



H2020-LC-SC3-2018-2019-2020
 EUROPEAN COMMISSION
 European Climate, Infrastructure and
 Environment Executive Agency (CINEA)

Biofuels production from Syngas
 Fermentation for Aviation and maritime
 use
 Grant Agreement No 884208

Deliverable D6.3

Process layout and cost engineering of the BioSFerA biorefinery plant

Document Details

Due date	31/12/2023
Actual delivery date	31/01/2024
Lead Contractor	SFW
Version	Final
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Document Details

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Abbreviations

BtL

DFBG

CAPEX

PSA

ATR

WP

TAGs

Biomass-to-liquid

Dual Fluidized Bed Gasification

Capital Expenditures

Pressure Swing Adsorption

Autothermal Reforming

Work Package

Triglycerides



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1 BioSferA concept

The BioSferA concept was initially introduced in Deliverable D2.5 'Full process basic definition' (Figure 1). The overall process can be separated in three distinct parts: the thermochemical part, the biotechnological part and the thermocatalytic part. The thermochemical part is based on the Dual Fluidized Bed Gasification (DFBG) unit, followed by a catalytic reformer and a proper syngas conditioning (contaminants reduction). The biotechnological part contains the double stage syngas fermentation (syngas → acetic acid → lipids) and the subsequent lipids purification system. Finally, the thermocatalytic part refers to the hydrotreatment activities that are applied for the conversion of the obtained lipids (TAGs) into drop-in liquid fuels.

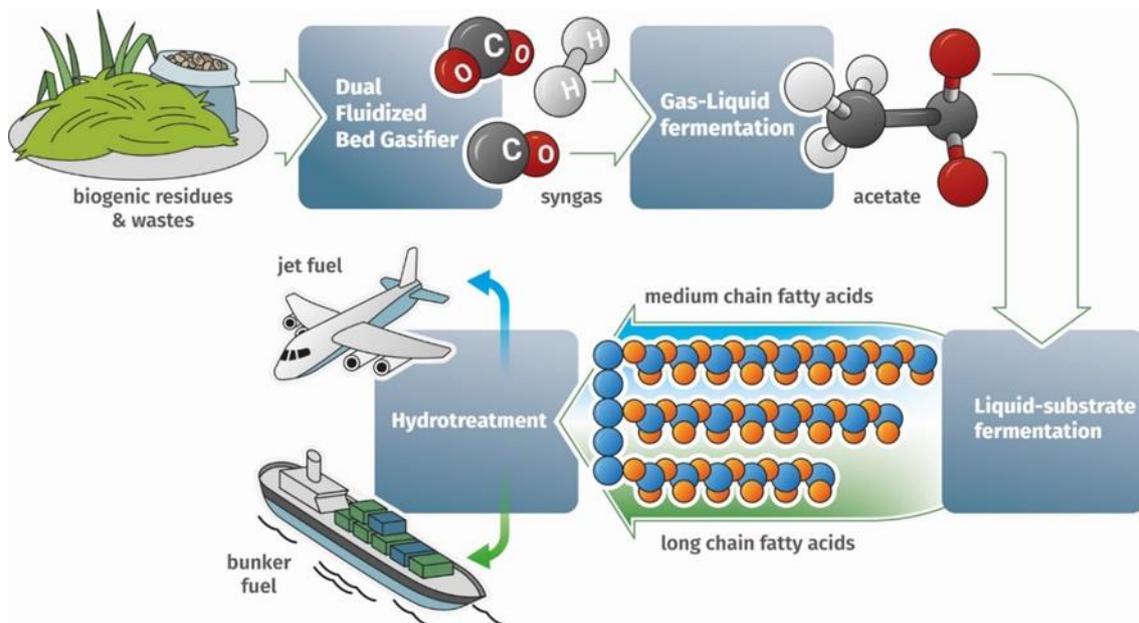


Figure 1. The BioSferA concept.

This report presents the industrial layout of the overall concept and provides a capital cost estimation for the whole BioSferA value chain.



2 Thermochemical part

A simplified flowsheet of the Biosfera project indirect gasification island is shown in Figure 2.

The first part of the report provides an overview of the scale-up study of a full-scale 2 x 100 MW / 200 MW gasification island for the Biosfera project.

During normal operation, the gasification plant will use 100% wood chips (crushed bark) as a feedstock. The feedstock is gasified in a CFB gasifier close to atmospheric pressure using steam as a fluidizing agent to form a syngas mixture containing mainly hydrogen (H₂), carbon monoxide (CO), carbon dioxide (CO₂) and hydrocarbons (C_xH_y). Limestone and sand are used as bed materials in the CFB cycle.

Downstream the gasifier, syngas is cooled down by a gas cooler acting as a boiler feed water preheater (or as an evaporation surface), filtered with the barrier filters, and cleaned out of tars with a dedicated catalytic tar reformer unit. After the catalytic tar removal reactor, syngas is cooled down with a product gas cooler benefiting from evaporation and superheating surfaces, and after that, syngas is flowing into a scrubber for further gas cooling and removal of its contaminants, such as ammonia and chlorides.

In this case the basic design of a combustor reactor is almost equal to the gasifier design. Combustor, cyclone and return leg to gasifier are also completely refractory lined structure without any heat transfer surfaces that would normally be part of the combustor furnace.

In the combustor, pre-heated air is acting as a fluidizing media and oxidizing agent for the combustion process. Air is distributed both through the grid nozzles and secondary/tertiary nozzles into the furnace.

The primary fuel in the combustor is the ungasified char derived from the gasifier, in conjunction with recirculated combustible gas. However, in instances where the heat balance necessitates, fresh biomass may also be employed as a supplementary fuel. The supplementary fuel is introduced into the lower part of the combustor, and the feeding arrangement closely resembles that of the gasifier. During normal operation all fresh bed make-up material is fed into the combustor.

Bottom ash from the combustor is extracted through the grid with similar arrangement to the gasifier bottom ash extraction system.

In this application, focusing on use of fresh biomass only, the back pass of the boiler follows a standard biomass design. This design incorporates components such as an air pre-heater, feed water pre-heater, steam evaporation, and superheating sections. Subsequently, the flue gas undergoes cleaning in a flue gas filter after passing through the back pass.

Furthermore, the indirect gasification plant lay out is presented in Figure 3.

