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FLEXIBILITY OF NUCLEAR POWER:
SMRs, COGENERATION AND OPTION TO SWITCH

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Ai miei genitori,

che hanno permesso tutto questo.

Grazie

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EXECUTIVE SUMMARY

Despite many criticalities mainly arisen after Fukushima accident, the nuclear power can still play a relevant role in the worldwide future energetic panorama [1]. Certainly nuclear energy represents an attractive alternative to tackle most of the issues born in the last century [2], such as the growing energy demand, the greenhouse gas emissions, the scarcity of fossil fuels sources at cost-effective price, etc. On the other hand, to be competitive in the long-term energy needs, various problems must be addressed, regarding waste, safety, security, and non-proliferation issues. Therefore an important goal to increase the acceptance of nuclear energy is to improve its sustainability.

Considering this energetic-economic-social background, the goal of this work is to study the future role of nuclear energy. In particular, Small and Medium-size Modular Reactors (SMRs) are currently at the centre of many researches of scientific community and nuclear industry [3] [4] [5], thanks to their key advantages in matter of economics and alternative applications. For these motivations, the aim of this thesis is to investigate the possibility of “flexibility” of SMRs, given by the possibility to work in load following mode. In order to satisfy this goal, the suitability of nuclear energy for not-electrical application has been evaluated, such as the production of biofuels or desalinated water.

Hence, the idea of this work is to take full advantage of the thermal power generated within the reactor vessel. Historically, Nuclear Power Plants (NPPs) have been mainly seen as a base-load source of energy. However, the important share of nuclear energy in some national electric portfolio (e.g. 75% in France) and the share in the same grid with intermittent source of energy (i.e. solar, wind) (e.g. in Germany) imposed the requirement to work in load following mode even with NPPs [6]. The requirements are specified in [7], that in matter of load following reports:

“a unit must me capable of continuous operation between 50% and 100% of its nominal power (P_n), […]. Load scheduled variations (should be) *2 per day, 5 per week and 200 per year”.*

Nowadays, nuclear reactors follow the grid inserting control rods and neutrons absorbers onto the coolant, then modifying the reactivity within the core [6]. This essentially represents a waste of fuel, a decrease in load factor and it introduces in primary loop some thermo-mechanical stresses and a smooth but constant fatigue cycle.

Since the quality of the produced steam in nuclear technology is intrinsically low and consequently the net efficiency of the Rankine cycle is low as well, the idea is to always work at full nominal power rate, leaving unaltered the primary loop. In order to work in load following mode, during high-load hours (day) the NPP provides at most electric energy to the grid, whereas during low-load hours (night) the not required thermal energy is directed to the auxiliary plant (red area in Figure ES 1).Then, the secondary loop switches production: the low enthalpy heat is directly sent to the coupled plants to produce alternative products (i.e. biodiesel, ethanol and desalinated water). The most suitable not-electrical applications would mostly require low quality thermal energy. In this way, the heat is not converted through a not-efficient electric conversion.

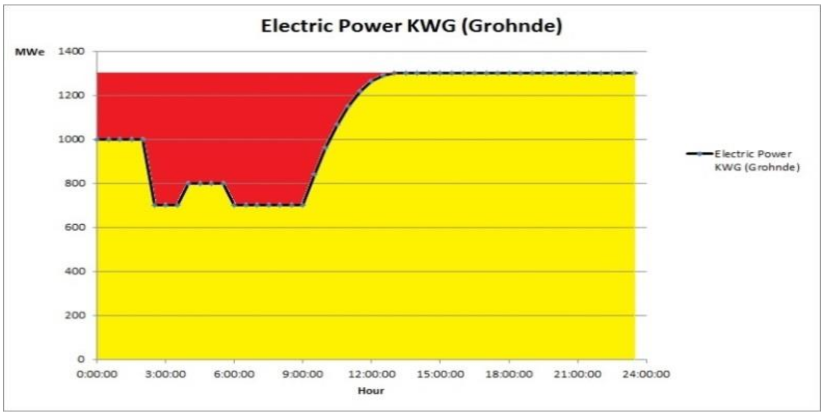


FIGURE ES 1. THE YELLOW AREA IS THE APPROXIMATED EFFECTIVE ELECTRIC POWER SUPPLIED TO THE GRID WITH A PLANT IN GERMANY. THE RED AREA IS THE AMOUNT OF POWER AVAILABLE FOR AUXILIARY PLANTS IN ORDER TO WORK IN LOAD FOLLOWING MODE. IT CORRESPONDS TO ABOUT THE 20% OF THE THEORETIC POWER.

In addition, due to the huge capital costs of the nuclear plant, it is possible to take more advantage of the already built capital-intensive plant.

This work is not going to discuss whether, or not, to build the nuclear site. This work tackles the possible investment's profitability of building a nuclear site, composed by "n" SMRs, in a cogenerative layout with a biorefinery (for the production of biodiesel and bioethanol from microalgal biomass) or a desalination plant, with the technical aim to work in load following mode.

A biorefinery is a plant that has as input mainly a biomass, thermal and electrical energies and as output a biofuel. Many biomasses are used to produce biofuels. Literature divides them in three generations: first generation is composed by conventional crops (corn, soy bean, rapeseed, sugarcane, etc.), second generation is composed by lignocellulosic biomasses (mainly forestry and agricultural waste), whereas third generation is represented by algae [8].

Nowadays the almost entirety of biofuels is produced by first generation feedstock. However, the biorefinery considered in this work is based on a microalgal biomass, because of great recent interest by scientific community and their many advantages (listed in Table ES 1). Because of the modernity of this technology, in literature data related to production of biofuels from algae are reported only from laboratory experiences or pilot plants [9]. Commercial scale plants are under development and only some companies have already started the construction – e.g. [10]. Here, to simulate the biorefinery, the more promising, currently adapted technologies are chosen to simulate every processes involved to produce biofuels from microalgae. The biorefinery modelled provide a whole cycle. The main production phases are [11]: cultivation, harvesting and dewatering, oil extraction, biodiesel production (via transesterification reaction) and bioethanol production (via fermentation). Hence, from the technical point of view, a black-box is obtained (Figure ES 2), characterized by various inputs and various outputs. The black box contain all data collected in matter of processes, chemical transformations, efficiencies and yields.

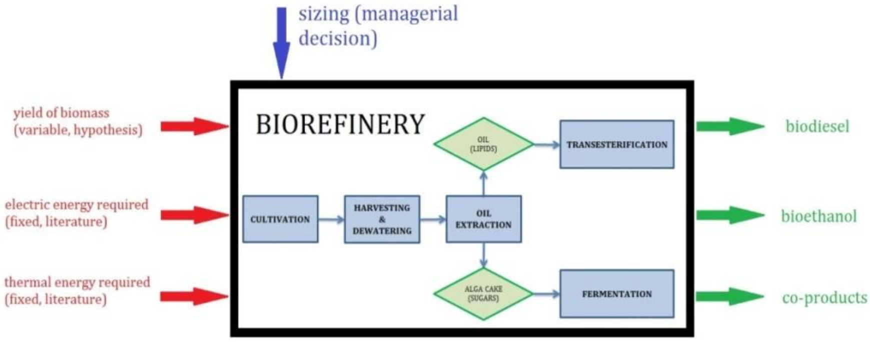
The sizing of the plant is chosen to address the same amount of energy hypothesised also for the case of desalination. In particular, in order to work in load following mode, about 70-80% of the energy produced must be directed to the auxiliary plants. The biggest capacity taken into account in this study is a biorefinery that produces approximately 270 million of litres per year (MLPY) of biodiesel and 45 MLPY of ethanol, using about 340 MWe and 180 MWt. However, once the size is fixed, the IRR of the investment is studied even with different sizes of the plant.

However, the critic analysis of technical data and power requirements suggests that is not possible to install a biorefinery that is working in a flexible way. Indeed, most of the required energy is for the beginning phases of the chain. Because of time needed to grow the biomass and issues of perishability of biomass, this energy must be provided continuously. Though this application represents an interesting opportunity of "static" cogeneration. For this reason the profitability of this plant has been further evaluated by an economic analysis.

TABLE ES 1. BENEFITS OF MICROALGAE, IN COMPARISON WITH OTHER FEEDSTOCKS.

Advantage	Benefit[1]	Compare[2]	References
High photosynthetic efficiency	E, I	I	[12], [13]
Fast growth	I	I	[14], [15], [16]
Short cultivation-harvesting cycle (only 3-10 days)	I	I	[17]
High starch content	I	General	[18], [19]
High lipid content	I	I, II	[15], [20], [13]
Possibility to produce many products and co-products	I	I, II	[12], [21], [17]
Wastewater treatment	E	I, II	[22], [11]
CO_2 fixation	E	I, II	[12], [23], [13]
Not compete with food	E	I	[24], [25], [26]
Not influence food prices	E	I	[24], [25]
Low water requirement	E, I	I	[27], [25]
Low land requirement (high biofuel yield per ha)	E, I	I	[13], [25], [28]
Not arable land requirement	E, I	I	[29], [16], [21]
Easy adaptability to climate	I	I	[26], [30]
Growing in many different environment (water condition)	I	I	[21], [31]
Conventional pretreatment	I	II	[32], [33], [34]

FIGURE ES 2. THE BIOREFINERY MODELLED IS LIKE A BLACK-BOX THAT CONTAINS EFFICIENCIES, YIELDS AND TECHNICAL ASSUMPTIONS FOR EVERY PROCESSES INVOLVED.



