactor for 2 hours. This extraction is assumed to yield an efficiency of 95%[4]. While this may vary slightly for different species of algae, due to the nature of cell walls or lipid location, these conditions are used as a representative extraction. The depressurisation steps will typically be carried out at 100-150 Bar with the potential for second step at 50-70 Bar.

In the model $CO_2$ is assumed to be supplied from an outside source, however if there was an emitter on site (e.g. from a CHP plant) then $CO_2$ could be recovered once impurities have been scrubbed from the flue gas. The percentage of unrecovered $CO_2$ is caused by depressurisation of the system for cleaning and down-time, and by small losses when batches are complete.

In this scenario, the model simplifies the process by examining the extraction of only one product, triacylglycerides from the algal biomass. In practice, other products such as pigments could be extracted simultaneously, and a second depressurisation step added to collect a different fraction. Owing to the huge diversity of microalgae, a single product probably is not correct but it is assumed here for simplicity reasons.

The amount of lipid in the algae determines the yield of oil and, obviously, a higher content result in a higher efficiency of the whole process. Values of up to 70% lipid have been used in literature when estimating algal biodiesel productivity[5]. The biomass that remains in the extraction vessel can be removed and sold as a protein-rich feed additive.

The amount of oil extracted (OIL in kg per year) with this procedure can be calculated as follows:

$$OIL = BMdry * DM\%dry * LIPID * EFFextract \quad [36],$$

Where LIPID equals the lipid content (%) of the algae dry matter and $EFF_{extract}$ equals the extraction efficiency (%) which is assumed as 95 %.

4. *Refining*

Lipid material extracted from algal biomass must be refined prior to transesterification, to separate any membrane lipids and chlorophyll from the triacylglycerides (TAG). The process for this is modelled on the degumming step for other plant oils. While the use of supercritical $CO_2$ can be optimised for extraction of neutral lipids, the presence of chlorophyll must be minimised to meet the EN 14105 regulations on biodiesel. In this model, the presence of contaminants is assumed to be such that this additional step is necessary, with the associated losses of material, and energy consumption.

Refining is carried out by adding a small quantity of water (4 % w/w lipid) and 85 % phosphoric acid at 0.25 %. An addition of 0.05 % citric acid is also included to complex with any non-

[2] Commercial information provided by SciMed, suppliers of $sCO_2$ equipment
[3] http://www.isasf.net/fileadmin/files/Docs/Versailles/Papers/N1.pdf
[4] This is based on personal communication with Professor Ray Marriott, Bangor Biocomposites Centre, who operates an $sCO_2$ pilot.
[5] c.f. Chisti 2007

hydratable phosphatides. Caustic soda is used as a neutralising agent. The refining step is carried out at 65 °C in a mixer settler tank. This uses gravity to separate the two phases. The model assumes that all of the salts will be recovered in the aqueous phase, and that fatty acids/phosphatides/TAG representing a 1% by volume loss of lipid from the input stream will occur. The quantity of fatty acids will vary as a function of the algal species, extraction methodology and storage conditions. The model assumes a best case scenario.

The energy required for this process step is the sum of the energy needed to warm up the material from ambient temperature ($T_{amb}$) to the 65 °C target value ($E_{warm}$) and the energy to mix the material for a period of time for mixing in the mixer settler tank ($E_{mix}$).

$$Ewarm = 2.2 * (65 - Tamb) * (OIL + W + CHEM)/3{,}600 \quad [37],$$

Where W equals the amount of water added in kg.$y^{-1}$ and CHEM being the weight of chemicals added (kg.$y^{-1}$) and 2.2 the specific heat capacity of plant oil (soy bean oil) in KJ.$K^{-1}$.$kg^{-1}$. The parameter 3,600 is used for conversion of kJ to kWh. $E_{mix}$ is calculated with:

$$Emix = CAPmix * VESSEL * Nbatch * TIMEres \quad [38]$$

in which $CAP_{mix}$ equals the mixing capacity required (5 kW.m-3 matching the base case with 10 tonnes algae dry mass), $TIME_{res}$ being the residence time in the mixer settler tank, $N_{batch}$ equals the number of batches based on 8,000 operational hours and 8 hours feeding time for the mixer settler (FT) while VESSEL is the mixer settler volume calculated as:

$$VESSEL = (FT * \frac{(OIL+W+CHEM)}{8{,}000})/950 \quad [39],$$

in which 950 equals the density of plant oil (kg.$m^{-3}$).

It is assumed that in this process 1% of the oil is being lost and that only 99% is entering into the last phase, the transesterification (meaning: 0.99*OIL represented as $OIL_{ref}$).

5. *Transesterification*

There are two possible routes for transesterification: acid and base catalysed. In the model, base catalysed (using potassium hydroxide) transesterification is used. Optimisation experiments suggest that methanol must be present in excess as the reaction between the alcohol and lipid is reversible. The model assumes a 6:1 molar ratio of methanol to lipid. This is based on the transesterification of palm oil (Darnoko & Cheryan, 2000). Potassium hydroxide is the catalyst, and is added at 1% w/w basis of lipid entering the process. The reaction is assumed to run for 1 hour, whereby 98% conversion of triacylglyceride to FAME (fatty acid methyl esters) is reached. For the purposes of estimating quantities, the molar mass of triacylglyceride is assumed to be that of triolein. In practice there will be a mixture of monoacylglycerides, diacylglycerides and triglycerides, but for simplification, triacylglycerides are assumed to be the only species. Power consumptions for mixing is estimated from the range expected for a stirred tank[6] of 1-10 $Wm^{-3}$.

The energy required for mixing ($E_{mix,trans}$ in kWh per year) is calculated as:

$$Emix,trans = CAPmix,trans * VESSELtrans * Nbatch,trans * TIMEres,trans \quad [40]$$

[6] Bioprocess Engineering Principles, 2nd Edition | Pauline Doran | ISBN 9780122208515

This equation is equal to [38] as the refining step uses a settler mixer as well. The capacity of the mixer ($CAP_{mixer,trans}$) also equals that of the refinement step. $N_{batch,trans}$ equals the number of batches based on 8,000 operational hours and 8 hours feeding time for the mixer settler ($FT_{trans}$) while $VESSEL_{trans}$ is the mixer settler volume calculated as:

$$VESSELtrans = FTtrans * \frac{(OILref/950)+((W+CHEM)/900)}{8{,}000} \quad [41],$$

in which 950 equals the density of a comparable triacylglyceride mixture, plant oil ($kg.m^{-3}$) and 900 represents the density of the mixture of water (25% of the incoming oil stream w/w) and chemicals (methanol, potassium hydroxide and sulphuric acid) required for the process. The water is used for washing biodiesel.