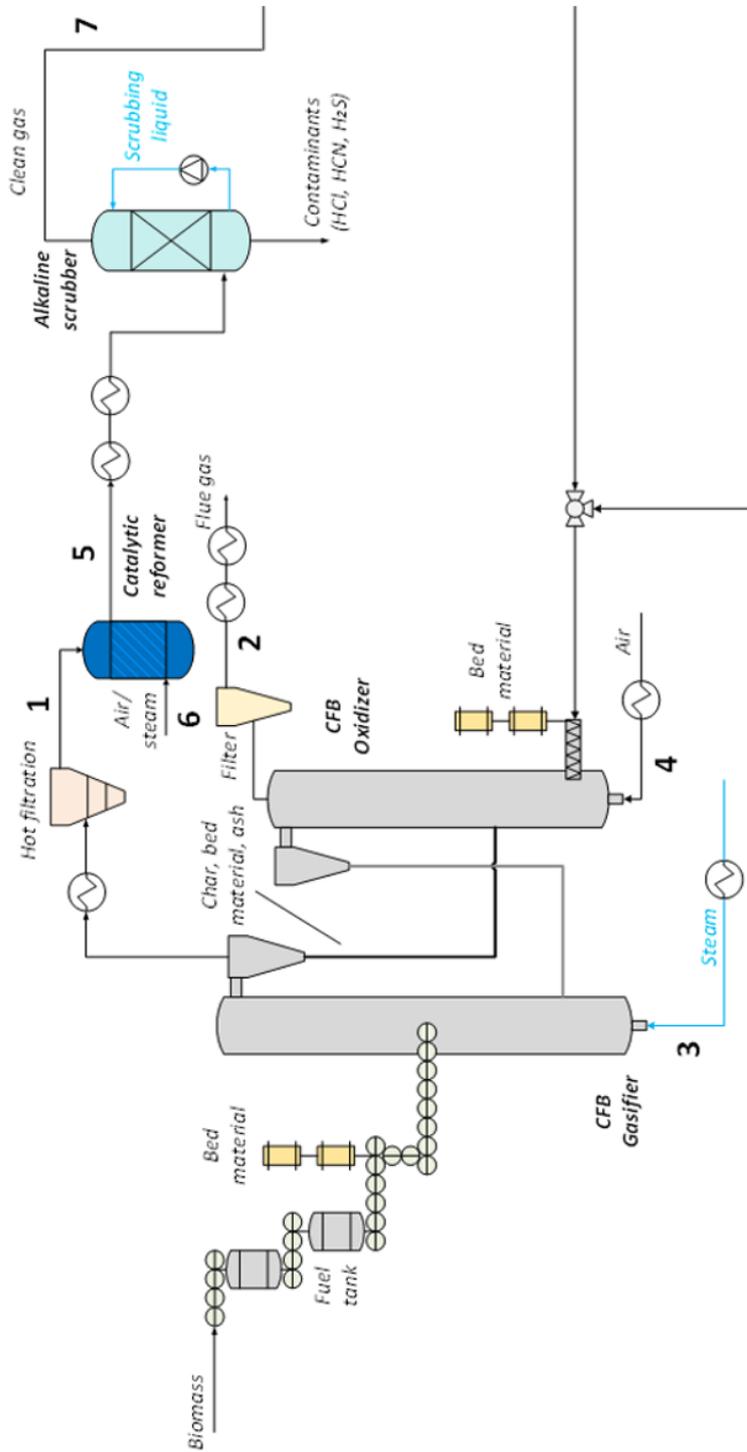


Figure 2. Simplified flowsheet of the complete Biosfera thermochemical part



2.1 Indirect gasification

The dual-fluidized bed gasification section consists of the following units:

Gasification section

- Biomass pre-treatment resulting in feedstock ready to be gasified and combusted (not included in this study)
- CFB gasifier with auxiliaries, based on close to atmospheric pressure circulating fluidised bed technology
- Product gas cooler
- Hot gas filtration system with auxiliary units
- Catalytic tar removal unit for reforming tars in the product gas
- Second gas cooler and scrubber for heat recovery and final removal of potential contaminants from the syngas, mainly ammonia, chlorides and possibly sulfuric components to some extent.

Combustion section

- CFB combustor with auxiliaries, based on close-to-atmospheric pressure circulating fluidised bed technology
 - The design of the combustor is different from normal boiler furnace since there are no heat transfer surfaces in the furnace, but it is all refractory lined
 - The CFB combustor is fluidized with air
- Normal boiler back pass with air preheater, water preheating, evaporating and superheating sections
- Flue gas filtration system with auxiliaries

Preliminary mass and energy balances of the whole Biosfera cycle were originally calculated by CERTH with an Aspen model and a reference biomass. Regarding the gasification island, the initial VTT piloting results underwent validation primarily through 3D modeling by LUT, and to a certain extent, through calculation models by SFW.

The following operating conditions were considered in this work:

Table 1. Example of biomass (wood chips) analysis used in Biosfera project. Several different biomass feedstocks were evaluated/discussed during project.

Item	Unit	Value
C	% in d.s.	50.80
H	% in d.s.	6.00
O	% in d.s.	41.79
S	% in d.s.	0.01
N	% in d.s.	0.40
Ash	% in d.s.	1.00
Moisture	%	15
LHV	MJ/kg in d.s.	19.50
LHV	MJ/kg as rec	16.20



Different operating conditions were tested and simulated in the calculations, but conditions described below were selected for the process scale-up.

- Biomass drying to 15% moisture content
- Gasifier and syngas temperature: 800 °C
- Combustor maximum temperature 920°C (can be lower depending on the plant heat balance)
 - Fuel and bed material agglomeration potential must be taken into account when determining the maximum temperature with different feedstocks
- Combustor inlet air after preheater: 100-150 °C
- Steam to dry ash free fuel ratio (S/F) of the gasifier is 0.7
 - Process calculations have been made also with S/F ratio of 1.0 which is close to S/C ratio used in VTT pilot tests

The main operating conditions are summarized in the following table.

Table 2. Summary of thermochemical part operating conditions.

Item	Unit	Value
Operating pressure	bara	1.3 - 1.5
Gasifier temperature	°C	780-800
Oxidizer temperature	°C	880-910
Steam-to-biomass ratio	kg/kgdaf	0.7 – 1.0
Steam pre-heating temperature	°C	150 / 350
Air pre-heating temperature	°C	150
Reformer (ATR) temperature	°C	900

The main process parameters based on CERTH simulations are presented in the following table.



Table 3. Main process parameters for 2x100 MW / 200 MW gasification island as per CERTH and SFW simulations (D 6.2).

Biomass feed to gasifier	Unit	1	2	3
Mass flow rate	kg/s	11,2		
Moisture content	%	15		
LHV	MJ/kg dry basis	21,37		
LHV	MJ/kg as rec	17,79		
Fuel effect	MW	200		
Fresh bed materials				
Sand	kg/s	0,2		
Limestone	kg/s	0,1		
Fluidizing media				
Steam	kg/s	7,9		
Syngas				
Mass flow rate	kg/s	17,0	24,2	19,2
Temperature	°C	800	900	15
Pressure	bara	1,4	1,1	1,1
H2	vol-%	30,9	32,6	41,5
CO	vol-%	12,8	15,3	20,3
CO2	vol-%	14,6	13,2	15,9
H2O	vol-%	32,0	22,7	1,7
N2	vol-%	2,0	15,4	19,6
H2S	ppm	178	127	21
CH4	vol-%	5,3	0,8	1,0
NH3	ppm	1900	273	84
HCN	ppm	12	0	0
COS	ppm	10	7	9
C2H4	vol-%	1,7	0	0
C6H6	ppm	3700	27	34
C10H8	vol-%	0,17	0	0
O2	vol-%	0	0	0
Ash and circulating material				
Bottom ash	kg/s	0,19		
Fly ash	kg/s	0,44		
1 = after gasifier				
2 = after reformer				
3 = after scrubber				