[1] It means if the advantage represents a benefit from the industrial (I) point of view (i.e. technical or economical advantage) or from the environmental (E) point of view (i.e. reducing global warming or more sustainability).
[2] This column refers in comparison to which other biofuel generation the benefits represent an advantages. "I" indicates the first generation; while "II" indicates the second generation.

The desalination technology instead is an already reliable application. First desalination plants were built in 1960s and the growth rate is increasing very rapidly, especially in the last decade. It is currently about 55% per year [35]. The main processes to produce desalinated water can be classed in two groups: membrane or thermal. For the purpose of this thesis and other advantages explained in chapter 3, a thermal process is selected. In particular the correct technology is a MED-TVC plant. Technical considerations allow to affirm that with this plant is possible to work in a flexible mode. The limitations regard the minimum power level that must be supplied to the MED-TVC and to the steam turbines when they are not used. In particular, for the desalination plant, this level has been fixed at 25%. This percentage guarantees the immediate availability of standard quality water production, during the nocturne hour.

To proceed into the economic analysis, firstly the technical parameters (size, power, products, etc.) are found, studying a possible layout. A site composed by 4 IRIS reactor, i.e. 4000 MWt, is considered. If roughly the aim is to direct 50% to the auxiliary plants, the layout of the site is designed in the following way:

- Two IRIS always produce only electricity. They are always connected only to the grid, working at full power load
- Other two IRIS instead are connected to the desalination units, as well.

Therefore, considering also the minimum no load level for the steam turbine, 1844 MWt must directed to MED-TVC. Since we assume a thermal energy need of 50 kWh/m^3, having a plant availability of 90%, the size has to be fixed to 885120 m^3/d, a similar size of the MED-TVC in Jubail (Saudi Arabia).

Subsequently the economic analysis is developed using both the Discounted Cash Flow (DCF) method and the Real Options Approach (ROA) analysis. DCF methodology, although is simple and easy to implement, is a static, very rigid analysis and presents three substantial criticalities:

1. The choice of the discount rate (that represents the only object owned by the investor to capture the risk associated to the investment)
2. The weak consideration of stochastic nature of the cash flows: DCF can't capture any market uncertainties, like electricity and fuel prices, or technical uncertainties, like construction costs that vary considerably along construction time [36]
3. The passivity of the management, unable to take advantage from the resolution of some uncertainties [37]

To overcome these issues, especially regarding uncertainties and management passivity, a new more dynamic tool was proposed: the *Real Options Approach (ROA)*. The key differences between DCF and ROA are listed in Table ES 2. However, ROA is not a substitute for DCF. It is an auxiliary tool that fills the gaps that DCF cannot address [38]. ROA uses DCF as a building block and captures the value of the options. The main advantage of the real options is that, if properly managed, options can create an extra value and reduce risk for the investors that own them [39].

Hence, two options are used to study the profitability of the investments: the Option to Build and the Option to Switch. In order to complete the analysis two algorithms are developed for both the options. In particular, the option to build analyses the investments to build a plant that is working "statically", both in case of the biorefinery and of the MED-TVC. Then the option to switch is used to study the eventual extra profitability, given by the possibility to work also in a flexible way, only for the case of the desalination.

TABLE ES 2. DIFFERENCES BETWEEN DCF AND ROA.

Discounted Cash Flow	Real Options Analysis
Does not capture the value of managerial flexibility during the project life cycle	Recognizes the value of managerial flexibility to alter the course of a project
Uncertainty with future project outcomes not considered	Uncertainty is the key factor that drives the option's value
Undervalues the asset that currently (or in the near term) produces little or no cash flow	The long-term strategic value of the project is considered because of the flexibility with decision making
Expected payoff is discounted at a rate adjusted for risk. Risk is expressed as a discount premium	Payoff itself is adjusted for risk and then discounted at a risk-free rate. Risk is expressed in the probability distribution of the payoff
Investment cost is typically discounted at the same rate as the payoff (risk-adjusted rate)	Investment cost is discounted at the same rate as the payoff (risk-free rate)

The main uncertainty that is modelled for the option to build is the yearly fluctuations of products' prices (i.e. biofuels, water and electricity). On the contrary for the option to switch the daily and weekly trends of the E.E.'s price are simulated. On the other hand the price of water is kept constant within a year.

The main goals of the algorithms are the following:

- for the <u>Option to Build</u>: to solve the uncertainty regarding market prices, for unstable and unknown markets. The algorithm should capture the profitability of the investments, when the uncertainties are evolving in an attractive (positive) way
- for the <u>Option to Switch</u>: to take advantage of the daily fluctuation of prices. When the E.E. is cheaper, the plant can produce more water than the hours in which the E.E. is more expensive. This advantage exists only whether the prices of electricity and water are combined in a particular way, providing comparable revenues.

Finally the results obtained with his work, can be divided in technical and economical.

TECHNICAL RESULTS

For the biorefinery, the technical analysis in matter of biology of the microalgae and power requirements allows to affirm two main statements:

- 5 scenarios are evaluated. Each one is characterized by a different yield and/or technology of cultivation of biomass. Only one scenario is suitable for a coupling with a 4 IRIS site, that one characterized by a cultivation phase in fermenters. All the others scenarios need a huge land requirements (thousands of hectares) for a reasonable coupling with a nuclear site. With fermenters scenario, a part of the sustainability of the project is lost (regarding CO_2 absorption/emission)
- the biorefinery can't work in discontinuous mode and therefore is not suitable for the load following

Concerning the desalination instead:

- it is possible to work in load following mode
- the required size of the plant is similar to the largest plants worldwide
- a minimum quantity of steam must be supplied to the steam turbines, even when they are not producing E.E. (no load value equals to 7,8%), in order to prevent superheating issues
- it is also advisable to supply a minimum quantity of power even to the MED-TVC plant during the daily hours. This fact is not mandatory to prevent technical problems on the system or on the material, but it is reasonable from a managerial point of view. Actually the plant begins to produce a sealable quality of water, at least at 20-30% of the power load. Consequently, in order to prevent losses and to take better advantage of a capital intensive plant, it is assumed to work with a minimum power level of 25% during the daily hours.

ECONOMICAL RESULTS

The economical results affirm that the products' prices that are requested to reach the break-even point of the investment are in the same order of magnitude of the market prices of the real world (Figure ES 3).

The sensitivity analysis (Figure ES 4) for both the plants confirm that the most important parameters that affect the final results are the prices of the involved products. Other important variables are the discount rate (WACC) and capital costs of the plant (especially for the desalination case).

The results obtained with the option to build confirm the theory of real options: the ROA can add an extra value to the investment's profitability. In addition, the graphs show that after about 4-6 years most of uncertainties are solved and the NPV reaches a maximum or an asymptotic value (Figure ES 5). This is caused by the fact that the advantage to wait to solve further uncertainties, is taken over by the higher discount factor, that lows the future inlet cash flows. Moreover, algorithm is broadly more promising with scenarios with a wide NPV standard deviation around zero. Actually, if the NPV calculated with the Monte Carlo DCF is already very positive because the investment is very attractive, the ROA theory can't add an extra value to the found result (Figure ES 6). However the value of the option to build strongly depends on the scenarios, both for the biorefinery and for the desalination plant.

FIGURE ES 3. TREND OF IRR AT VARYING O BIODIESEL PRICE. VALUES CALCULATED WITH A CLASSIC DCF METHOD. THREE TRENDS ARE OBTAINED FOR THREE DIFFERENT CAPACITIES OF THE BIOREFINERY.

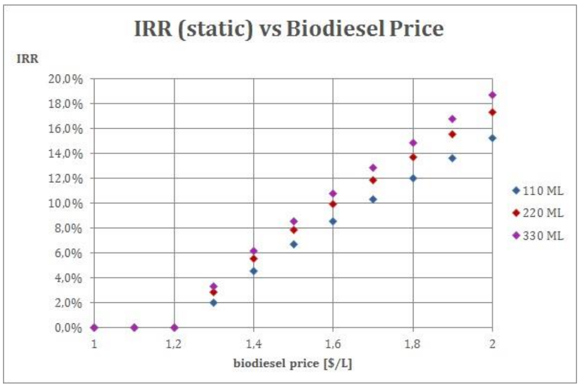


FIGURE ES 4. SENSITIVITY ANALYSIS FOR THE BIOREFINERY CASE. ANALYSIS DONE FOR THE NPV CALCULATED WITH A DCF MONTE CARLO, AT YEAR 0.