A final step subsequent to transesterification is included to remove the catalyst and any remaining impurities, necessary to comply with EN14105. This has been simplified to consider the main process interventions. At the end of the reaction the catalyst is neutralised with sulphuric acid. Phase partitioning and a water washing step is used to remove glycerol and salts present in the biodiesel. Separation of the biodiesel and aqueous phases is assumed to take place in a mixer settler unit. In larger scale biodiesel plants, centrifugal separation is used to “dry” the biodiesel, i.e. remove water from the FAME. This has a significantly higher energy consumption, therefore the mixer-settler unit is presented as a best-case option for the final purification step.

The amount of biodiesel produced (BIOD in litres per year) is calculated as follows:

$$BIOD = \frac{1{,}000}{880} * \frac{1}{0.9993} * OILref * 0.98 * 3 * 296.49 * \frac{1{,}000}{885.5} \quad [42]$$

in which:

- 880 ($kg/m^3$) is the density of biodiesel;
- 0.9993 is a parameter introduced to correct for losses
- 0.98 represents the loss during the transesterification phase (2 %);
- 296.49 represents the molecular mass (g/mol) of oleic acid (but the molecular mass of any other fatty acid or weighted average could be included here);
- the number 3 represents three moles of fatty acids from 1 mole of TAG;
- 885.5 represents the molar mass of vegetable oil (g/mol)

Besides biodiesel, this transesterification process also results in the production of glycerol (GLYC in kg/year), calculated as follows:

$$GLYC = 0.98 * OILref * \frac{92.09}{885.5} \quad [43],$$

in which the parameter 0.98 equals those in equation [42], 885.5 equals the molecular mass (g/mol) of vegetable oil and 92.9 that of glycerol.

The amount of methanol needed (MOH in kg per year) is estimated in the calculation process as follows:

$$MOH = 0.9 * (\frac{OILref}{885.5} * 6 * 32.04 * 0.5) \quad [44],$$

in which 885.5 is the molecular weight of vegetable oil and 32.04 that of methanol and the number 6 represents the molar ratio of methanol to oil as mentioned earlier. The number 0.5

represents 50% of the methanol not being used in the chemical process. The calculation also accounts for 10% losses of MOH during the recovery process and a 1% methanol loss in the biodiesel and the outgoing glycerol stream (not being shown in equation [44]). The transesterification process is assumed to require 157 hours of low labour and 12 hours of high value labour per year, assuming the base case of 10 tonnes of algae dry mass with 15% dm content in the paste.

The products from the biodiesel process accounted for in the economic calculations are:

1. *Biodiesel*
   From 10 tonnes of input biomass with 20% lipid, the expected output of biodiesel is 2,107 L. Prices for biodiesel vary according to the feedstock, and fluctuate depending on supply and demand at various times of year. The selling price for biodiesel as a substitute for diesel is estimated at 0.518 €/L (Voort *et al.*, 2014).

2. *Protein-rich biomass*
   The de-fatted biomass from sCO2 extraction could have a number of applications – as a source of pigments or other bioactive compounds or as a source of protein. Due to the biodiverse nature of algae, the model does not assume a specialty application for this biomass. The selling price for protein rich biomass as a substitute for soybeans is estimated based on the soy protein price of 350 € per tonne of soy bean meal with 50 % protein content on the dry matter and accounting for the protein content of the protein rich side stream produced. This assumes that the quality of the algae protein equals that of soybean and this has not been established yet. The protein content of the side stream could be 62% assuming 20% lipid content and 50% protein content in the algae. These numbers are used in the economics below. This represents a base case estimate, as at present algal biomass is a niche supplement for animal feed. If a move to large scale production (e.g. for biodiesel) was to happen, it is anticipated that the price of the algal material would decrease to reflect availability.

   Biomass with a residual amount of omega-3 lipids could be perceived as more valuable, with fishmeal is currently priced at € 1,000/tonne (Indexmundi – 65 % protein). Dry biomass would be compatible with conventional pelleting equipment for feed manufacture, making it theoretically, competitive with soy or other protein meals.

3. *Glycerol*
   Glycerol is a by-product of the transesterification reaction. The crude product may contain water, free fatty acids, and residual salts. Distillation may be carried out to produce a more pure product, and as such increase the value. Pure glycerol has applications in the pharmaceutical and personal care industries. A crude glycerol product may be used as an additive to anaerobic digestion to enhance biogas yield. For the model, subsequent refining of the glycerol is not considered, and a representative price € 0.5 /kg for crude glycerol is assumed.

***Table 13:*** *Estimated labour requirement for base case of 10 tonnes algae dry matter in hours per year.*

| Biodiesel Production | LQ labour | HQ labour |
|---|---|---|
| **Drying** | 400 | 12 |
| **Milling** | 400 | 12 |
| **$sCO_2$ extraction** | 391 | 12 |
| **Refining** | 85 | 12 |
| **Transesterification and purification** | 157 | 12 |

**Labour Costs**
From the base case of 10 tonnes of input biomass, the labour required was calculated. The number of potential batches was calculated, and a time assigned for LQ labour per batch. An overview of all labour requirements are listed in Table 13.

**Capital Equipment**
Estimated investment costs for the biodiesel downstream process are listed in Table 14. Dryer costs are estimated by using the e-commerce website Alibaba.com. The cost of a 10 L supercritical $CO_2$ rig is based on quotes from SciMed, who supply extraction equipment (http://www.scimed.co.uk/). Estimates for the refining equipment vessels and transesterification reaction are taken from heuristic values for tanks[7] as well as process information for soybean biodiesel production (Haas, McAloon, Yee, & Foglia, 2006). Process control costs are estimated at 10% of the total, apart from infrastructure. Infrastructure costs are estimated for a small simple structure or space in a larger process hall with piping connecting all equipment.

***Table 14:*** *Investment costs of capital goods for base case of 10 tonnes algae dry matter.*

| | Biodiesel route | |
|---|---|---|
| | **investment** | **life span (y)** |
| Dryer | € 5,000 | 20 |
| Ball Mill | € 10,000 | 15 |
| $SCO_2$ extractor | € 120,000 | 15 |
| Refining | € 15,000 | 20 |
| Transesterification equipment | € 22,520 | 20 |
| Process control | € 16,000 | 5 |
| Infrastructure | € 30,000 | 20 |
| **Total** | **€ 218,520** | |

**Scale up**
Brunner (2005) has estimated that the cost of operating a $sCO_2$ extraction decreases linearly, therefore inputs to the process have been scaled up proportionally. The cost of the physical unit operations have been scaled up using a factor of 0.6.

When scaling up $sCO_2$ extraction, one method to improve throughput is to have parallel feedstock reactors. This enables one to be loaded and prepared ready for a rapid changeover once the first batch has finished the pressurisation cycle. When that route has been maximised, another means of scaling up $sCO_2$ is to increase the vessel size. Industrial scale $sCO_2$ extraction is used for in decaffeination of coffee beans.