2.2 Plant lay out

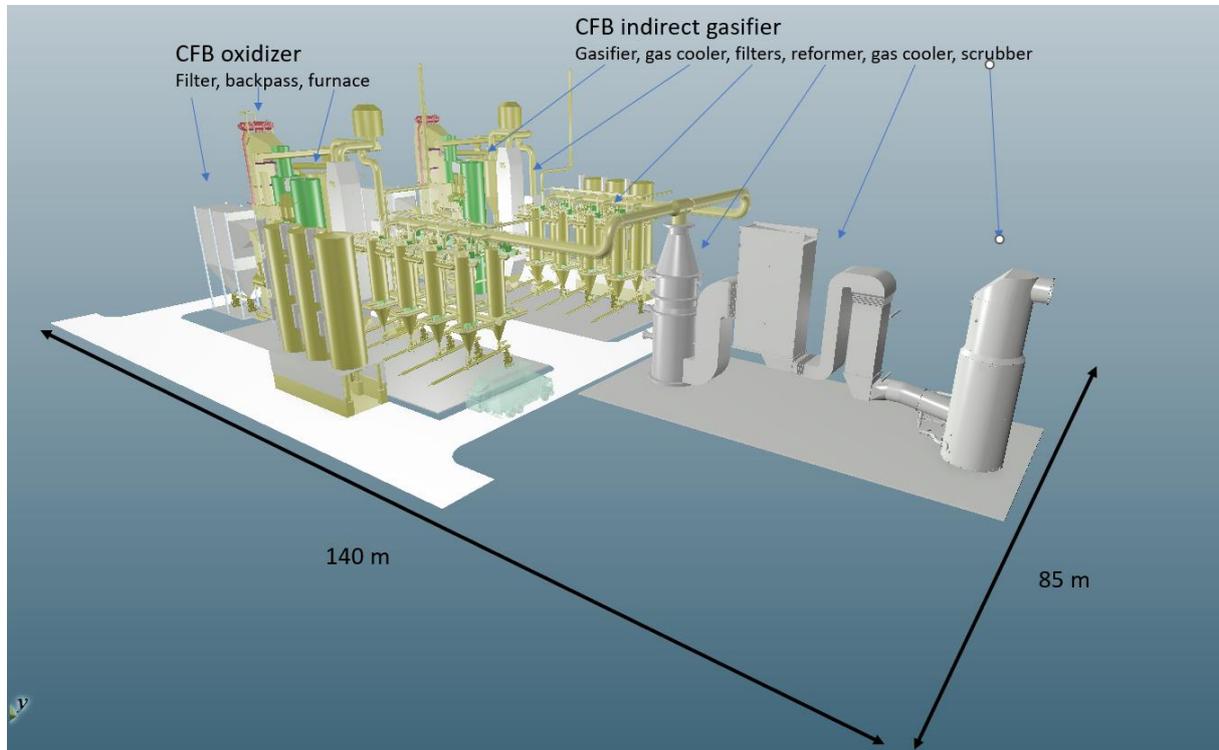


Figure 3. Plant lay out with main components. 2 x 100 MW gasification trains and a 1 x 200 MW reformer, gas cooler and scrubber.

As indicated in the figure above the footprint of the plant is 140 m x 85 m and the height of the steel structure or the building is approx. 65 m.

Various components of the plant are briefly presented in the following sections.



3 Biomass pre-treatment

Normally, biomass pre-treatment section consists among other things of the following stages (order and number of components may vary):

- crushing
- metal / impurity separation
- sieving
- drying

3.1 Biomass crushing

Biomass crushing is not included in the scope and cost estimate, but fuel is assumed to be in a proper particle size when received on site.

Maximum particle size should be limited at this point to 30-40 mm. Additionally, it is possible that the share of very fine material (< 3 mm) is beneficial to reduce in the feedstock for minimizing soot and tar formation and for maintaining carbon conversion at high enough level.

3.2 Biomass drying

Biomass drying is not included in the scope and cost estimate, but fuel is assumed to be dry enough when received on site.

In general, for the indirect gasification process with a synthesis process it is necessary to have relatively low moisture fuel fed into the gasifier since the gasifier is fluidized with steam. The target content for the moisture is approx. 15%. As biomass has typically higher moisture content (around 30-50%), a drying system for the feedstock is needed. There are different technologies available for biomass drying (drum driers, rotadisc driers, tube bundle driers, belt driers, etc).

For this commercial 2x100 MWth case, a tube bundle dryer could be considered to be the optimum solution. However, depending on the energy source available (water/steam, t, p), the final decision for the optimum technology should be made only after the water/steam balance of the whole cycle is evaluated, and the lay out and possible space requirements for the fuel receiving and handling station are considered. I.e. the decision is made and optimized case by case.

Dried biomass is stored in a dedicated storage system, including a conveyor from the dryer to the storage silo with discharging devices and a feeding conveyor transporting dried biomass into the daily storage silo located in each gasifier building.

3.3 Impurity separation and sieving

In general, fuel shall be free of non-fluidizing particles (stones, gravel, metals or similar). A separate limit value for allowed amount of non-fluidizable particles must be defined during the project phase.

For metal separation applications such as magnets and Eddy current systems are used. Suitable particle size distribution is set with fuel crushing and sieving devices. Crushing can be organized in different ways and can include several crushing stages. As regards sieving, disc screens are often used in biomass applications (especially wood based biomass).



4 Gasification

4.1 Indirect-CFB gasification process

The dual fluidized bed gasification process consists of two interconnected reactors: a steam-blown gasifier reactor (at 800°C) and an air-blown combustor reactor (at 910°C). Both reactors operate at a slight overpressure and gasifier pressure is a function of the plant design/scope (overall pressure difference) and the operating conditions, in practice the load of the unit. Combustor pressure is controlled separately and follows the gasifier pressure. Hot bed material consisting of Ca based sorbent (CaCO₃ / CaO), sand, fuel ash and char, is transported from the air-blown combustor to the gasifier, where it provides heat for the gasification reactions. Temperature difference between reactors affect the needed bed material flowrates between the reactors. In the gasifier, syngas travels upwards together with the fluidized solids comprising of ash, char and bed materials. The gas/solids mixture enters the cyclone and most of the solids are separated and directed back to the combustor through a return leg.

Syngas from the steam-blown gasifier consists of combustible gases (CO, H₂, C_xH_y), water vapour (H₂O), and carbon dioxide (CO₂). Since the gasifier is steam-blown, there is no airborne nitrogen in the gas, but all nitrogen based components are originating from the fuel nitrogen and nitrogen used as a sealing and purging gas in the system. Cooled bed material from the gasifier is transported back to the combustor to be re-heated. The remaining unreacted char in the circulating solids is combusted in the combustor together with recirculated gas in this very case. Additional biomass fuel can be fed to the combustor to maintain the required temperature level of the process if the energy from char is not enough.

As regards the gasification process itself, the feedstock is immediately dried and pyrolysed as it enters the reactor. The final composition of the syngas leaving the gasifier is a result of several homogeneous / heterogeneous reactions. At the moderate gasification temperatures of the CFB gasifiers, and depending on the fluidizing media and bed materials, also CH₄ content in the syngas is typically around 5-8% (dry basis).