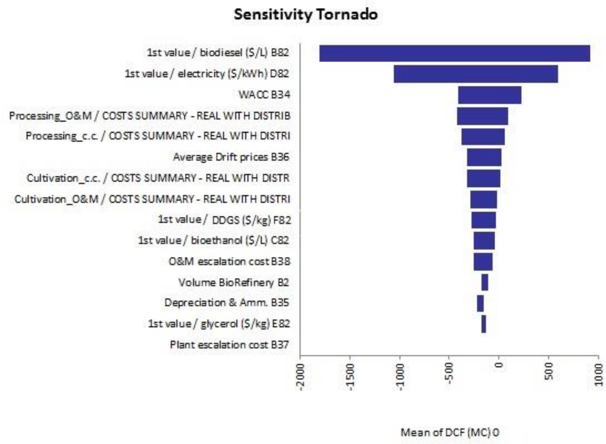


FIGURE ES 5. TREND OF CALCULATED NPV FOR DCF METHODS AND ROA AT DIFFERENT YEARS. CASE OF SCENARIO 7 (BREAK-EVEN CASE) FOR MED-TVC PLANT.

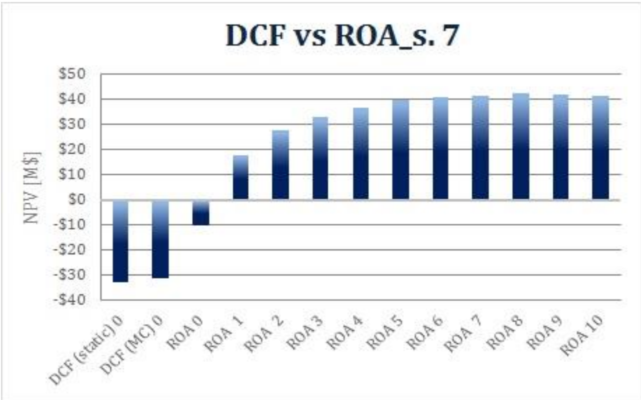
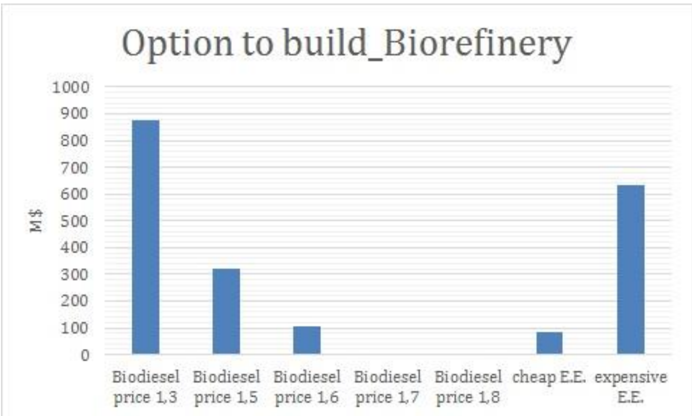


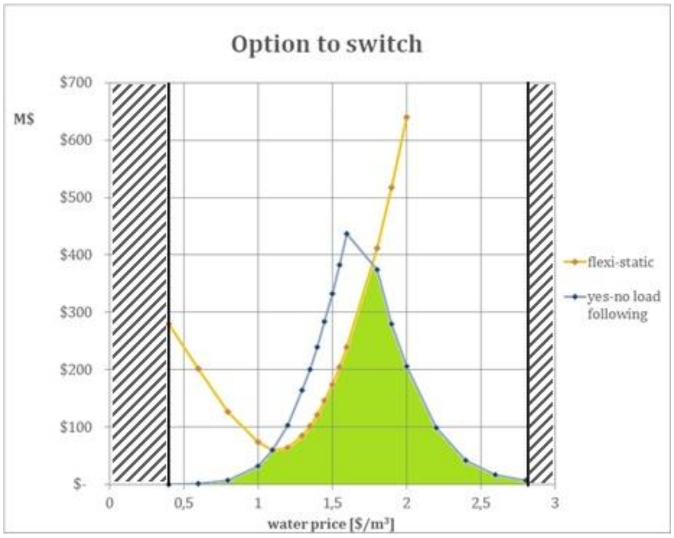
FIGURE ES 6. THE VALUE OF THE OPTION TO BUILD FOR DIFFERENT EVALUATED SCENARIOS FOR THE BIOREFINERY.



Finally the option to switch assert that, if the revenues of water and E.E. are comparable, because the prices are combined in a particular way, then there is an effective advantage to work in a flexible load following mode. The result is drawn in Figure ES 7. The price of E.E. is chosen standard (average daily price of 0,08 $/kWh). Then, varying the price of water, the advantage of switching changes. In particular, if the price of water is to low or too high, there is not advantage to switch production. This is because to produce E.E. is always more profitable than producing water (dotted area on the left) or vice versa (dotted area on the right). Instead in the center, the value of the option to switch (green area) is obtained comparing two curves: the yellow one is obtained comparing a static load following with a flexible load following; the blue one is obtained comparing a flexible load following with a no-load following arrangement.

It is interesting to note that effectively it is possible to add an extra value, simply switching on/off the plants. The range of water price in which the option has a value belongs is strongly comparable with the price of water in the real market.

FIGURE ES 7. VALUE OF THE OPTION TO SWITCH, IN RELATION TO THE PRICE OF BIODIESEL. GRAPH OBTAINED FOR SCENARIO 5 "MARKET".



CHAPTER 1. INTRODUCTION

In the first chapter, the main themes faced in this thesis are introduced. In particular, a brief description of small modular reactors peculiarities permit to focus the current situation in the development of nuclear technology. Subsequently, also the modern issue for power plants to work in load following mode is explained. This is the main goal of the dissertation. Indeed, in this thesis the possibility to work in load following mode with SMRs thanks to a diverse methodology has been suggested. In the following of the chapter, the descriptions of both the auxiliary plants are summarized, chosen to couple with the nuclear power site: a "microalgal" biorefinery and a "thermal" desalination plant. Finally, the economic approach of Real Options, used to evaluate the investment, is presented at the end of the chapter.

1.1 THE ROLE OF THE SMALL MODULAR REACTORS

"The Fukushima nuclear accident, in March 2011, seemed to squelch the so-called "nuclear renaissance", but many countries – including the U.S., South Korea, Russia, China, and even Japan – are moving ahead with plans for small reactors that can be factory-crafted (thus "modular") and assembled on-site" [40].

From this statement it is clear that, despite many criticalities, nuclear energy can still play a relevant role in the worldwide future energetic panorama [1], [41]. Both for the continuous innovative challenges that such an energy has faced in its history and is still going to face in the nearby future, and for the modern global issues that have arisen in the last century: growing energy demand, growing population, environmental and climatic problems, e.g. greenhouse gas (GHG) emissions, and all the well-known issues related to the usage of fossil fuels. Nuclear energy is representing an attractive alternative to tackle most of these problems.

On the other hand, to be competitive in the long-term energy needs, either globally or regionally, some mandatory solutions have to be found, regarding waste, safety, security, and non-proliferation issues (sustainability of nuclear energy) [42], as well as the capital costs of construction [2]. Small and Medium-size Modular Reactors (SMRs) could successfully address several of these difficulties. The term "modular" in this context refers to a single reactor that can be grouped with other modules to form, eventually, a larger nuclear power plant. SMRs offer simpler, standardized, and safer modular design by being factory built, requiring smaller initial capital investment, and having shorter construction times. Moreover SMRs could be small enough to be transportable by conventional lines, and offer a wide number of possibilities of applications: they could be used in remote locations without advanced infrastructures and for small electric grids, or could supply auxiliary thermal and electric energy for non-electric applications, replacing the existing fossil fuels boiler, or, as said, could be clustered in a single site to provide a multi-module, large capacity power plant. In the following section, the current trend in the development of nuclear technology and the main features of the SMRs are briefly explained.

1.1.1 Current trend in nuclear plants

After the construction of the first commercial nuclear power plant (NPP) in 1957, Shippingport (USA), of 60 MWe, the trend in the design of the subsequent reactors has been to increase their sizes, as it is possible to note in Figure 1 [43]. Much of this order of magnitude's design scale-up occurred over a 15 years period without the benefit of operating experience from their smaller predecessors. As plant sizes grew and as operational issues began to curb the industry's confidence in the ultimate safety of the plants, more stringent safety requirements were imposed and the simplicity of the original light water reactors (LWR) gave way to a complex layering of redundant safety and auxiliary systems. The main reasons that suggested to design bigger NPPs was to take advantage of economies of scale, and therefore to curb the costs. Unfortunately, this escalation of plant complexity increased rapidly the costs, and created licensing, construction and operational delays, and, in addition, decreased the confidence of the investors in the profitability of this energy source.