The main challenge for this process route is to scale the downstream process equipment accordingly. For 10 tonnes of algal biomass feedstock per year, the theoretical quantity of lipid is only 3 tonnes. Post $sCO_2$, this effectively means that mixer-settler tanks would not be operating at capacity.

[7] (Coulson & Richardson, 1996)

### 2.1.3 Economic results

The potential biodiesel yield from 10,000 kg algae biomass is 2,107 L according to the model. The income from diesel is only €982, while the assumed costs for the algae at this scale are €100,000. Although the rest products namely a protein-rich biomass and glycerol generate some income too, the economic results are negative, see Table 15.

***Table 15:*** *Return and costs ethanol production from 10,000 kg dm algae paste.*

| return | amount | | selling price | | return |
|---|---|---|---|---|---|
| Biodiesel | 2,107 | l | 0.52 | €/L | **€ 1,091** |
| | | | | | |
| **By-product** | | | **selling price** | | **by-product** |
| Protein-rich biomass | 8,100 | kg dm | 0.44 | €/kg | € 3,544 |
| Glycerol | 192 | kg | 0.50 | €/kg | € 96 |
| | | | | | |
| **Variable costs** | **amount** | | **price per unit** | | **total variable costs** |
| Algae biomass | 10,000 | kg dm | 10 | €/kg dm | € 100,000 |
| Electricity | 62,857 | kWh | 0.107 | €/kWh | € 6,726 |
| Chemicals | 37 | kg | 0.2 | €/kg | € 7 |
| Methanol | 408 | kg | 0.282 | €/kg | € 115 |
| Carbon dioxide | 6,253 | kg | 0.142 | €/kg | € 888 |
| Wastewater | 1 | $m^3$ | 0.1 | €/$m^3$ | € 0 |
| Labour LQ | 1,443 | hr | 11.53 | €/hr | € 16,632 |
| Labour HQ | 60 | hr | 25.57 | €/hr | € 1,534 |
| **Total** | | | | | **€ 125,903** |
| | | | | | |
| **Capital goods** | | | | | **total costs capital goods** |
| Depreciation | | | | | € 12,063 |
| Interest | | | | | € 6,584 |
| Maintenance | | | | | € 4,154 |
| Insurance | | | | | € 1,088 |
| **Total** | | | | | **€ 23,889** |
| | | | | | |
| **Total costs** | | | | | **€ 146,153** |
| | | | | | |
| **Total results** | | | | | **-€ 141,422** |

The cost price of one litre biodiesel from 10,000 kg algae dry mass is €69.38, while the selling price is only €0.52. 68% of the cost price covers algae biomass, but capital good costs (16%), labour (12%) and electricity costs (5%) are high as well. The major part of the capital good cost relates to the supercritical $CO_2$ extractor. Drying of algae paste takes a lot of energy.

***Table 16:*** *Biodiesel cost price structure (from 10 tonnes algae).*

| Cost | Amount | Percentage |
|---|---:|---:|
| Protein-rich biomass | -€ 1.68 | -2.4 % |
| Glycerol | -€ 0.05 | -0.1 % |
| Algae biomass | € 47.47 | 68.4 % |
| Capital goods | € 11.34 | 16.3 % |
| Labour | € 8.62 | 12.4 % |
| Electricity | € 3.19 | 4.6 % |
| Carbon dioxide | € 0.42 | 0.6 % |
| Methanol | € 0.05 | 0.1 % |
| Chemicals | € 0.00 | 0.0 % |
| Wastewater | € 0.00 | 0.0 % |
| **Total** | **€ 69.38** | **100 %** |

***Table 17:*** *Biodiesel cost price structure per process step (from 10 tonnes algae).*

| Cost | Amount | Percentage |
|---|---:|---:|
| Protein-rich biomass | -€ 1.68 | -2.4 % |
| Glycerol | -€ 0.05 | -0.1 % |
| Algae biomass | € 47.47 | 68.4 % |
| Drying | € 4.30 | 6.2 % |
| Dry ball milling | € 3.83 | 5.5 % |
| Supercritical CO2 | € 8.92 | 12.9 % |
| Refining | € 1.29 | 1.9 % |
| Transesterification | € 2.03 | 2.9 % |
| Process control | € 1.98 | 2.8 % |
| Infrastructure | € 1.29 | 1.9 % |
| **Total** | **€ 69.38** | **100 %** |

Downstream processing of larger algae biomass volumes will decrease the cost price for biodiesel.

On the basis of 1,500 tonnes algae biomass (assuming 20% lipid and 50% protein content) 316,000 litres biodiesel can be produced, see Table 18. Because the algae cost price will decrease from €10 to about €5 per kg (see chapter 1.5.5), the costs for algae will be halved and the cost price reduction for downstream processing in percentage will be even higher.

*Table 18: Upscaling effect on biodiesel production volume and cost price biodiesel in €/L diesel.*

| | | | | |
|---|---|---|---|---|
| **Algae** | **10 t** | (10 €/kg DM) | **1,500 t** | (5 €/kg DM) |
| **Biodiesel** | **2,107 l** | | **316,000 l** | |
| Algae biomass | € 47.47 | | € 23.37 | |
| Downstream processing | € 21.91 | | € 1.83 | |
| | **€ 69.38** | | **€ 25.56** | |

Obviously, a cost price of € 25.56 per litre diesel is still much too high in comparison with the selling price of € 0.52. Related to this price level, the downstream cost price of € 1.83 is also too high and the major contributors in this cost price are supercritical $CO_2$ extraction (€1.57) and drying (€0.58) and ball milling (€0.52).

## 2.2 Ethanol

### 2.2.1 Downstream processing of algae to ethanol

For the production of ethanol from algae, the first step is to disrupt the algal cell walls so that the cell's content is released. Several cell wall disruption technologies exist, and among these, ball milling is relatively common, which is why it is applied in this model. In case of dry milling, the resulting product is rehydrated. Then, the next step is enzymatic hydrolysis of the released polymeric carbohydrates. Of the resulting monomeric carbohydrates, only glucose is readily fermented into ethanol by common yeast. Some recently developed yeast strains also ferment other C6 sugars, and even C5 sugars, but less readily so and glucose is usually still preferred. Through distillation the ethanol can be upgraded from a concentration reached in the fermentation broth to 94%.

Protein is a (high-value) by-product of the process. The algae paste can be milled directly after harvesting (wet milling) or can be milled after a drying step (dry milling). The fact that wet milling makes drying unnecessary means that drying energy is saved, and the risk of damaging the cell's content – protein – is avoided. However, the energy requirements for wet milling are quite high as well, and the milling equipment is more expensive than a simple dry mill of comparable throughput. Moreover, as wet milling requires more or less immediate processing to avoid spoilage, it would also require a certain overcapacity of the entire downstream process, at least in regions with seasonal fluctuations in algae production. The drying step may be costly and potentially damaging to proteins present, but it would facilitate storage. Smoothing out production peaks in this manner could mean that a somewhat smaller – and cheaper – downstream processing process may be sufficient.