Gasification of biomass at the moderate temperature conditions of the fluidised bed gasifier results also in olefin and tar content in the syngas exiting the gasifier. In fact, tar decomposition rate at temperature lower than 1200°C is not adequate to completely reduce the tar component in the syngas, thus requiring proper downstream removal treatment.

The following chapters describe the process equipment in more detail.

4.1.1 CFB gasifier

The CFB gasifier consists of the following sub-systems: CFB gasifier with cyclone and return leg to gasifier and combustor, start-up burner, solid fuel feeding system, fluidizing media (steam) system (including the equipment required for operation in air blown mode during the start-up phase) and bottom ash system. The gasifier is a refractory lined structure with steel shell. Also, the cyclone and the return leg are completely internally refractory lined



Steam is fed into the wind box and further through the grid into the reactor. The dried biomass is fed into the lower part of the gasifier. Bottom ash is removed from the reactor through the openings in the grid and furthermore through the cooled extraction screw into the bottom ash container. During the normal operation gasifier is operated as a steam-blown unit, but during the start-up phase, air is needed for the initial heating and for the start-up burner. Additionally, during the normal operation bed make-up is fed only into the combustor, but during the start-up phase it is assumed that part of the bed material is fed directly into the gasifier to speed up the start-up phase.

Under normal operation mode, the synthesis gas exits the gasifier through the cyclone at approximately 1.3 – 1.5 bar (absolute) and around 800°C, subsequently proceeding to the downstream cooling and filtration systems. As indicated earlier, the operating pressure depends on the final scope and design (filter pressure difference, tar removal concept, etc).

4.1.2 Biomass feeding system

The solid fuel feeding system inside the gasifier building consists of a daily silo, distributing conveyers, feeding lines and a feeding chute with (special design) rotary valve feeders leading into the lower part of the gasifier.

Since the gasifier is in the slight overpressure it is important to form a pressure lock so that combustible gases are not flowing upwards into the fuel feeding system or even through the whole fuel feeding system

4.1.2.1 Alternative feeding system considerations

Since the feedstock is solid wood based biomass (with relatively high bulk density), a lock hopper feeding system could be considered as well. This system is proven technology for feeding solid material into the pressurized systems. It is more costly and requires some more space, but an obvious alternative. A draw-back of the lock hopper system is extra inert gas consumption when pressurizing and de-pressurizing actual lock hopper vessels.

4.1.3 Bed material feeding system

Bed make-up, consisting of limestone and sand, is fed continuously into the combustor during the normal operation. However, as mentioned earlier during the start-up phase it is beneficial to feed part of the bed material also directly into the gasifier.

4.1.4 Fluidizing gas feeding system

Gasifier is assumed to be operated in an air blown mode during the gasifier pre-heating phase and during the **start-up phase** when increasing temperatures before turning the process into the steam blown gasification mode. As regards air temperature, it is beneficial to pre-heat air as much as possible.



During **normal operation** gasifier is fluidized with steam. Steam can be extracted from higher pressure and pressure is then reduced and adjusted in the steam pipe well before the gasifier.

4.1.5 Bottom ash extraction system

Bottom ash is extracted through the gasifier grid and ash chute and is cooled in a water-cooled extraction screw before being dropped into an ash container. Bottom ash is discharged on an intermittent basis, depending on the bed pressure, i.e. on the solids inventory at the bottom of the gasifier. Also, the solid impurities are extracted from the gasifier through the bottom ash extraction system. Gasifier bottom ash extraction rate is balanced with the combustor bottom ash extraction rate.

Also in here, as in fuel feeding system, if needed a lock hopper based system can be considered.

The level in the bottom ash container is controlled through an automation system. When the container reaches full capacity, it is emptied using a truck.

4.2 Raw syngas cooling and filtration

In practise the cooler arrangement, number of coolers and cooling effect depend on the selected tar removal system.

4.2.1 Syngas cooler I

Raw and dirty gas from the gasifier cyclone enters the syngas cooler which cools syngas down to the required temperature. The temperature level depends on the downstream filter and tar removal technology. In this case the gas temperature after the cooler is designed to be approx. 550°C. Syngas cooler I is called also a dirty gas or raw gas cooler as the syngas from the cyclone still contains impurities such as fuel ash, char and tars. Due to high impurity loading, cooler must be equipped with a suitable cleaning system removing deposits from the cooler heat transfer surfaces.

Ash from the syngas cooler is discharged on an intermittent basis, depending on the solid level in the cooler hopper, which acts as a buffer storage.

4.2.2 Hot gas filter

Depending on the tar removal process there can be different requirements for the particulate removal process efficiency. In this case the particulates are removed with a barrier filter system (full ceramic or metallic filter elements), which is almost an absolute filtration system, i.e. particulate concentration after the filter is extremely low.

A hot gas filter consists of a number of filter elements located in a filter casing divided by a tube plate into a raw gas side and a clean gas side.

Ash removed from the surface of the filter elements falls into the hopper part of the filter vessel and is discharged through the filter vessel bottom valve to a dedicated fly ash removal system.



Fly ash from the bottom of each filter vessel is discharged through a water-cooled fly ash extraction screw(s) to the downstream lock-hopper and finally to the fly-ash silo for final ash disposal. The water-cooled screw(s) cool the fly ash down to preferred temperature by a constant cooling water flowrate continuously entering the screw jacket and also the shaft, if needed. Continuous monitoring is maintained for the gas atmosphere within the fly ash silo, and in practice, it is kept inert through the use of N₂ purging.

4.3 Tar removal systems

At the moment there are several options for syngas tar removal process under development or under demonstration. Processes such as catalytic tar reforming or thermal reforming are interesting. For the clean biomass based concepts, feasibility of catalytic tar removal system has been demonstrated in a pilot scale and in a demonstration scale in other projects. Obviously, depending on the overall scope, the catalytic tar removal system could be one of the practicable solutions from the overall efficiency point of view.

4.3.1 Catalytic tar removal

For synthesis purposes, secondary cleaning measures are required to convert tar compounds and light hydrocarbons to synthesis gas. Reforming is widely studied and shown to be an efficient technology for gas cleaning. [3]

Steam reforming of hydrocarbons is a well-established industrial process for natural gas and naphtha feeds. Reforming of biomass gasification gas, however, has some special features compared to natural gas or naphtha reforming. The hydrocarbons that need to be reformed from biomass gasification gas are aromatic tar compounds and light olefins, such as ethene, which are more prone to carbon formation than typical steam reformer feedstocks. [3]

Another special feature is that the gas contains sulfur compounds. Sulfur is mostly in the form of H₂S (with biomass around 100 ppm) and a few ppm of COS. Sulfur compounds complicate reforming by poisoning the catalyst. However, studies show that catalyst activity can be retained by using reforming temperatures above 900 °C. [3]

Several types of catalysts have been studied for reforming hydrocarbons in biomass gasification gas. Typical catalysts are nickel and precious metal catalysts. Nickel catalysts have been traditionally used in industrial steam reforming of natural gas. Steam reforming nickel catalysts are active in tar and methane conversion and, in addition, have WGS activity and are effective in ammonia conversion. Their disadvantage is that they are prone to coke formation. Precious metals are less prone to this, but their price is high. [3]

**Table 4.** Gas composition before and after catalytic reformer by CERTH simulation (D6.2).

Syngas	Unit	Before reformer	After reformer
Mass flow rate	kg/s	16,99	24,19
Temperature	°C	800	900
Pressure	bara	1,37	1,1
H ₂	vol-%	30,89	32,61
CO	vol-%	12,79	15,33
CO ₂	vol-%	14,58	13,15
H ₂ O	vol-%	31,95	22,73
N ₂	vol-%	2,04	15,38
H ₂ S	ppm	178	127
CH ₄	vol-%	5,26	0,75
NH ₃	ppm	1900	273
HCN	ppm	12	0
COS	ppm	10	7
C ₂ H ₄	vol-%	1,74	0
C ₆ H ₆	ppm	3700	27
C ₁₀ H ₈	vol-%	0,17	0
O ₂	vol-%	0	0

4.4 Clean syngas cooling

The clean syngas cooling section consists of the following sub-systems: syngas cooler(s) and a syngas scrubber.