Nowadays, in the "nuclear renaissance" [44], much effort is made to design modern NPPs of a reduced size: indeed many designs of Generation IV belong to the category of Small and Medium-size Reactor [45], [46]. The definition that the

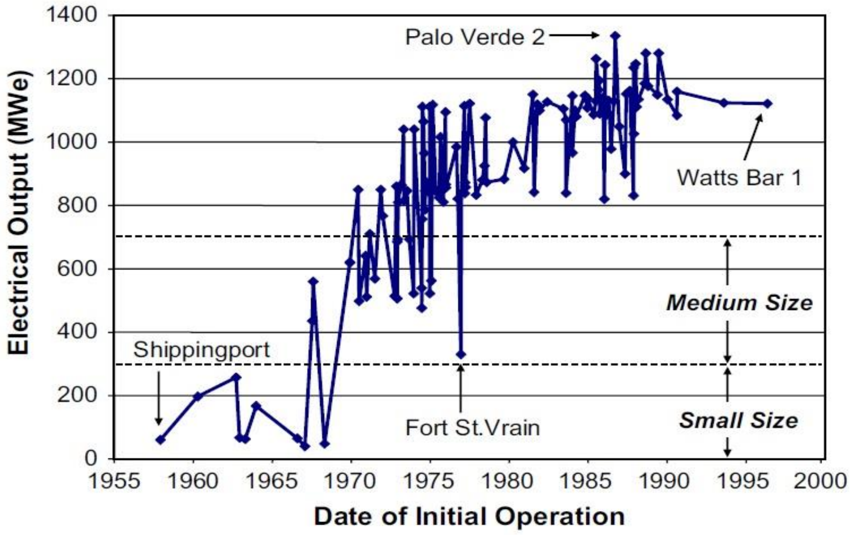


FIGURE 1. ELECTRIC OUTPUT OF U.S. COMMERCIAL NUCLEAR POWER PLANTS [3]

International Atomic Energy Agency (IAEA) gives for SMRs is the following: a “small” reactor as one having electrical output less than 300 MWe whereas a “medium” reactor as one having output between 300 and 700 MWe. Therefore a “large” plant is assumed to produce more than 700 MWe. This definition is given in term of power output; not surprisingly, most SMRs are also physically smaller than large plants. Both their lower power capacity and higher compactness contribute to many benefits, especially in matter of plant safety, fabrication, operations, and economics (see next section). To clarify, the acronym SMR has also started to be used for Small Modular Reactor. It is not clear where this semantic shift originated, but the effect is confusing [47]. Hence in this text “SMR” refer both to Small and Medium-size Modular Reactor, and to Small Modular Reactor.

According to the definition of IAEA, currently out of the 437 commercial power plants operating all over the world [48], about one third are SMRs. However, most of these are simply a scale-down design of a plant which is more conventional and larger in size. The most innovative effort is designing innovative layouts of reactors of reduced size: so called deliberately small reactors [43]. The aim of such current studies is to take advantage of what similar small plants can offer and therefore achieving specific performance characteristics. Their smallness confer them to be suitable both to produce a more conventional base-load electricity generation, and to suit some specialized non-electric applications, in cogenerative layouts.

There are three major groups of SMRs’ design:

1. Reactors based on the design concepts of proven and widely utilized light water reactors (LWRs)
2. Gas-cooled reactors
3. Advanced SMRs, with more exotic design, being cooled either by liquid metals or liquid salts.

This study will analyse the suitability of a cogeneration with other plants to supply, thanks to the NPP, both the electric and thermal requirements, the latter directly via steam. Therefore the main designs considered belong mostly to the group number 1 and the group number 3 (Rankine cycle). Moreover, because of the easier applicability of an already well-known technology in water reactor, SMRs of the first group seem closer to a practical application. Hence the attention of this work is more focused on small and medium sized modular, light water reactors (e.g. IRIS). However since the only parameters considered for the suitability of this study have been the nominal power of a single modular unit and the total power installed in a site, the results of this thesis are easily applicable even to NPPs with different designs.

1.1.2 Benefits of SMRs

The advantages that these small and compact nuclear reactors offer can be categorised in four classes: fabrication and construction logistics, plant safety, operational flexibility, and economics [43].

The physical compact size of such reactors confer many improvements in matter of fabrication, transportation and final construction: in large reactors, huge components like the reactor vessel or the steam generators require very specific manufacturers, difficult lines for transportation and complicated procedures for the final assembly. In addition these issues restrict the number of sites suitable for the allocation of the plants. In contrast, small reactors greatly reduce or eliminate the need for forgings, as they can be transported by more conventional ways (i.e. truck, rail or barge), and also they can be fabricated in a more controlled factory environment and, at last, shipped to the site for the final assembly (so called *"plug and play"* installation).

The plant safety for a SMR is intrinsically increased by the following factors: less quantity of radionuclides produced by nuclear fissions, less number of vulnerable parts that can be involved in accident (i.e. circulation pumps, pipes, etc.), and the opportunity to passively respond to unexpected transients. SMRs have some design features that can be adopted solely thanks to their compactness: for example, in IRIS design, the main vessel incorporates steam generators, the reactor core, an internal control rod drive mechanism and the pressurizer, eliminating the need for many large pipes and auxiliary pumps. Moreover, most of the designs are based on passive emergency systems, can enjoy anti-seismic protections for more traditional buildings (like seismic platforms), and can be hollow underground for a more efficient defence against accidental or terroristic aircraft crashes or other external accidents. In addition, even in case of an accident, SMRs have a higher efficiency in decay heat removal. Reducing the quantity of radioactive material within the plant and reducing the possibility of accident scenario, enhances the general safety. In particular, the higher the intrinsic safety, the lower the licensing requirements. For this reason, SMR can enjoy reduced shielding, reduced site boundary and reduced emergency planning zone (EPZ). Furthermore the latter factor eases the construction of the plant in a cogenerative layout, reducing thermal losses due to long pipes, and eases the decision on the construction site.

Other technical advantages are that the reduced size can adhere better to the grid demand, which provides grid stability, and has a lower requirement for water.

Also interesting considerations exist from the economic point of view [49]. Many researches are conducted by companies, institutions and universities in order to study the profitability of building SMRs [5]. In order to evaluate the attractiveness of an investment in nuclear power plant, one should take into account the economic side of the investment. Firstly there are financial parameters: net present value (NPV), internal rate of return (IRR), LUEC (economic evaluation production cost), etc. Secondly other issues to be considered are the associated risk and cash flows, capital cost, construction time, etc. In forecasting and analysing construction costs, there is a prevailing attitude that economy of scale (EOS) dominates all other considerations and consequently that smaller sized plants are not economically viable. But an associated study by Westinghouse Electric Company is quantifying several factors that can mitigate EOS for SMRs [50]: factors that are independent of plant size, but that are favourable in case of small design (modularization, factory fabrication, shared site infrastructure, process learning, etc.); factors that are unique for SMRs (simplifications, demand matching, economy of replication, etc.). In a work of Politecnico di Milano (Italy) in SMR Economics Evaluation the code INCAS ver 1.0 was used for evaluating the generation cost and key financial and economic parameters for small modular reactors in comparison with other sources of energy and NPP of bigger scale [51]. The conclusions of this study assert that the economy of scale law could be overcome by other SMR features, leading SMRs to competitiveness. Basically, the peculiarities of small and modular reactors create possible "economies of multiples" instead of economies of scale [52]. An example of a SMRs advantage is the cash flow profile, that is smoother and with a lower upfront investment; so the SMR projects can be more attractive for investors. In other words, the maximum cash outflow is lower: as one module is finished and starts producing electricity, it will generate positive cash flow for the next module to be built (peculiarity named "self-financing"). Lower overnight costs and reduced maximum cash outflow, permit also to expand the market, making nuclear reactors affordable even by many utilities and small developing countries. Regarding net present value and IRR, four SMRs are comparable with one large LWR. In addition, successive construction of SMR units compared to a single large plant is much less sensitive to construction delays and market variations (prices of electricity, discount rate, etc.). A modular investing strategy with a step-by-step power block deployment process allows lower financial exposure and less capital at risk, and may mitigate the impact of scenario uncertainties on a project's profitability. Therefore, even if more conventional big size non-modular reactors can be more profitable in scenarios where conditions are more predictable, SMRs appear to be more suitable as an option to control financial risk [5].

Then, this thesis wants to study eventual further investments in auxiliary non-electric applications to couple with the NPP, especially in unpredictable/flexible scenarios/markets. As explained in paragraph 1.5, sources of uncertainties are the basis to implement and to take advantage of an economic analysis with Real Options. Therefore, here the construction of a SMR site is taken into consideration. In particular, in this work, we are not going to discuss whether or not build the nuclear site. We are going to analyse if, thanks to the features discussed in this paragraph, it's possible to couple a further plant for thermal applications (cogeneration), to add an extra worth to the initial investment of the NPP. The economic analysis faced in chapter 5, tackles the possible investment profitability of building a nuclear site, composed by "n" small modular reactors, in a cogenerative layout with a biorefinery (for the production of biodiesel and bioethanol from microalgal biomass) or a desalination plant.