### 2.2.2 Explanation of the models

Two Down Stream Processing models have been developed. The main difference is that one uses dry milling to disrupt dried cells, while the other one uses wet milling, hereby avoiding a drying step (see 9). Starting point for the base case of both processes is 10 tonnes dry matter, in the form of harvested algae paste (15% dry matter; 66.7 tonnes). The operating time of the envisaged processing plant is 8,000 hours per year. It should be noted that the processing capacity required will depend mostly on the production peak, in summer (assuming a North West European algae production site). That is, unless the drying step facilitates storage, resulting in a consequently smaller required maximum processing capacity. In this model we ignore this aspect by stating that the algae processor is a different company than the algae producer.

**Table 19:** *Process steps for dry milling and wet milling.*

| | Dry milling | | Wet milling |
|---|---|---|---|
| 1 | Drying | | |
| 2 | Cell disruption by dry milling | 1 | Cell disruption by wet milling |
| 3 | Fermentation | 2 | Fermentation |
| 4 | Distillation | 3 | Distillation |

**Drying**

It is assumed that dry milling takes place at a dry matter content of 80% w/w. The required energy for evaporation at 100°C is 2,260 kJ per litre water to be evaporated. Electric heating is assumed to be used at 100% efficiency (1 MJ = 0.28 kWh), without heat recycling.

**Cell disruption by dry milling**

The capacity of the ball mill is based on 12.5 tonnes dried algae paste (10 tonnes or 80% w/w dry matter) and 8,000 operational hours. Electricity required by the ball mill is 1.87 kWh/kg dry biomass (Balasundaram et al, 2012), and the assumed cell disruption efficiency is 95%.

**Cell disruption by wet milling**

For the wet milling process, a ball mill type Dyno Mill AP is used for the model. A test performed by Dyno Mill in Basel with the AP05 mill and algae from the Lelystad open pond pilot plant resulted in +/- 95 % of the cells disrupted, after 2 passes (visually assessed). For the base case of 67.7 tonnes of paste (10 tonnes dry matter at 15% dm content), a Dyno Mill AP20 is assumed, which is driven by an 11 kW drive and requires 1,167 L cooling water per hour. This means 9,336 $m^3$ cooling water on a yearly basis, and consequently 9,366 $m^3$ waste water per year as well.

**Fermentation**

In case of dry milling, the water that was removed in the drying step is recycled back to the dried disrupted algae, resulting in the original 15% dry matter content. It would be possible to add less water, resulting in a higher dry matter content during fermentation, but because the consistency of the rehydrated paste is unknown, 15% dry matter is maintained.

For the base case of 10 tonnes dry matter, 4 fermentation batches of 50 kg dry matter and 6 hours each are run every 24 hours. Clearly, the carbohydrate content and composition of the used alga species has a large influence on the fermentation result. A number of assumptions and estimations are made for use in the model. Firstly, the alga *Porphydrium cruentum* has been described as having a carbohydrate content of 49% w/w in the dry matter of the growth medium. In practice, about a third of this can be present outside of the algal cell, as polysaccharides. But for the model, a 'best case' scenario of 49% w/w carbohydrates is maintained. Secondly, it is assumed that 65 % of the carbohydrates are enzymatically hydrolysed to monomeric glucose, becoming available for ethanol fermentation by yeast. Some yeasts may be capable to ferment xylose or arabinose to ethanol, but they do so less readily. Thirdly, the ethanol fermentation is assumed to be 100% efficient, meaning that all available glucose is actually fermented to ethanol. This means that the theoretical ethanol-from-glucose weight yield of 51% (w/w) is applied in the model. The model allows all of these parameters to be changed to study the effect on the end result.

The fermentation process can be described as follows:

$$C_6H_{12}O_6 \rightarrow 2\,C_2H_6O + 2\,CO_2 \qquad [45]$$

Accounting for molar mass of the molecules involved in this chemical equation, the involved weights would be: 180.16 kg of glucose would yield 92.14 kg of ethanol and 88.02 kg of carbon dioxide. So, per kg of glucose, the process would 0.51 kg of ethanol.

The enzyme dosage is 2 mL per kg wet paste, based on practical experience with a pilot bioethanol plant fed with maize. Of course, the enzyme used for algal cell wall carbohydrates will differ from the amylase used for starch hydrolysis. The applied yeast dosage is 0.35 g per L ethanol, again as used in the aforementioned bioethanol pilot plant.

All in all, and considering the rather positive assumptions, the maximum ethanol yield in this model resulting from 10 tonnes algal dry matter is 1,546 kg ethanol per year. The maximum ethanol concentration would be around 2.7% (w/w). If the carbohydrate content would be around 80 % (comparable to starch content in maize on dry matter basis) and if the algae paste would contain 20 % dry matter, the ethanol concentration would rise to almost 7% and this would obviously decrease the processing cost considerably. The ethanol concentration (ETH%) is calculated as follows:

$$ETH\% = ALGdry * RUPT\% * CH\% * CHferm * 0.51/WATER \qquad [46],$$

in which $ALG_{dry}$ equals the dry amount of algae processed (kg/year), RUPT% equals the percentage of cell rupture in the milling process, CH% equals the carbohydrate content in the dry matter, $CH_{ferm}$ equals the percentage of fermented sugars and WATER is the amount of water (kg per year) in the fermentation broth.

**Distillation**

For water removal from 2.7% to 94% (w/w) ethanol, 10 MJ energy is needed per kg ethanol, based on data from Vane (2008)). The energy required for distellation ($E_{distil}$ in kWh per kg of ethanol) is calculated using equation 47:

$$Edistil = 22.569 * ETH\%^{-0.817} * \frac{1}{3.6} \qquad [47]$$

The number 3.6 in equation [47] serves to convert MJ into kWh.

The step from 94 % pure ethanol to fuel grade (+/- 99 %) is not included in this model. The selling price for ethanol as a substitute for petroleum is estimated at 0.518 €/L (Voort et al. 2014), this is 0.409 €/kg.

**Protein**

Algal protein is an interesting by-product from the described process. Protein should preferably be extracted from the algae after the cell disruption step. As proteins denature at elevated temperature and hereby loose functionality, they become less valuable after a heating step. To estimate a protein value for the produced *Porphydrium cruentum*, only feed value is assumed. The price of the protein rich biomass is calculated according to paragraph 2.1.2.

**Labour and wastewater**

The estimated amounts of labour required are listed in Table 20.