4.4.1 Syngas coolers

Before being sent to the scrubber, syngas is cooled down in one or two syngas coolers to the maximum temperature acceptable by the scrubber materials. This temperature can be in the range of 100-150°C. Syngas temperature exiting the final cooler is controlled to that range in order to feed syngas to the scrubber at the lowest temperature possible, provided that no water condensation occurs in the cooler itself and in the pipeline upstream the scrubber. This arrangement, as well as water/steam data of the syngas coolers depend to some extent on the selected tar removal process and the water/steam cycle of the plant.



4.5 Scrubber

The final cooling of the syngas, down to approximately 40°C or less, occurs in a scrubber, where a significant portion of the water in the syngas is condensed. The scrubber unit consists of a counter-flow fixed bed absorption column and water circulation system in two stages. The scrubber is able to separate both acidic and alkali components from the gas flow, since the lower part of the scrubber is operating in basic conditions, as the ammonia absorbed from the gas is keeping the water solution pH above neutral, and the top section is kept slightly acidic by adjusting the pH with formic acid. Should it be necessary, the scrubber is designed to capture residual tars that may pass through reformer in case of malfunction of tar reforming. Depending on the design and pH control, also NaOH can be used as an additive.

Depending on the pressure, the scrubber can be considered to be made of heat resistant plastic material (LMVEN) or coated from inside with it. This type of material enhances the chemical resistance especially for chlorine with respect to metal materials. However, LMVEN does not withstand high temperatures as effectively as metal materials. Consequently, the syngas inlet temperature is brought below 150°C in syngas cooler II. The excess water, condensed due to the cooling of the syngas, is sent by gravity to the wastewater treatment system.



5 Combustion

5.1 CFB combustion process

The CFB combustor consists of the reactor part, and its auxiliary component.

5.1.1 CFB Combustor reactor

In this case the basic design of the combustor reactor is almost equal to the gasifier design. Combustor, cyclone and return leg to combustor/gasifier are also completely refractory lined structure without any heat transfer surfaces that would normally be located in the combustor furnace.

Pre-heated air acts as a fluidizing media and oxidizing agent for the combustion process.

Combustor is operated in an air blown mode and in oxidising conditions all the time. During the start-up period air is distributed into three sections: primary air, secondary air and start-up burner air, while during normal operation only primary and secondary airs are in use. Primary air is fed into the combustor wind box whereas secondary air is distributed into the furnace to two or three different levels.

The ungasified char coming from the gasifier is the main fuel (with recirculated gas) in the combustor, but if the heat balance requires fresh biomass can be used as a secondary feedstock to balance the temperatures. Fresh biomass is fed into the bottom part of the combustor and feeding arrangement is almost equal to the gasifier feeding arrangement. Only one fuel feeding line is needed for a combustor of this size.

As discussed earlier in the gasifier section, bed make-up, consisting of limestone and sand, is fed during normal operation into the combustor only

The combustor bottom ash extraction system is equal to the gasifier bottom ash extraction system. Bottom ash is extracted through the combustor grid and ash chute and is cooled in a water-cooled extraction screw after which it is dropped into an ash container.

Alternatively, if proper pressure lock is not formed, a lock hopper system can be considered as an ash extraction system. In that case a lock hopper vessel would be located after the cooling screw. After purging with nitrogen, ash would be discharged from the hopper into the ash container.

The level in the bottom ash container is controlled via automation system and when the container is full it is emptied with a truck.

5.2 Backpass and flue gas cooling

In this commercial scale application, flue gas cooling can take place in a standard boiler backpass. Typically, this kind of a backpass consists of three sections: superheater section, economizer section for water pre-heating and air preheater section for preheating combustion air (primary air). In this very case also an evaporation surface is included since there is no such a surface in the combustor furnace. Backpass is equipped with necessary auxiliaries, e.g. soot blowing systems etc. The actual design and



heat transfer surface arrangement is considered always case by case and it depends on issues like fuel type, size of the plant and the steam/water data. Also, in this case the integration to the entire Biosfera process water/steam cycle must be considered carefully. The flue gas exit temperature from the air preheater depends on the fuel properties.

As regards the overall material selection there are no special features in this case. The selection and design can be made based on normal biomass boiler backpass design.

5.3 Flue gas cleaning

5.1 Filtration system

For the flue gas filtration system there are two options: A bag filter solution and an electrostatic precipitator system. The selection depends on the local circumstances and emission limits. From the particulate emission point of view the bag filter based solution is more efficient.

In the bag filters the filtration efficiency is typically very high and the operation principle is equal to the syngas filters discussed earlier. Filter is split to a dirty side and a clean side with a tube plate. In such a filter flue gas flows through the filter bag and ash particles of the gas are filtrated and form a “cake” on the surface of the filter bag. This cake increases both the filtration efficiency and the pressure difference of the filter bag. Thus, to ensure continuous operation, the filter bags must be cleaned on regular basis with a pulse gas (=air) cleaning system. A short-duration compressed air pulse causes a momentary expansion of the bag and loosens the cake from the bag surface. This cake or ash falls then into the hopper part of the filter house and from there it is transported via pneumatic transportation system into a separate fly ash silo. In addition to filter bags also so called light ceramic candles can be used as filter elements.

In electrostatic precipitation, suspended particles in the gas are electrically charged, then driven to collecting electrodes by an electrical field; the electrodes are rapped to cause the particles to drop into collecting hoppers. Effective separation of particles can be achieved with lower power consumption, with negligible draft loss and with little or no effect on the composition of the gas.

The combustor filter ash system is very much similar to that of gasifier filter ash handling system. Fly ash from each filter vessel hopper is discharged into the downstream lock-hopper and finally transported pneumatically into the fly-ash silo for final ash disposal. So, a traditional and a very straight forward design and solution which is widely used in the ash handling and transportation systems.

Filter ash from the gasifier and filter ash from the combustor are not mixed with each other, and have separate silos.



6 Consumables

The following utilities/consumables and service fluids, required for the gasification plant normal operation, are needed:

- Biomass feedstock
- Limestone
- Sand
- Nitrogen from storage / air separator;
- Cooling water;
- Raw/Service water;
- Gaseous / liquid fuel for start up burners
- Electric power

Steam, demineralised water and instrument air required for the plant operation are generated internally.

In cold conditions (temperature $<0^{\circ}\text{C}$) must be kept in mind to use glycol mixed water in the ash screw cooling circuits.