1.2 WHAT IS LOAD FOLLOWING?

During their history, nuclear power plants (NPPs) have been mainly seen as a base-load source of electricity. The main reason for this fact is that operating a

NPP at the rated power level is usually more efficient economically, and simpler. This mode of operation was possible because the share of Nuke in most countries' energy mixes was very small, and thus the manoeuvring capabilities of the plant were typically limited to safety needs (e.g. safe shutdowns in case of load rejection) and frequency regulation required by the electric grid operator. However that situation is changing in several countries [53]. The share of nuclear power in the national electricity mix of some of these countries had become so important that the utilities had to implement or improve the manoeuvrability capabilities of their NPPs, to be able to adapt the electricity supply to daily or seasonal variations of the power demand (load following). For example, this is the case in France where more than 75% of electricity is generated with atomic energy. Another motivation for load following with nuclear power plants comes from the large-scale deployment of intermittent electricity sources (mainly renewables, like solar energy or wind farms). This could be the case of Germany. If there is a significant share of intermittent and nuclear power sources on the same electric grid, NPPs must be able to operate in a load following mode to balance the fluctuations of the total power generation, and in this case unexpected large and rapid modulations of the power demand could occur [6].

1.2.1 Different modes to operate for a power plant

Therefore, there are four operation modes currently used by nuclear power plants:

- Base-load generation mode
- Primary and secondary frequency control
- Load Following mode

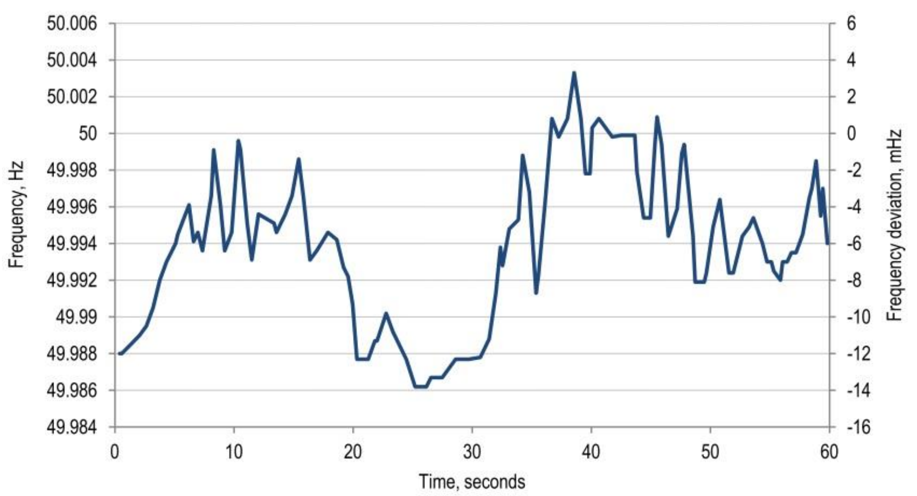
BASE-LOAD MODE

In base-load operation mode, the nuclear power plants operate at constant nominal power (P_n) during the majority of the time.

PRIMARY AND SECONDARY FREQUENCY CONTROL

The power demand can never be exactly evaluated in advance and thus there is a certain random variation of demand resulting in frequency fluctuations (see Figure 2), typically of less than 20 mHz. The power plants have to monitor the

FIGURE 2. EXAMPLE OF THE FREQUENCY VARIATION ON THE GRID IN EUROPE. TARGET IN EUROPE IS FIXED AT 50 HZ.



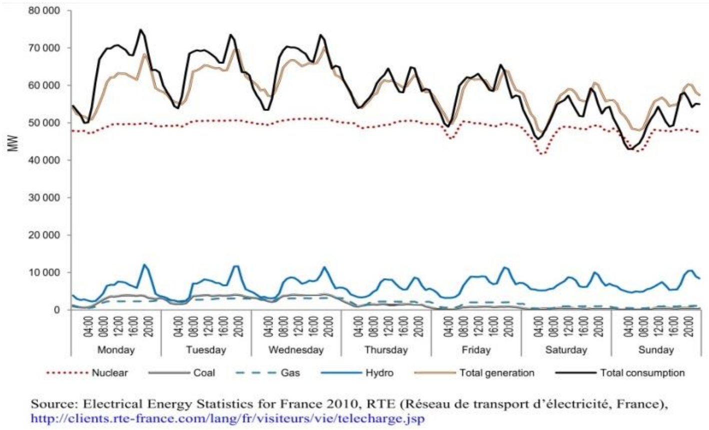


FIGURE 3. EXAMPLE OF ELECTRICITY GENERATION IN FRANCE DURING A WEEK IN NOVEMBER, 2010 [7].

frequency on the grid and immediately adapt their level of generation in order to keep the frequency stable at the desired value. This kind of regulation is called *primary control*. The power modulations for the primary control in frequency regulation are performed in the interval of ±2% of P_n. In addition, the primary frequency control allows short-term adjustment of electricity production and demand in the time frame of about 2 to 30 seconds after the deviation is observed.

Another type of frequency regulation, named *secondary control*, acts over a longer timeframe (say, from several seconds to several minutes) and restores the exact frequency by calculating an average frequency deviation over a period of time. The secondary control is particularly important because of the interconnection of the grid of a country with other European grids. In order to adjust the frequency, taking into account the balance of electricity exchanges with other European grids, the grid operator sends a digital signal to the NPP to modify their power level in the interval of ±5% P_n [6].

LOAD FOLLOWING

Nuclear power plants operating in load following mode follow a variable load program with one or two power changes over a period of 24 hours. The load pattern is determined by the grid operator and the utilities, depending on the power demand and the manoeuvring capabilities of the plant itself. Depending on the load pattern, several intervals of power ramps are authorised ranging from 1% of P_n per minute to approximately 5% of P_n per minute [6].

According to [54], currently slow ramps of less than 1,5% of nominal power per minute are most often used in France and the typical low power level is about 50%.

1.2.2 The current situation worldwide

Modern nuclear light water reactors (LWRs) have strong manoeuvring capabilities. Nowadays, NPPs can operate in load-following mode, i.e. participate in the primary and secondary frequency control, and some units follow a variable load program with one or two large power changes per day. For example, in countries such as France, load-following is needed to balance daily and weekly power variations in the supply and demand of electricity, since nuclear energy has a large share in the national mix (see Figure 3). In addition, in countries such as Germany, load-following has become important in recent years due to a large share of intermittent sources of electricity generation (e.g. solar or wind farm), introduced into the national mix (see Figure 4).

The minimum requirements for the manoeuvrability capabilities of the modern reactors are defined by the utilities needs that are based on the requirements of the grid operators. For example, according to the current version of the European Utilities Requirements (EUR) the NPP must, at least, be capable of a daily load cycling operation between 50% and 100% of its nominal power P_n, with a rate of change of electric output of 3-5% of P_n per minute.

To satisfy the grid requirements, a NPP has two possibilities to balance the thermal powers of the primary and the secondary loops:

- Maintain the average temperature of the primary circuit constant, or
- The pressure in the secondary loop remains constant

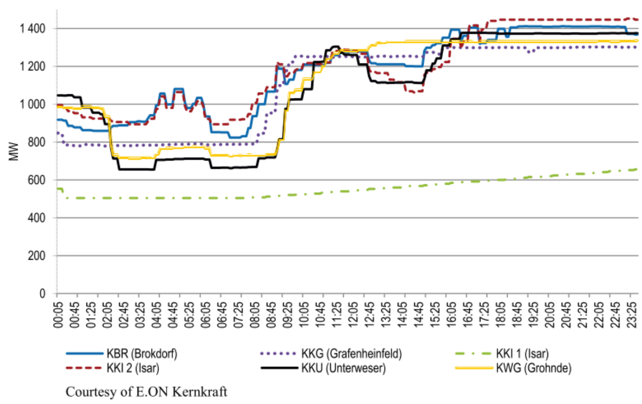
Various combinations of these options are possible. However, the most important thing is that, for regulating the primary coolant temperature or the secondary loop pressure, a reactivity control within the vessel has to intervene.

A reactivity control can be done by essentially moving the control rods and/or managing the boric acid in the primary circuit. Both of these ways mainly consist in introducing strong neutrons adsorbing materials, to manage the neutronic population within the reactor core, and therefore the power produced.

1.2.3 Requirements to follow the grid

In 1991, five European utilities[3] considered that a more open specification would be needed to cover a wider range of designs, and thus the European Utilities' Requirements (EUR) were created. The EUR cover a broad range of conditions for a NPP to operate efficiently and safely. They include such areas as plant layout and specifications, systems, materials, components, probabilistic safety assessment methodology and availability assessment.

FIGURE 4. EXAMPLE OF THE ELECTRICITY GENERATION WITH SOME GERMAN NUCLEAR POWER PLANTS (PWR AND BWR) [7].