**Table 20:** *Estimated labour requirement for base case of 10 tonnes algae dry matter in hours per year.*

| Dry milling | Labour low qualified | Labour high qualified | Wet milling | Labour low qualified | Labour high qualified |
|---|---|---|---|---|---|
| Drying | 400 | 12 | | | |
| Cell disruption by dry milling | 400 | 12 | Cell disruption by wet milling | 667 | 12 |
| Fermentation | 400 | 12 | Fermentation | 400 | 12 |
| Distillation | 400 | 12 | Distillation | 400 | 12 |

The amount of waste water produced annually is 57 $m^3$ per year assuming the base case of 10,000 kg of algal dry matter processed.

**Capital goods**

Estimated investment costs for both downstream processes are listed in Table 21.Dryer costs are estimated by using the e-commerce website Alibaba.com, while investments costs for the ball mill are based on information in Balasundaram *et al*., 2012. The wet ball mill costs are estimated using a quote for the Dyno Mill AP05 mill. Costs for the fermentation and distillation are both based on the investment costs for an ethanol pilot plant in Lelystad (see Figure 29) producing 150 thousand litres ethanol from corn per year, linearly scaled down to the base case of 10 tonnes dry matter algae. As the ethanol pilot plant distils to 50-60% (w/w) ethanol, costs for distillation to 94% (w/w) are roughly estimated.

Process control costs are estimated at 10% of the total, apart from infrastructure. Infrastructure costs are estimated for a small simple structure or space in a larger process hall with piping connecting all equipment.

**Table 21:** *Investment costs capital goods for base case of 10 tonnes algae dry matter*.

| | Dry milling | | Wet milling | |
|---|---|---|---|---|
| | Investment | Life span (y) | Investment | Life span (y) |
| Dryer | € 5,000 | 20 | | |
| Ball mill | € 10,000 | 15 | € 76,000 | 10 |
| Fermentation | € 6,532 | 20 | € 6,532 | 20 |
| Distillation | € 6,532 | 20 | € 6,532 | 20 |
| Process control | € 2,306 | 5 | € 8,906 | 5 |
| Infrastructure (building and piping) | € 30,000 | 20 | € 30,000 | 20 |
| **Total** | **€ 60,371** | | **€ 127,971** | |

*__Figure 29:__ Pilot of a bio-ethanol corn plant in Lelystad (Netherlands).*

**Scaling up; costs and economic calculations**
The upscaling equation applied in the models of the algae cultivation systems (equation [11]) is also used in the downstream process models and prices for water, electricity, waste water and labour are listed in Table 1. The cost price for algae biomass (15% dry matter is assumed to be €10 per kg dry mass. Based on the Lelystad bioethanol plant, enzymes cost are 6 €/L and yeast (DSM Fermiol) cost 9 €/kg.

### 2.2.3 Economic results

From 10,000 kg algae dry mass 1,546 kg ethanol can be produced according to the model and there is 3,230 kg protein available as a by-product of the process (based on the assumption that the algae dry matter contains 34 % protein). However the return from ethanol is only €632, while the costs for algae biomass are much higher, in this case €100,000. The possible income from the protein gives a small decrease of the costs, but the results are negative, see Table 22 and Table 23. Downstream processing of algae to ethanol by dry milling is cheaper than by wet milling.

***Table 22:*** *Return and costs ethanol production from 10,000 kg dm algae paste by wet milling.*

| **return** | **amount** | | **selling price** | | **return** |
|---|---|---|---|---|---|
| Ethanol | 1,546 | kg | 0.41 | €/kg | **€ 632** |
| **by-product** | | | **selling price** | | **by-product** |
| Protein | 3,230 | kg | 0.47 | €/kg | **€ 1,507** |
| | | | | | |
| **variable costs** | **amount** | | **price per unit** | | **total variable costs** |
| algae biomass | 10,000 | kg dm | 10 | €/kg dm | € 100,000 |
| Water use | 9,336 | $m^3$ | 0.878 | €/$m^3$ | € 8,197 |
| Electricity | 95,185 | kWh | 0.107 | €/kWh | € 10,185 |
| Enzyme | 133 | L | 6 | €/L | € 800 |
| Yeast | 1 | kg | 9 | €/kg | € 6 |
| Wastewater | 9,393 | $m^3$ | 0.1 | €/$m^3$ | € 939 |
| Labour LQ | 1,467 | hr | 11.53 | €/hr | € 16,911 |
| Labour HQ | 36 | hr | 25.57 | €/hr | € 921 |
| **Total** | | | | | **€ 137,958** |
| | | | | | |
| **capital good costs** | | | | | **€ 17,288** |
| | | | | | |
| **total costs** | | | | | **€ 153,478** |
| | | | | | |
| **Total results** | | | | | **-€ 151,600** |

***Table 23:*** *Return and costs ethanol production from 10,000 kg dm algae paste by dry milling.*

| **Return** | **Amount** | **Unit** | **Selling price** | | **Return** |
|---|---|---|---|---|---|
| Ethanol | 1,546 | kg | 0.41 | €/kg | **€ 632** |
| **By-product** | | | **Selling price** | | **By-product** |
| Protein | 3,230 | kg | 0.47 | €/kg | **€ 1,507** |
| | | | | | |
| **Variable costs** | **Amount** | | **Price per unit** | | **Total variable costs** |
| Algae biomass | 10,000 | kg dm | 10 | €/kg dm | € 100,000 |
| Water use | 0 | $m^3$ | 0.878 | €/$m^3$ | € 0 |
| Electricity | 59,889 | kWh | 0.107 | €/kWh | € 6,408 |
| Enzyme | 133 | L | 6 | €/L | € 800 |
| Yeast | 1 | kg | 9 | €/kg | € 6 |
| Wastewater | 57 | $m^3$ | 0.1 | €/$m^3$ | € 6 |
| Labour LQ | 1,600 | hr | 11.53 | €/hr | € 18,448 |
| Labour HQ | 48 | hr | 25.57 | €/hr | € 1,227 |
| **Total** | | | | | **€ 126,895** |
| | | | | | |
| **Capital good costs** | | | | | **€ 6,152** |
| | | | | | |
| **Total costs** | | | | | **€ 131,279** |
| | | | | | |
| **Total results** | | | | | **-€ 129,401** |

The ethanol selling price is € 0.41 per kg, while the cost price with dry milling is €85.08 and with wet milling €99.43 per kg ethanol. The cost for algae biomass (€64.68 per kg ethanol) is the largest contributor to the cost price. However, the cost price for the downstream processes alone already exceeds the selling price, see Table 24 and Table 25.

Costs for labour, electricity and capital goods are relatively high in comparison with the income from ethanol. Although no drying step is required for wet milling it is more expensive than dry milling because of higher capital good, water, electricity, process control and waste water costs.