6.1 Inert gas balance

Both CO_2 and N_2 can act as inert gas on site. However, in this case only nitrogen is used.

Nitrogen can be transported to site as liquid and then evaporated for use. Other option would be to generate nitrogen e.g. with PSA or corresponding technology. Obviously, the purity of liquid nitrogen is better than PSA based nitrogen. This is a design, safety and OPEX issue to be taken into account when optimizing the inert gas supply.

The storage capacities and flow rates are to be defined separately based on the actual needs.

6.2 Equipment list of thermochemical part

The total cost estimate includes the following items:

Fuel feeding to gasifier and combustor

Gasifier and combustor + hot gas ducts including all refractories

Gasifier + combustor steel structure

Air system with auxiliaries

Steam and nitrogen systems including piping and valves

Start-up burners



Ash systems

Bed material systems

Flare and stack

Gas coolers + combustor backpass

Hot gas filters with auxiliaries

Catalytic tar reformer

Scrubber

Gas ducts

Instruments / EIC

Control and on/off valves

Local control boxes

Roof + steel structure

Engineering, erection and project management etc.

Commissioning

Spare parts for two years



7 Biotechnological part

The following chapters provide a summary of the process description and basis for an estimation of the direct and total costs for the biotechnological part of the 200 MWth full-scale BioSFERA Biomass-to-Liquid (BtL) process chain concept. Process is introduced in more detail in Deliverable D2.5 (Full process basic definition), and further optimized in Deliverable D6.2 (Results of full-chain process simulations).

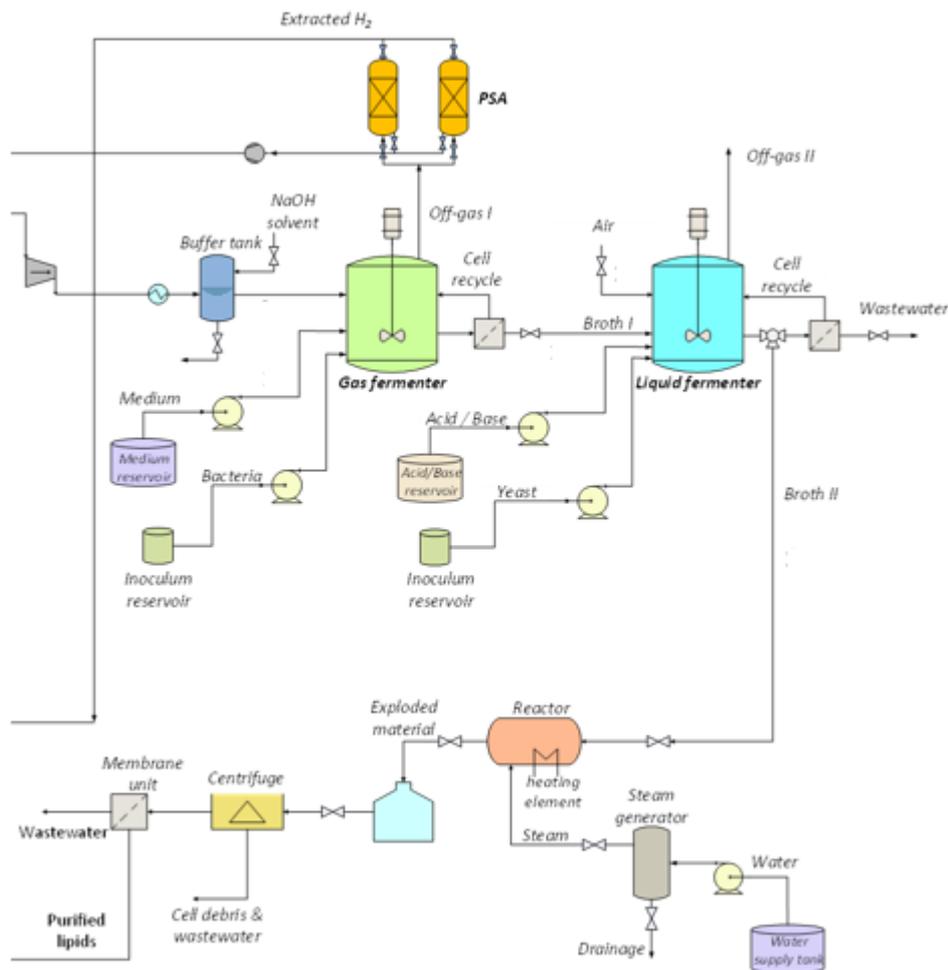


Figure 4. Biotechnological part of the BioSFERA value chain.

7.1 Process description

In the first step of the biotechnological part of the process, the interaction of syngas with the acetogenic bacteria under anaerobic conditions leads to acetic acid production. For the syngas fermentation stage, after the extended experimental testing, *Moorella thermoacetica* was selected as the most efficient



acetate producer strain. The operating temperature is set around 55 °C, since the optimal temperature range for these strains is 55-60 °C. The operating pressure of the reactor was considered to be 5 bar in order to achieve higher solubility of the reacting gases. Two critical factors, that highly influence the fermentation kinetics and consequently the acetate productivity, are the gas solubility and the ratios of CO₂/CO/H₂. The unconverted syngas components (off-gas) can be either recycled back to the fermenter or utilized elsewhere in the plant. The broth containing the produced acetate in low concentration is extracted in continuous way, and the liquid volume is kept constant by adding fresh culture medium. A cell recycling system (hollow fiber membrane) is required to keep the cells in the fermenter while extracting the liquid effluent.

The second fermentation step refers to the production of TAGs via aerobic fermentation of the diluted acetic acid stream. Taking into account the experimental trials of WP3, *Yarrowia lipolytica* is the yeast strain that has been selected to be involved in the liquid substrate fermentation of acetate. The diluted acetate effluent stream from the syngas fermentation enters the aerobic fermenter, where the targeted TAGs are formed as intracellular products in the presence of oxygen, additional nutrients, salts and the oleaginous yeast (*Y. lipolytica*). A cell recycle system (hollow fiber membrane) can be installed to recirculate the cellular biomass in the bioreactor while extracting the effluent. At the same time, a gaseous CO₂-rich stream is produced and leaves the bioreactor from the top.

Lipids extraction from the oleaginous yeasts is an important step before hydrotreatment. As oleaginous yeasts store lipids in intracellular forms, extraction is required to obtain TAGs. Cell disruptions alongside lipid extraction steps are critical for large-scale biofuel production in terms of cost adequacy. Mechanical disruption requires energy inputs such as shear forces, electrical pulses, waves or heat. Mechanical processes generally provide high products recovery yields with good management and scalability, but they are energy intensive. Among the options actually available, there are novel technologies with considerably lower power consumptions such as steam explosion, centrifugation and membrane separation considering different process parameters and extraction procedures. Within BioSFERA project, based on the insights gained in Deliverable D3.6 'Lab scale downstream processing for TAGs recovery and purification using conventional and novel strategies', a scalable DSP (downstream processing) train was defined for the efficient lipids recovery from the fermentation broth.