[3] British Energy/Nuclear Electric, EDF, Tractabel and groups of German and Spanish utilities

In matter of power fluctuations (in particular in matter of load following and primary and secondary control[4]), the EUR requirements are [7]:

1. *"The unit must be capable of continuous operation between 50 and 100% of its nominal power P_n. The standard plant design shall allow the implementation of scheduled and unscheduled load following operation during 90% of the whole fuel cycle. Restrictions are due to fuel conditions at the end of the cycle. The rate of change of electric output shall be 3% of P_n/min. In addition, the unit shall be expected to go through the following number of load scheduled variations, each variation being defined as a transient from full power to minimum load and back to full power: 2 per day, 5 per week and cumulatively 200 per year".*
2. *"The unit shall be capable of taking part in the primary control of the grid. This is a prerequisite for connection to the grid. The primary control range shall be ± 2% of the nominal power (mandatory), but higher values may be agreed between system operators and plant operators, though not higher than ± 5% of P_n."*
3. *"The standard plant design shall allow the implementation of a secondary control (optional). Participation in secondary control is based on an agreement between the grid operator and the electricity production company. The secondary control is a central control (manual or automatic) of selected regulating plants or units within an area to restore the frequency and the net power exchanges to their scheduled values (on a time scale of a few minutes). The specifications in detail are part of the agreement. The minimum control range for secondary control operation shall be ±10% of P_n above the minimum load taking into account the control range. The variation rate shall be 1% of P_n/min. Higher values may be agreed between system operator and plant operator, though not higher than 5% of P_n /min."*

1.2.4 Why proposing an alternative to the current state-of-the-art?

Introducing modifications in the primary loop to follow the grid, i.e. mainly inserting control rods, also introduce some not negligible issues. They could create some problems from both the technical and the economical point of view. Technical aspects of load following involve counter reactions of reactivity (like moderator effects, Doppler effects and change in the power distribution in the core), poisoning by fission products (i.e. Xenon) and consequences of the fuel burn-up. In particular, regarding the latter point, at the end of the fuel cycle the boron concentration is almost zero within the coolant, and the control rods are in upper position. Thus the margins for the manoeuvrability decrease. In addition poisoning and counter reactions of reactivity make the operations rather delicate. Moreover, fluctuations in thermal power inside the core create thermo-mechanical stresses, submitting main components of the plant to a continuous fatigue cycle.

The economic consequences of load-following are mainly related to the reduction of the load factor. Basically, introducing neutron absorbers means to continue the fission chain, burning fuel, without producing energy. Unfortunately, in case of nuclear plants, fuel costs represent a small fraction of the electricity generating costs, especially if compared with fossil fuel sources (i.e. gas and coal). Thus, operating at higher load factors is more profitable for nuclear power plants, since they cannot make savings on the fuel cost while not producing electricity

[4] There are also other requirements in matter of power fluctuations, for example in case of emergency load variations; but they are not investigated in this study.

(fixed costs are definitely more important than variable costs in nuclear technology). Moreover, there is some influence of the load-following on the ageing of some operational components (e.g. valves), and thus one can expect a slight increase in the maintenance costs. Therefore, to retrieve as much as possible the huge investment costs already shouldered, the load factor has to increase. If it is not possible to sell electricity because in some hours there is less demand, we suggest to produce something different (not-electrical applications). The auxiliary applications are also chosen to improve the sustainability of nuclear energy, supporting the "green economy".

For all these motivations, the purpose of this study is to suggest a different way to work in load following mode, leaving unaltered the primary circuit. The idea is to direct the steam of the secondary loop in a different plant, whenever it is necessary to follow the grid. In this way, the reactor core always operate at full nominal power, and therefore no stresses are introduced in the components of the plants, and management of the operations within the vessel are kept constant and simpler. In addition, it is possible to take full advantage of the fuel and the plant in general, without wasting neutrons, and using thermal power to produce alternative co-products instead of electric energy, whenever is necessary.

In this study a nuclear site composed by 4 SMRs is considered. To reach the goal of working in load following mode, it has been set that in less demanding hours, for example during the night, only 50% of the nominal electric power is given to the grid. The rest of thermal energy produced is switched to other plants, like a biorefinery and a desalination plant.

1.3 BIOFUELS AS RENEWABLE SOURCE OF ENERGY

In order to work in load following mode, two plants were selected to couple with the nuclear power plant: a biorefinery and a desalination plant. In this paragraph biofuels are introduced, with their main features and a brief historical review. A similar presentation about desalinization will be done in paragraph 1.4.

1.3.1 Why interest in biofuels?

In 2012 the annual world primary energy consumption was estimated at 12479 million tonnes of oil equivalent (mtoe). Fossil fuels accounted for 86,9% of the primary energy consumption, with oil (33,1% of share), coal (29,9%) and natural gas (23,9%) as the major fuels, while nuclear energy, hydroelectricity and other renewables accounted for 4,5%, 6,7%, and 4,7% respectively of the total primary energy consumption [55]. Excessive consumption of fossil fuels, particularly in large urban areas, has resulted in generation of high levels of pollution during the last few decades. Fossil fuels are the largest contributor of greenhouse gases (GHGs) to the biosphere [56]. Furthermore, it is well known that their combustion releases in the air a sizable quantity of particulates, i.e. PM 5 or PM 10, and nitrate oxides (NO_x) [57]. Moreover, with the expansion of human population and increase of industrial prosperity, especially in new world-powers, global energy consumption also has increased gradually. By 2030, CO_2 emission from road transport in China are expected to have increased by a factor of 3.4, whilst in India they are predicted to have raised by a factor of 5.8 [58]. Additionally, transport fuels (i.e. diesel and gasoline) are affected by limited reserves of hydrocarbons, and the annual global oil production will begin to decline within the near future [59]. Even if current capacity of extracting hydrocarbons will be improved, accepting for example to pay a higher price for boring, fossil fuels prices are always subjected to a great volatility, mainly due to speculations and politics and historical-cultural background of main producing countries (Middle East, East Europe, China).

In this context, countries across the globe developed state policies toward the increased and economic utilization of biomass for meeting their future energy demands, in order to meet the target of a carbon dioxide reduction of 5.2% from 1990 values, as specified in the Kyoto Protocol of 1997 [60] as well as to decrease reliance and dependence on the supply of fossil fuels . The focus on Climate Change in Europe that led to the 20-20-20 targets is a consequence of this trend. These targets require countries to decrease GHG emissions, increase electricity generation from renewable sources, and decrease whole consumption by 20%. All EU-27 countries intend to include biomass in their portfolios to obey these targets by 2020 [22]: the European Union directive of 2006 had set the goal that by 2010 each member state should have achieved at least 5.75% biofuel usage of all fuel used for transportation. By 2020 this figure should be increased to 10% [61].

1.3.2 First, second and third generation of biofuels

Biofuel is a fuel obtained from a biomass. Literature usually splits the history of biofuels into three generations. The first one consists in biofuels derived from sugar, starch and vegetable oil, extracted by conventional crops. Worldwide ethanol is the most common biofuel, especially in USA and Brazil, whereas biodiesel is the most common biofuel in Europe. Other biofuels that are commonly produced are biogas (methane) and syngas (a mixture of carbon monoxide, hydrogen and other hydrocarbons). Ethanol is produced via a fermentation reaction of sugars with yeast, while biodiesel is obtained via a different chemical reaction, called transesterification, of oil extracted from the biomass. The most common first generation feedstock for production of biofuels are listed in Table 1.

Nearly all ethanol is derived from starch- and sugar-based feedstock. Corn is the leading crop used in USA to produce ethanol, while sugar cane is the most widely employed feedstock to obtain ethanol in Brazil. On the other hand the main feedstock to produce biodiesel are rapeseeds (in Europe) and soybeans (in USA). Table 2 and Table 3 list the top countries in matter of production of ethanol and biodiesel, respectively [62].

Currently, about 1%, 17 million hectares according to [63], of the world's available arable land is used for the production of biofuels, providing 1% of global transport fuels. Competition with food market, and notable demands of water and land were the main issues that induced to explore different feedstocks.

The advent of second generation biofuels was intended to produce fuels from the whole plant matter of dedicated crops, agricultural residues, forest harvesting residues or wood processing waste [64]. The crops under consideration are mostly grasses (switchgrass), general residues, conventional crop straws (i.e. corn stover) and trees (willow). Second generation biofuels are therefore produced by cellulosic (also called lignocellulosic) feedstocks, i.e. non-food based feedstocks. These feedstocks usually contains cellulose, hemicellulose, and lignin. However, technologies using straws or agricultural and forestry waste do not appear to reach an economical suitability, because of the complexity of the energy intensive pretreatment processes that such a raw material requires [34]. Indeed, cellulose, hemicellulose and lignin are carbohydrates characterized by a complex structure [65]. Hence even if these carbohydrates do not have use as food, lignin require a large amount of thermo/mechanical energy to destroy its bonds and make simple sugars accessible to the fermentation reaction for the production of biofuel [66].

TABLE 1. LIST OF MOST COMMON FIRST GENERATION FEEDSTOCKS FOR BIOFUELS PRODUCTION.