***Table 24:*** *Ethanol cost price structure for dry and wet milling.*

| | **Dry milling** | | **Wet milling** | |
|---|---|---|---|---|
| Protein | -€ 0.97 | -1.1 % | -€ 0.97 | -1.1 % |
| Algae biomass | € 64.68 | 76.0 % | € 64.68 | 65.0 % |
| Labour | € 12.73 | 14.9 % | € 11.53 | 11.6 % |
| Electricity | € 4.14 | 4.9 % | € 6.59 | 6.6 % |
| Capital goods | € 3.98 | 4.7 % | € 11.18 | 11.2 % |
| Enzyme | € 0.52 | 0.6 % | € 0.52 | 0.5 % |
| Water use | € 0.00 | 0.0 % | € 5.30 | 5.3 % |
| Wastewater | € 0.00 | 0.0 % | € 0.61 | 0.6 % |
| Yeast | € 0.00 | 0.0 % | € 0.00 | 0.0 % |
| **Total** | **€ 85.08** | **100.0 %** | **€ 99.43** | **100.0 %** |

*Table 25: Ethanol cost price structure for dry and wet milling per process step.*

| | Dry milling | | Wet milling | |
|---|---|---|---|---|
| Protein | -€ 0.97 | -1.1 % | -€ 0.97 | -1.0 % |
| Algae biomass | € 64.68 | 76.0 % | € 64.68 | 65.0 % |
| Drying | € 5.86 | 6.9 % | € 0.00 | 0.0 % |
| Ball milling | € 5.22 | 6.1 % | € 24.30 | 24.5 % |
| Fermentation | € 4.29 | 5.0 % | € 4.29 | 4.3 % |
| Distillation | € 3.86 | 4.5 % | € 3.86 | 3.9 % |
| Infrastructure | € 1.75 | 2.1 % | € 1.75 | 1.8 % |
| Process control | € 0.40 | 0.5 % | € 1.53 | 1.5 % |
| **Total** | **€ 85.08** | **100.0 %** | **€ 99.43** | **100.0 %** |

Downstream processing of larger algae biomass volumes will decrease the cost price for bioethanol.

In case of 1,500 tonnes algae 231,924 kg bioethanol can be produced, see Table 26. Because the algae cost price will decrease from €10 to about €5 per kg (see chapter 1.5.5), the costs for algae will be halved and the cost price reduction for downstream processing in percentage will be even higher.

*Table 26: Upscaling effect on bioethanol production volume and cost price ethanol in €/kg ethanol (dry-milling).*

| | | | | |
|---|---|---|---|---|
| **algae** | **10 t** | (10 €/kg dm) | **1.500 t** | (5 €/kg dm) |
| **bioethanol** | **1,546 kg** | | **231,924 kg** | |
| algae biomass | € 64.68 | | € 32.34 | |
| downstream processing | € 20.40 | | € 2.61 | |
| | **€ 85.08** | | **€ 34.78** | |

Obviously, a cost price of €34.78 per kg bioethanol is still much too high in comparison to the selling price of €0.41 and even the downstream processing cost alone is very high. One of the reasons for this has already been explained on the basis of the low percentage of ethanol in the fermentation broth reached.

## 2.3 Methane

### 2.3.1 Co-digester with green gas unit

Digestion is a fermentation process in which organic matter is broken down by microorganisms to methane ($CH_4$) and $CO_2$ under anaerobic conditions. Co-fermentation is fermentation of various biomass flows in a fermentation installation (co-digester) at the same time, whereby biogas is produced. In practice, co-fermentation is often seen as manure being fermented with other organic material such as maize, barley, potatoes et cetera. These materials are added to increase the profitability of the fermentation process. In our calculations the digester is to be fed with algae paste only. The economic aspects concerning producing green gas were studied on the basis of an economic model in Excel. The model is based on a pilot co-digester (Figure 30) with a green gas unit in Lelystad in the Netherlands.


*Figure 30: Pilot of a co-digester at ACRRES in Lelystad (Netherlands).*

### 2.3.2 Explanation of the model

**Feed**

The co-digester can be fed with a mixture of manure and (one or more) other organic materials, with manure accounting for at least 50% of the total. The characteristics of the feed substrates are listed in Table .

**Table 27: Characteristics of some co-digester feed substrates.**

| Raw material | Dry matter | Organic dry matter | Ch4 | Price |
|---|---|---|---|---|
| | *%* | *% of dm* | *Nm$^3$/kg organic dry matter* | *€/tonne* |
| **Cattle slurry** | 8.0 % | 80.0 % | 0.170 | -€ 10 |
| **Algae paste** | 15.0 % | 95.0 % | 0.276 | € 750 |
| **Corn** | 32.0 % | 96.0 % | 0.375 | € 52 |

**Biogas production**

Biogas from a co-digester contains 52 % $CH_4$ and 48 % $CO_2$ by volume.

1 tonne of cattle slurry produces 8 % * 80% * 0.170 * 1,000 m$^3$ $CH_4$ = 11 Nm$^3$ $CH_4$ per year, equalling 21 Nm$^3$ biogas per year.

1 tonne of algae paste produces 15 % * 95 % * 0.276 * 1,000 m$^3$ $CH_4$ = 39 Nm$^3$ $CH_4$ per year. This represents 76 Nm$^3$ biogas per year.

The selling price for bio-methane as a substitute for LPG is estimated at 0.22 €/m$^3$ (Voort *et al.*, 2014).

**Digestate production**

The calculated mass of the produced amount of digestate is the total mass of the feed minus the mass of the produced biogas (methane and carbon dioxide). The density of methane is 0.668 kg/$m^3$ and for $CO_2$ 1.842 kg/$m^3$, so the calculated density for biogas is 1.232 kg/$m^3$. To market the produced digestate the costs are €20 per tonne, according to the algae model, see Table 1.

**Capital goods**

The investment costs are made scalable by use of a known investment level and the equation that allows estimating costs at different scale levels (equation [11]).

The minimum capacity of a co-digester is 1 million $m^3$ biogas per year. Typical investment costs for a co-digester tank are presented in Table 28. The cost of a unit that separates the $CO_2$ from the methane to produce green gas is also shown.

***Table 28:*** *Investment costs co-digester and green gas unit at different scale levels.*

| Capacity | Co-digester | Green gas unit |
|---|---|---|
| **Biogas (x1,000 $Nm^3$/year)** | **M€** | **M€** |
| 1,000 | 1.70 | 0.44 |
| 2,000 | 2.31 | 0.66 |
| 3,000 | 3.50 | 0.84 |
| 4,000 | 4.46 | 1.00 |
| 5,000 | 5.31 | 1.14 |
| 6,000 | 6.77 | 1.28 |
| 7,000 | 8.04 | 1.40 |
| 8,000 | 9.19 | 1.52 |
| 9,000 | 10.26 | 1.63 |
| 10,000 | 11.25 | 1.73 |

**Other cost assumptions**

The life span of the digester is estimated to be 25 years and 10 years for the green gas processer. Depreciation is dependent on the estimated lifetime for each item and based on a 10% resale value. The interest for capital goods is 5.5 % which averages 0.55*5.5% = 3% per year at a resale value of 10%. Insurance, maintenance and labour costs are respectively 0.5%, 2.0% and 2.5% of the investment. The land use for the smallest co-digester is 0.2 ha at 1,041 €/ha.