Steam explosion is an innovative method with reduced environmental impact, lower costs and energy demand, compared to other techniques that are widely used. In steam explosion, raw material exposed to steam at 180-240 °C for several minutes and then subjected to depressurization to ambient conditions. This generates an explosion that causes cell-wall disruption. In context in which heat flows are available as downstream of other processes, and so steam could be generated at low cost (such as the BioSFERA concept), steam explosion should be considered as potential technology for cellular biomass fractionation with high yields of recovery. The relevant experimental activities of WP3 revealed that steam explosion should be performed at low pressure and temperature (about 5 bar and 150 °C) in order to avoid the TAGs disruption. Microfiltration/Centrifugation have been positively evaluated for their ability to separate oil from the broth deriving from steam explosion. The solvent extraction step should be further investigated because of the difficulties associated with the formation of emulsions.

The choice between using the high-pressure homogenization or steam explosion for the cell-breaking stage for future scale up of the DSP process should be made following some technical and economic considerations. Before solvent extraction, a microfiltration step can be included to remove the cell debris.



The main operating conditions for the biotechnological part at industrial scale, as derived from the up to now experimental/pilot activities of BBEPP, CARTIF, CSIC, and ENVIPARK are presented in

Table 5.

Table 5. Main operating conditions for the biotechnological part of the BioSferA concept

Parameter	Input
Gas Fermentation Pressure (bar)	5
Gas Fermentation Temperature (°C)	55
Liquid Fermentation Pressure (bar)	1
Liquid Fermentation Temperature (°C)	28
Steam Pressure for Steam Explosion (bar)	5
Steam Temperature for Steam Explosion (°C)	150

7.2 Equipment list of biotechnological part

The biotechnological part is further divided into 3 main steps: gas fermentation, aerobic fermentation, and TAGs recovery. For the definition of the main equipment, Deliverable 6.2 was used as a baseline. The main equipment is described in Paragraph 3.4 and Figure 7 (Process Flow Diagram (PFD) for the BioSferA BtL process chain), while for the number of upstream (reactors) and downstream equipment, we considered the outcome of the sensitivity analysis included in paragraph 4 (Optimization of model parameters and model adaptations), and the Heat & Mass balances from paragraph 5. For example, for the gas fermentation, we considered an Acetic acid productivity of 3 g/L/h (at a fixed Acetic Acid Titer of 30 g/L). For the TAGs fermentation, we considered a TAG productivity of 0.8 g/L/h (at a constant 30 g/L of Acetic acid in the feed).

According to the above, the number of bioreactors for gas fermentation is 30, and for liquid fermentation is 30. For both steps, the same bioreactor type and geometry were used. A continuous stirred tank reactor is the most common type of bioreactor, in this kind of reactor gas solubilization is favored by mechanical agitation.

As described in the PFD (Figure 7, Deliverable 6.2), the downstream of the TAGs consists of several steps among them there are microfiltration, solvent extraction centrifugation, and evaporation. After the first microfiltration step, cells have to be broken to recover the bio-oil from the inside of the cell. There are different techniques available to break cells. In this report, we decided to use high-pressure homogenization instead of steam explosion. For the selection of each downstream equipment, the results of the mass balances (Paragraph 5.1, Deliverable 6.2) were used.

Additionally, to the main equipment, side equipment was included. Here below the side equipment is listed out:

- Heating/Cooling unit to control the temperature of the bioreactors.
- Inoculum reactor and transfer pump
- Steam generator (for bioreactors and lines sterilization)
- Acid/base reservoir (for pH control) and transfer pump



The list of the equipment with the dimensions of each system can be found in the equipment list. The list is divided according to the main steps of the biotechnological part: Acetic acid fermentation, TAGs fermentation, and the subsequent lipids purification system. Media preparation, and seeds (AA, and TAG) are also included as they are an essential part of the process.

EQUIPMENT LIST OF BIOLOGICAL PART

Item	Quantity	Name	Description	Operating Volume	Dimension	Remark
<u>Media preparation</u>						
	30	Substrate/Powder handling	Storage			
	30	Chemical Dosing	dosing NaOH and H2SO4 for pH-correction, Antifoam			
	30	Substrate preparation:	dissolving, sterilisation			
	30	Substrate dosing system				
<u>Acetic Acid fermentation</u>						
	30	Main fermenter	CSTR	250 m ³	D=4.7 m, H=14 m	H/D=3
	30	Agitator				
	30	Circulation Pump				
<u>Seed AA</u>						
	3	Main fermenter	CSTR	250 m ³	D=4.7 m, H=14 m	H/D=3
	3	Agitator				
	3	Circulation Pump				
<u>TAG fermentation</u>						
	30	Main fermenter	CSTR	250 m ³	D=4.7 m, H=14 m	H/D=3
	30	Agitator				
	30	Circulation Pump				
<u>Seed TAG</u>						
	6	Main fermenter	CSTR	15 m ³		H/D=3
	6	Agitator				
	6	Circulation Pump				
<u>Utility</u>						
	6	Steam generator				
	6	Heating cooling unit				
	6	acid/bast tanks				
<u>TAG separation</u>						
	1	Microfiltration	feed flow max 50 kg/s		W=7 m, L=3 m, H=5 m	
	3	High pressure homogenization	feed flow max 10 kg/s		W=2 m, L=3 m, H=2 m	
	1	Microfiltration	feed flow mas 50 kg/s		W=7 m, L=3 m, H=5 m	
	1	Solvent extraction	NA		NA	
	3	Centrifugation	feed flow max 20 kg/s		W=2 m, L=2 m, H=2 m	
	1	Evaporation	NA		NA	

In general, costs can be divided among the steps of the biotechnological part as follow: 50% gas fermentation, 40% liquid fermentation, 10% TAGs recovery. To build up the total equipment costs, prices of the key equipment were collected or estimated based on in-house information, literature, and other projects. The cost of piping, valves, accessories, and civil works have been calculated according to ratios coming from other projects.



7.2.1 Influence of acetic acid & TAG productivity on cost estimation

The cost of the biological part estimated in the paragraph above take into account Acetic acid and TAG productivities estimated from the sensitivity analysis from the chapter 4.4 deliverable 6.4. In the present report, we considered Acetic acid productivity of 3 g/L/h (at a fixed Acetic Acid Titer of 30 g/L) and TAG productivity of 0.8 g/L/h (at a constant 30 g/L of Acetic acid in the feed).

In a view of a process intensification of the biotechnological part of BioSferA concept, the table below shows the cost estimation at higher productivities. Since in the BioSferA concept gas fermentation and liquid fermentation are in series, to simplify the cost estimation, it was decided to consider the same number of bioreactors for both steps.

AA		TAG		Cost estimation
Productivity	Number of reactors	Productivity	Number of reactors	
g/l/h		g/l/h		%
3	30	0.8	30	100
4.5	20	1.1	20	67
6	15	1.4	15	53

As for before, the cost estimation includes only the direct costs (equipment, piping, valves and accessories, civil works).



8 Thermocatalytic part

The final section (thermocatalytic part) of the BioSFerA value chain includes the upgrading of microbial oil into drop-in aviation and marine biofuel. The thermocatalytic part refers to the hydrotreatment unit where the consecutive hydrogenation, deoxygenation, isomerization and fractionation procedures of the purified TAGs take place.