BIOFUEL	CLASSIFICATION	FEEDSTOCK
Alcohols: ethanol and butanol	Produced by starch	Corn (main feedstock in USA)
		Milo
		Cassava
		Wheat
		Sorghum
		Sweet potato
	Produce by sugar in general	Sugar cane
		Sugar beets
		Sugar palm
		Sweet sorghum
		Nypa palm
Oil based biofuels: biodiesel, pure plant oil, biogas	Edible oil plants	Rapeseeds (main feedstock in Europe)
		Soy beans (main feedstock in USA)
		Oil palm
		Sunflower seed
		Coconut oil
		Mustard seed
	Non- edible oil plants	Jatropha
		Camelina
		Jojoba
		Karanj
		Castor beans
		Field pennycress
	Waste feedstock	Animal fat
		Waste vegetable oil

TABLE 2. TOP 10 COUNTRIES FOR ETHANOL PRODUCTION. 2010 DATA.

RANK	COUNTRY	MILLION LITRES
1	USA	51416
2	Brazil	26888
3	China	2699
4	France	1821
5	Canada	1495
6	India	1421
7	Poland	1079
8	Germany	917
9	Thailand	869
10	Jamaica	833

TABLE 3. TOP 10 COUNTRIES FOR BIODIESEL PRODUCTION. 2010 DATA.

RANK	COUNTRY	MILLION LITRES
1	USA	5912
2	Germany	5048
3	Spain	5023
4	Indonesia	4262
5	Brazil	4160
6	Malaysia	4091
7	China	3906
8	Argentina	3636
9	France	2926
10	Thailand	2771

The current state-of-the-art of companies all over the world is the construction of small facilities that could be expanded in the next few years [67]. One of the leading companies worldwide is Sapphire Energy (San Diego, USA), that, in matter of one of its project called Green Crude Farm, on its website reports:

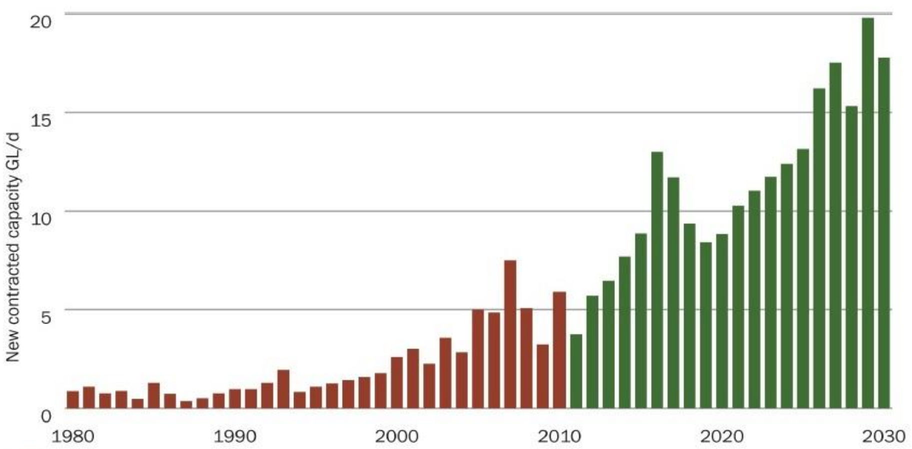
"As technology is proven and economies of scale are achieved at the IABR (Integrated Algal Biorefinery), the design and construction of the first commercial biorefinery will start in 2015, and by 2018, Sapphire aims to produce 5,000 barrels per day of green crude leading to million of gallons of renewable, domestically-produced liquid transportation fuels." [10]

In this thesis data available from literature are used. Since most of them come from experiments on laboratory scale, they were adapted with the aim to simulate and model an autonomous biorefinery, characterized by an entire cycle, from the cultivation of microalgae, until the production of biofuels. Thanks to their biological features, the biorefinery could produce both ethanol and biodiesel (see paragraph2.3).

1.4 DESALINATION IN THE WORLD: THE STATE-OF-THE-ART

The second type of plant to couple with a NPP is a desalination plant. This paragraph will explain the reasons that suggested this choice. Together with pollution and depletion of hydrocarbon resources, water scarcity is one of the most serious global challenges of our time. The challenge of providing ample and safe drinking water is further complicated by population growth, industrialization, contamination of available freshwater resources, and climate change. These motivations suggested the birth of desalination technology in 1960s and 1970s. In some countries, especially in the Middle East, desalination is no longer a marginal or supplemental water resource. For example, Qatar and Kuwait rely 100% on desalinated water for domestic and industrial supplies [68]. Kuwait was the first state to adopt seawater desalination, linking electricity

FIGURE 5. FORECASTED PRODUCTION CAPACITY FOR DESALINATED WATER [47].



generation to desalination plants. Kuwait began desalinated water production in 1957, when 3.1 million m^3 were produced per year. Saudi Arabia entered the field of desalinating water much later than Kuwait (in 1970), but it is currently the global leader for production of desalinated water.

Presently, the total global desalination capacity is around 66,5 million m^3/d [69] and it is expected to further increase in the next years, reaching about 100 million m^3/d by 2015 [70], that corresponds twice the rate of global water production by desalination in 2008. Indeed, desalination capacity is continuously increasing worldwide, not only in the Middle East and the North Africa (MENA) region, where water demand is high and other sources of supply are limited, but also in countries where desalination was unthinkable in the past, such as in Spain and Australia. Rising in desalinated capacity is well captured in Figure 5. The increase of desalination capacity is primarily caused not only by increases in water demand but also by a significant reduction in desalination cost as a result of substantial technological advances that contributed to make desalinated water cost-competitive with other water sources. However, desalination has great development potential on a global scale. This is attributed to the fact that, out of the 71 largest cities in the world that do not have local access to fresh water sources, 42 are located in coastal regions [68]. Out of the entire world population, 2400 million inhabitants representing 39% of the total, live at a distance of less than 100 km from the sea [71].

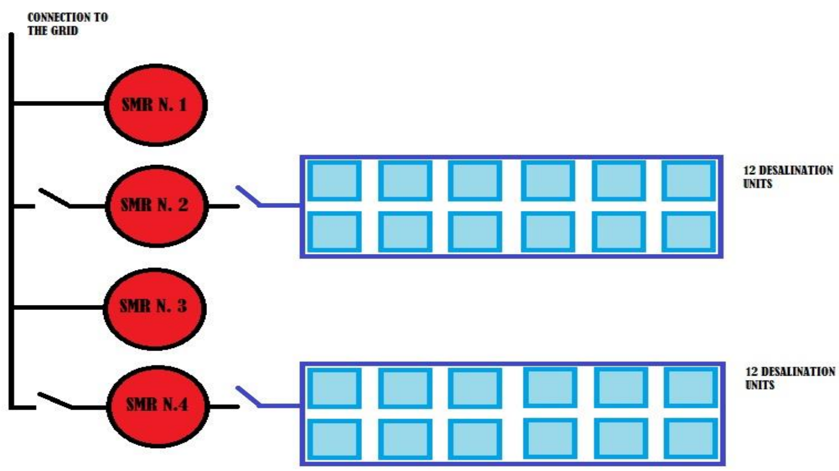
Other than the fact desalination may be the only option for some countries, there are some driving forces behind its potential development, making it more favourable than conventional resource development. Being independent of climatic conditions, rainfall and so on, a primary force is its identification as a secure source of supply. In addition, desalinated seawater has an essentially unlimited capacity, not subject to sustainability criteria, although perhaps limited only by energy requirements [70]

FIGURE 6. THE MARAFIQ PLANT IN SAUDI ARABIA. IN THE PICTURE, CLOSER TO THE SEA IT IS POSSIBLE TO SEE ALL THE 27 MODULES THAT FORM THE DESALINATION PLANT, AND BEYOND THEM, THE POWER PLANT.

Desalination processes could be classed essentially in two groups: thermal processes and membrane processes (paragraph 3.2). In history, the first developed technology was the thermal one. It consists in causing evaporation of water and subsequent re-condensation of steam, once it is salt-free. In the last few decades, this technology was overtaken by the membrane technology, that is based on the separation of salts using their physical-chemical properties. Nowadays, about 50% of the total desalination investments are for Seawater Reverse Osmosis (SWRO) projects due mainly to its lower overnight costs and total water costs compared to other conventional processes. Thermal processes will also continue to be utilized especially where energy is available at low-cost, but the tendency is that Multi-Effect Desalination (MED) will replace Multi-Stage Flash desalination (MSF) in future projects and could even compete with SWRO where raw water is highly polluted or of very high salinity (like Arabian Gulf seawater), or a higher quality of final water is needed. Thermal processes will remain in the market also because they have been widely accepted worldwide, they have a proven record of reliability, also they are dependable and have the potential for cogeneration of power and water (hybrid systems). Moreover, many efforts have been done to prove and confirm a good suitability of a cogeneration between a nuclear power plant and a desalination facility [2] [43] [72]. Thermal processes require mostly thermal energy (low temperature steam), while membrane facilities require only electricity. The purpose of this work is to direct energy produced by NPP in the desalination plant. Using the thermal energy is therefore more convenient because it would not require the conversion in electrical energy, losing about two thirds of the power, because of net efficiency. For this motivation, as discussed in section 3.2.4, in this thesis a MED-TVC plant has been chosen for cogeneration with the nuclear plant.