### 2.3.3 Economic results

To produce at least 1 million $m^3$ biogas per year (the minimum capacity of a co-digester), the digester has to be fed with 13,200 tonnes algae paste (15% dry matter). The methane (green gas) production is 520,000 $m^3$ per year.

The economic results are negative, according as shown in Table 29. With the previously described price assumptions costs do exceed income and there is no return on investment.

*Table 29: Return and costs methane production with algae paste.*

| **Return** | | | | |
|---|---|---|---|---|
| Green gas | 520,000 | $Nm^3$ | € 0.22 | **€ 114,400** |
| | | | | |
| **Variable costs** | | | | |
| Algae paste (15 % dry matter) | 13,221 | Tonne | € 750 | € 9,916,099 |
| Digestate | 11,990 | Tonne | -€ 20.00 | € 239,799 |
| | | | | |
| **Capital goods** | | | | |
| Depreciation | | | | € 100,360 |
| Insurance | | | | € 10,674 |
| Interest | | | | € 64,579 |
| Maintenance | | | | € 42,697 |
| | | | | |
| **Land** | 2,000 | $M^2$ | € 0.104 | € 208 |
| | | | | |
| **Labour** | | | | € 53,371 |
| | | | | |
| **Total costs** | | | | **€ 10,427,787** |
| | | | | |
| **Total results** | | | | **-€ 10,313,387** |

The cost price is € 20.05 per $Nm^3$ methane at a selling price of €0.22. The costs for the algae paste form the major part of the cost price, but even the algae downstream processing costs (€0.98 per $m^3$) alone are higher than the methane selling price.

*Table 30: Methane cost price structure.*

| Cost | Amount | Percentage |
|---|---|---|
| Algae paste | € 19.07 | 95.1 % |
| Digestate | € 0.46 | 2.3 % |
| Capital goods | € 0.42 | 2.1 % |
| Labour | € 0.10 | 0.5 % |
| Land | € 0.00 | 0.0 % |
| **Total** | **€ 20.05** | **100.0 %** |

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## Summary


Economic aspects concerning micro-algae production systems were studied on the basis of a bio-economic model in Excel named ‘AlgaEconomics’ which is developed within the EnAlgae project.

To produce energy products from algae further refinery steps are needed. Separate downstream processing Excel models for biodiesel, ethanol and methane are also developed within the project.


### The model ‘AlgaEconomics’

The model calculates the amount of algae biomass produced for each month of the year based on the global radiation at the location. All biomass produced is assumed to be harvested by means of direct centrifugation of the algae culture.

There are several easy accessible options in the models to study the effects of various parameters on the outcome. These options are:

- Scaling up: It can be assumed that certain cost will decrease when the production site would be scaled up to larger areas. An equation is included that allows the effect of up scaling to be calculated. This has an effect on capital good costs, labour requirement and extra land requirement for the surrounding equipment and infrastructure.
- Varying the location: The default location for the model is Lelystad, but changing to a different location changes the climate data set thereby changes the algae production level. The location also changes the day length used for the calculation of heat energy consumption.
- Seasonal operation: An option is given to stop the operation of the algae facility for part of the year. This was introduced with cold winter periods in mind when algae production might become limited and variable cost could outweigh the income from algae produced.
- Use of flue gas for $CO_2$ addition: Flue gas addition is considered optional for the open pond model. It is assumed that without $CO_2$ addition the algae growth is reduced or limited.
- Use of digestate as a nutrient source: If the option is not used, the model calculates the need for artificial fertiliser as separate nitrogen and phosphorus. If digestate is used, a small negative growth factor is introduced in the calculations as we assume that digestate increases turbidity thus decreasing light transmission and thus decrease growth.
- Application of heat: If the option is not selected, the temperature of the cultures is assumed identical to the average environmental temperature. Growth of the algae is influence by the ambient temperature regime. When heating is selected it is assumed that the ponds are heated to a temperature setpoint that is an input parameter located in the worksheet “parameters”. The setpoint temperature can vary per month. When heat use is selected the heat consumption is also calculated assuming that the ponds cool down at night and are heated to setpoint temperature during daytime. When there is no heat use the capital cost for heating equipment is ignored in the economic calculations.
- Varying the algae species: A maximum of three different algae species can be selected in the model. By changing the algae species the composition of the algae biomass changes which influences the algae production.
- Varying the prices: The assumed prices for in- and output can be changed to analyse the effects on the economics.

## Micro-algae production

Three types of micro-algae production systems were studied namely open ponds, tubular and flat panel photobioreactors (PBRs).

The yearly algae biomass production in a 1,000 $m^2$ open pond located in Lelystad (the Netherlands) is 1,538 kg of dry matter according to the model, equalling 15 tonnes dry matter per ha pond.

Biomass production per area in a tubular PBR is twice and in a flat panel more than three times as much as in an open pond, see Figure 31. Differences in production between production systems can be attributed fully to differences in photosynthetic efficiency (PE) on daylight (1.5% / 3% / 5%).

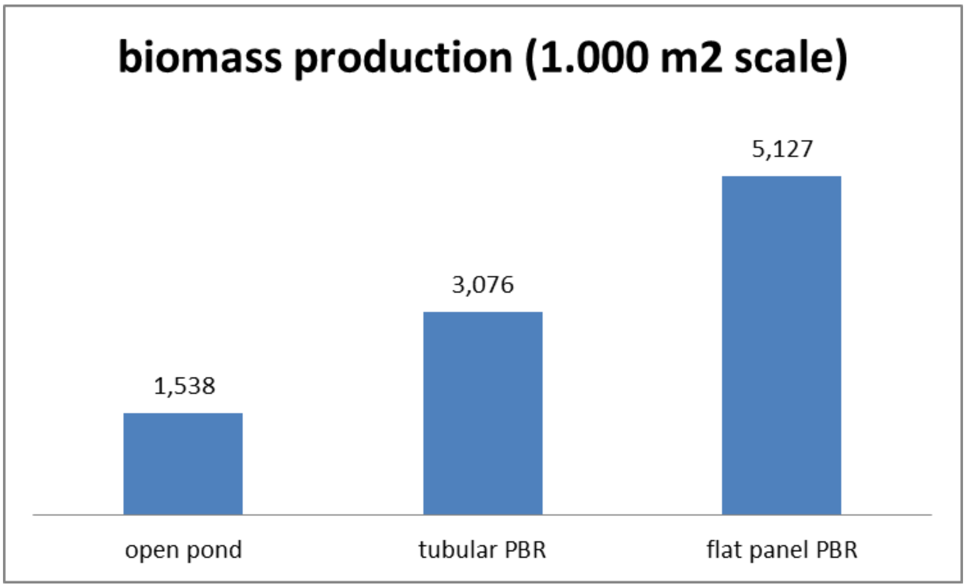


***Figure 31:*** *Biomass production for the three algae production systems (1,000 $m^2$ scale).*

In the Netherlands the average daily light intensity and temperature are low. When the model is fed by climatic data from Willemstad (Curacao) the yield is more than twice as high.