8.1 Process description

The final section of the BioSFerA value chain includes the upgrading of microbial oil into drop-in aviation and marine biofuel. The core of the thermocatalytic part of the concept is the hydrotreatment unit where the consecutive hydrogenation, deoxygenation, isomerization and fractionation procedures of the purified TAGs take place. Common catalysts for this process are Pt, Ni or other metals based on Al_2O_3 .

In particular, the saturated fatty acids are converted to straight long-chain alkanes by hydrodeoxygenation and decarboxylation, co-producing propane, methane, water, CO, and CO₂. The deoxygenated straight chain paraffins are selectively hydrocracked or isomerized yielding highly branched alkanes. The resulted organic product is a mixture of straight and branched C_nH_{2n+2} that can be suitably used as drop-in liquid fuel. Fractionation is necessary to separate the jet from marine fraction.

The envisaged configuration for the hydrotreatment unit is presented in Figure 5.

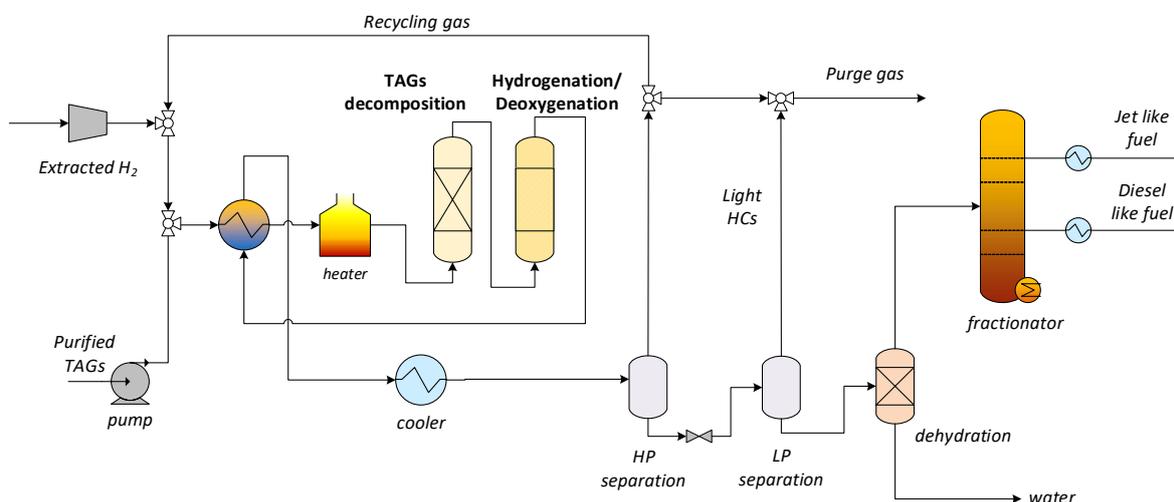


Figure 5. Hydrotreatment of the BioSFerA value chain.

The main operating conditions for the thermocatalytic part at industrial scale, as derived from the up to now experimental/pilot activities of CERTH and KPRT are presented in Table .

**Table 6.** Main operating conditions for the thermocatalytic part of the BioSferA concept

Parameter	Input
Reactor pressure (bar)	100
Reactor temperature (°C)	370
Hydrogen-to-TAGs ratio (kg/kg)	0.05

8.2 Equipment list of thermocatalytic part

This is the most technologically mature part of the process taking into account not only the vast expertise and equipment provided by conventional refineries but also the large number of recent investments in HVO plants. It would be quite accurate to claim that the thermocatalytic part of the BioSferA value chain is in essence a HVO plant with microbial oil as feedstock instead of other typical oils. Therefore, taking into advantage the availability of such plants at commercial level and the proven similarity of the BioSferA microbial oil with other typical feed oils (WP5), the already available and tested industrial layout of these plants is considered for the hydrotreatment part of the BioSferA process as well (see PFD). The boundary conditions that could serve as design guidelines for the industrial setup of the BioSferA hydrotreatment unit are provided in Table 77, as they emerged from the simulated 200 MWth Heat & Mass balances and the insight from the latest pilot hydrotreatment tests. Finally, it has to be noticed that in the case of BioSferA concept, as analyzed in Deliverable D6.2 'Results of full-chain process simulations', the required hydrogen for the hydrotreatment procedure is secured via PSA (Pressure Swing Adsorption) from the off-gases of the gas fermentation process. The main equipment list is presented in



Table 88.

Table 7. Design guidelines for the industrial setup of the BioSferA hydrotreatment unit

Parameter	Value
TAGs feed flow	1.82 kg/s
TAGs feed volume	7.14 m ³ /h
Catalyst volume (for LHSV = 1 hr⁻¹)	7.14 m ³
Fresh hydrogen flow	0.10 kg/s
Hydrogen consumed (per pass)	0.06 kg/s

**Table 8.** Main equipment list for the industrial setup of the BioSferA hydrotreatment unit

Component	Quantity
Hydrotreating reactor	2
Fractionating column	1
Flash tank (separator)	2
Dehydrator	1
Compressor (for hydrogen)	1
Liquid pump (for TAGs)	1
Reactor feed-effluent heat exchanger	1
Reactor feed heater	1
Reactor effluent air-blown cooler	1
PSA hydrogen unit	1

The capital cost for the BioSferA hydrotreatment unit was estimated in an aggregated form as a function of the annual capacity (~40 kt/year liquid products, new HVO-type of refinery, see Deliverable D6.2) by utilizing available technical reports [1,2]. The presence of the hydrotreatment unit in an integrated BtL scheme (i.e. BioSferA concept) comes up with design benefits compared to standalone refineries, such as:

- No need of ‘heavy’ pre-treatment of the feed (TAGs), since appropriate cleaning has been carried out in earlier stages of the value chain
- No need of hydrogen production unit, since the required hydrogen is extracted internally from the off-gases of the biological part (via PSA)

Thus, there are promising factors for further capital cost reductions. The claimed capital costs for the potential large-scale hydrotreatment unit will also be adopted for the forthcoming Task 7.1 ‘Techno-economic assessment’.

9 BioSferA value chain total cost estimate

As discussed the BioSferA process concept consists of 3 main steps:

- Thermochemical: biomass gasification in syngas,
- Biotechnological: syngas conversion in TAGs and subsequent TAGs recovery and purification,
- Thermocatalytic: TAGs upgrading into aviation and marine biofuels.

The total cost estimate (TCI, Total Capital Investment) for the establishment of a 200 MWth BioSferA commercial plant is approx. 570 M€ (plus-minus 40%). A more detailed cost breakdown for each section will be included in Deliverable D7.1 ‘BioSferA Techno-Economical Assessment’ along with the assessment of the financial competitiveness of the concept



10 References

- [1] Concawe. *Refinery 2050: Conceptual Assessment. Exploring opportunities and challenges for the EU refining industry to transition towards a low-CO₂ intensive economy*; Concawe: Brussels, Belgium, 2019.
- [2] Concawe. *A Clean Planet for all. Impact assessment on the potential implications for our refining system and the link with Refinery 2050*; Concawe: Brussels, Belgium, 2020.
- [3] Kaisalo Noora, *Tar reforming in biomass gasification gas cleaning*; Aalto University, 2017