## Cost price micro-algae

The cost price for algae biomass produced on the basic scale (1,000 $m^2$) is much lower in PBRs than in open ponds, mainly because of the lower costs for capital goods and labour, see Figure 32.

Only electricity costs are higher with PBRs. A flat panel PBR has the lowest algae cost price.

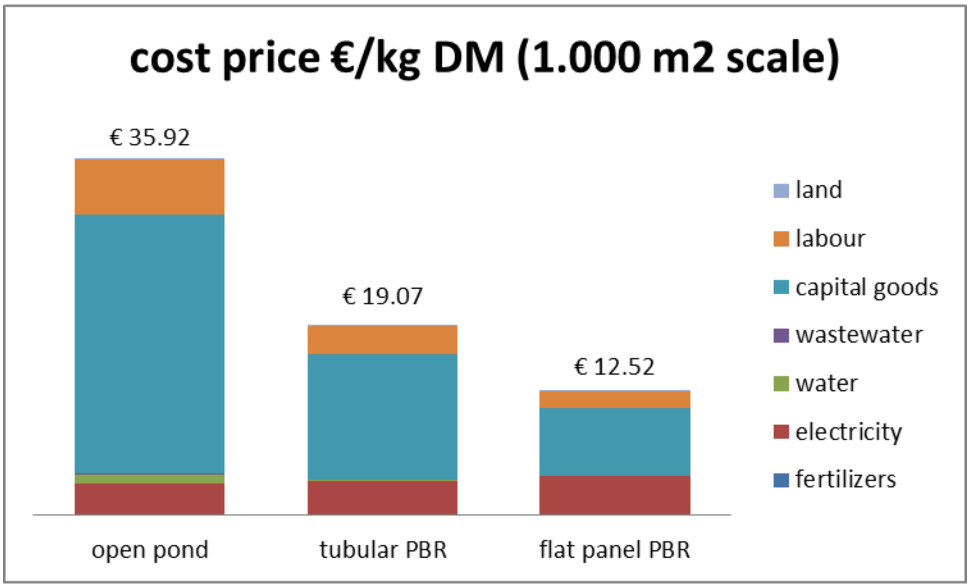


***Figure 32:*** *Algae cost price for the three algae production systems (1,000 $m^2$ scale).*

For the downstream processes to energy carriers very large algae production volumes are needed. To produce about 1,500 tonnes dry algae biomass (the minimum algae volume for a digester) an area of 100 ha open ponds, 50 ha tubular PBRs or 30 ha flat panel PBRs is needed.

At these higher production volumes, costs per kg biomass become more comparable for the three systems (about €5 per kg dry mass, see Table 31) where the main part consists of electricity costs.

***Table 31:*** *Land use and cost price for the production of 1,500 tonnes algae biomass for three algae production systems.*

| | open pond | tubular PBR | flat panel PBR |
|---|---|---|---|
| land use | 100 ha | 50 ha | 30 ha |
| cost price per kg | € 6.27 | € 4.88 | € 4.85 |

## Downstream processing to energy carriers

Separate downstream processing Excel models for biodiesel, ethanol and methane are also developed within the EnAlgae project.

### Diesel from algae

The downstream process of algae to diesel involves the recovery of intracellular lipid, and the subsequent conversion to the fatty acid methyl ester (FAME) via a transesterification reaction to make biodiesel.

Harvested algae must be dried and milled prior to supercritical carbon dioxide ($sCO_2$) extraction to separate lipid from the cells. Under conditions of high pressure and elevated temperature, $CO_2$ acts as solvent to remove neutral lipids, such as triacylglycerides. A pressure drop is used to collect various fractions from the biomass. In this example, only the lipid fraction is considered. Following $sCO_2$ extraction, the remaining protein-rich biomass can be sold for other purposes. The lipid is collected and reined prior to transesterification using methanol and a catalyst. Further purification is required to reduce the contaminants present in the FAME mixture to acceptable levels for EN14105 biodiesel. Glycerol, which is a by-product of the transesterification reaction, can be purified and sold.

### Ethanol from algae

First step is to disrupt the algal cell walls by ball milling so that the cell's content is released. The algae paste can be milled directly after harvesting (wet milling) or can be milled after a drying step (dry milling). The fact that wet milling makes drying unnecessary, means that drying energy is saved, and the risk of damaging the cell's content – protein – is avoided. However, the energy requirements for wet milling are quite high as well, and the milling equipment is more expensive than a simple dry mill of comparable throughput. In case of dry milling, the resulting product is rehydrated. The next step is enzymatic hydrolysis of the released polymeric carbohydrates. Of the resulting monomeric carbohydrates, only glucose is readily fermented into ethanol by common yeast. Through distillation the ethanol can be upgraded from a concentration of 2.7% to 94% ethanol. Protein is a (high-value) by-product of the process.

### Methane from algae

Digestion is an organic process where organic matter is broken down by microorganisms to methane ($CH_4$) and $CO_2$ under anaerobe conditions. Co-fermentation is fermentation of various biomass flows in a fermentation installation (co-digester) at the same time, whereby biogas is produced. In practice co-fermentation is often seen as manure being fermented with other organic material such as maize, barley, potatoes et cetera. These materials are added to increase the profitability of the fermentation process. In this case the co-digester will be fed with algae paste only.

The biodiesel and ethanol models are based on 10 tonnes algae dry matter, whereby the amount of biomass can be scaled up. Because the minimum algae volume to produce methane in a digester is about 1,500 tonnes algae, the comparison of the cost price for the energy carriers will be based on this large algae biomass volume.

## Production volume and cost price energy products from algae

In Table 32 the potential production volume of the energy products from 1.5 million kg algae are listed. The cost prices for the processing of these energy products from algae are very high and exceeds the selling price. (The cost price for the algae is assumed to be €5 per kg dry mass at this volume.)

***Table 32:*** *Production volume for biodiesel, bioethanol (dry milling) and methane from 1,500 tonnes algae biomass, cost price in comparison to the selling price.*

| | Biodiesel | Bioethanol | Methane |
|---|---|---|---|
| Production volume | 316,000 l | 231,924 kg | 520,000 $Nm^3$ |
| Cost price | 25.56 €/L | € 33.34 €/kg | 20.05 €/$Nm^3$ |
| Selling price | 0.52 €/L | 0.41 €/kg | 0.22 €/$Nm^3$ |

The raw material costs are the most important factor in the production price. But even if algae would be freely available the downstream processing cost still exceeds the selling price. If algae are to be used for energy production, an important decrease in the cost of production and processing is required.

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