Currently, the capacity of many commissioned plants exceeds 400000 m^3/d, with the largest plant of the world (Ras Azzour) that is going to produce 1,034 million m^3/d, with a hybrid layout [69]. The largest size of plant that is already in operation produces 8/8,5 hundred thousand cubic meters per day. The Marafiq plant is an example: it is currently the second largest plant operating in the world, but it is the biggest MED-TVC plant. It was built in 2009, and it is characterized by a cogenerative layout with a conventional power plant (with both gas and steam turbine). It is composed by 27 units producing 30 m^3/d each, for a total capacity of 800000 m^3/d (Figure 6).

FIGURE 7. AN EXAMPLE OF POSSIBLE LAYOUT FOR THE COGENERATIVE PLANT. THE SMRS AND THE DESALINATION UNITS CAN BE COUPLED IN DIFFERENT FLEXIBLE ARRANGEMENTS.



Moreover, this modularization of the desalination plant could be an ulterior attractive feature to give a higher flexibility degree to our cogenerative plant. Indeed, having many smaller units allows to couple various desalination units with each corresponding nuclear modular unit, in a very flexible layout.

For example Figure 7 features 4 SMRs: 2 of them are only connected to the grid and always produce electricity. The other 2 are connected even to 24 desalination units, 12 each. They can switch the connection from either producing electricity and/or desalinated water. The possibility to have different single unit size and different whole site size, permits to couple the plants with very flexible layouts.

1.5 A NEW ECONOMICAL VALUATION TOOL: REAL OPTION ANALYSIS

1.5.1 Uncertainties and risks in the energetic market

Broadly, two of the most clear characteristics of markets, especially the energetic one, are the unpredictability and the volatility. Both of them confer to the financial world a large source of uncertainty, that in turn creates a risk for the investment.

In the energetic field, possible sources of uncertainties are the price of the electricity, the price of fuels (i.e. coal, gas, uranium), overnights costs of the plants, construction times, etc. Most of them not only depend on unpredictable financial speculations, but also on unpredictable environmental accident or political scenarios. For example, an unexpected accident like the one of Fukushima can change the energetic policy of some countries (i.e. Germany, Italy, Swiss); similarly critical political scenarios that usually characterize Middle Eastern countries contribute to increase the volatility of fossil fuels prices. At the same time, also the overnight costs are always subjected to great uncertainty: indeed, even if a technology could be already proven, the raw materials to produce it (i.e. steel) could became more expensive. Whereas they can be classified as "global" sources of uncertainty, some others can belong to a "local" class of uncertainty: for example the construction time is a very flexible parameter, and frequently is subjected to delay. They can be caused by accident on the construction site, mistakes during the installation, issues regarding licensing, delay in shipping, etc. Having a delay means waiting for receiving revenues, although at the same time the investor is paying the labour, debts, installation, etc. Moreover, also environmental uncertainties can be gathered in the "local" group. For instance they are represented by the country in which the investment is going to be introduced, regarding with the technology proposed: to install a CCGT in Italy is definitely less economically hazardous in comparison with, for example, the installation of the first SMR site. This is because the confidence, experience and acceptance of the technology, can create an uncomfortable environmental background.

From the financial point of view, uncertainty means risk. When risky events occur, an investment evaluated profitable can suddenly become less worthwhile. Consequently, the cost of Equity and the cost of Debt increase for risky projects, because the parts that finance the project require a higher return. In other words, the higher the risk and the higher the returns that investors will expect. Moreover, the higher the cost of capital, the higher the discount rate (WACC in this study).

For this reason the future cash flows are discounted with a more demanding discount factor, and to reach the break-even point become more difficult.

Therefore, trying to predict the flexibility of markets, is a mandatory point for a complete and accurate economic analysis for any investment.

1.5.2 Approaches for economic analysis

Usually in capital budgeting most of the investment analysis is sorted out using the Discounted Cash Flow (DCF) method, that is a methodology focused on the time value of money. This method is better explained in paragraph 5.1. As an introduction, it is important to specify that there are several DCF models, that range from simple to more sophisticated, but they all are based on the same foundation that simply involves calculation of the Net Present Value (NPV) of a project over the entire life cycle, accounting for the investment costs and the production phase free cash flows. Because of the time value of money, each cash flow from the future (future value, FV) is converted into today's value (present value, PV). The project NPV is simply the summation of the PVs of all the cash inflows and cash outflows. Then the optimal investment rule is to proceed with a project if its NPV is greater than zero and, in case of a portfolio with two or more different projects, the priority will be given to that one with the larger NPV [73]. A project can be evaluated even with the Internal Rate of Return (IRR): this profitability indicator is the discount rate at which the NPV becomes equal to zero. The greater the IRR of a project, the more attractive the investment.

A fundamental parameter used in NPV evaluation is the discount rate: it is the rate that is used to convert the future value of the project cash flow to the present value. It is adjusted depending on the perceived risk associated with the project. As said in the previous section, business is basically about taking risks. The higher the risk, the higher the returns investors will expect. With DCF methods, the only way to adjust the analysis to the risks and uncertainties that surround the project is to adapt the future cash flows with an appropriate discount rate, that however remain a fixed, established, deterministic value. Therefore, the choice of the discount rate is a very delicate step.

Hence, the DCF methodology, although it is simple and easy to implement, presents three substantial criticalities:

1. Choice of the discount rate
2. Weak consideration of stochastic nature of the cash flows: DCF can't capture any market uncertainties, like electricity and fuel prices, or technical uncertainties, like construction costs that vary considerably along construction time [36]
3. Assumed passivity of the management, unable to improve the results after the resolution of some uncertainties [37].

To overcome these problems, capturing uncertainties, volatility and flexibility, in the recent decades a new valuation approach was developed. It is based on giving a value to the managerial ability of inserting various modifications or an important drastic choice during the life of a project. It means that the manager can choose a better solution for the company during the period of the investment. For example, thinking about an investment in the energetic field, it could be reasonable to reckon that the construction of a nuclear power plant can be profitable (and therefore start) only if the price of the electricity, in the particular country the investor has chosen, rises over an established value (say 0,11 $/kWh) [74]. If the scenario becomes more (or less) attractive, the manager can decide to build (or abandon) the plant. The meaning of this choice is that ***the manager has***

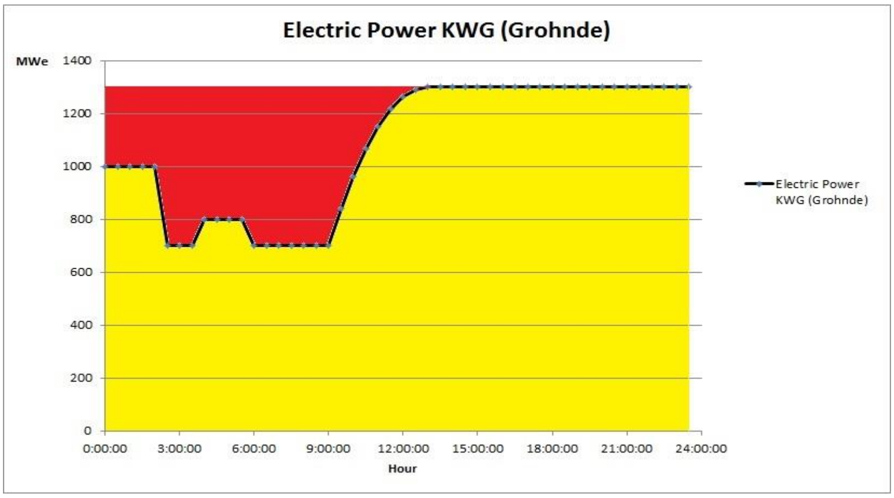


FIGURE 8. APPROXIMATION OF THE CURVE CONCERNING ENERGY PRODUCTION OF KWG PLANT IN GRONHDE (YELLOW LINE OF FIGURE 4). THE RED AREA IS THE AMOUNT OF POWER AVAILABLE FOR AUXILIARY PLANTS.

an option. This option is the right, but not the obligation, to take a managerial decision.

Thanks to this similitude, in the academic world researchers started to suggest evaluations models based on financial options, born in the 1970s. In finance, for the owner of the option, the option itself represents the right, but not the obligation, to proceed in the contracted transition (see paragraph 5.2.1): basically the shareholder has the option to buy or to sell stock. Prices involved in this trade are established by the financial markets. Similarly, in the real world the investor has some options, as said in the previous example. Because they refer to the real world, they are called *Real Options*.

Studying an investment with the Real Options Approach (ROA) gives to the investment itself a higher value. For example, calculating the NPV with DCF and ROA methods for the same project, the NPV evaluated with ROA will be higher than that one calculated with DCF. The difference between values is the worth of the option itself: it means that if the manager has a managerial option about a project, from the financial point of view this option has a value.

The extra value given by the option is due to the capability of the ROA to solve some uncertainties, to take advantage on the development of the scenario and to take a decision considering the evolving environmental conditions. Therefore the main difference between DCF and ROA is that the latter considers uncertainties and risks in a dynamic, stochastic, more interactive way. ROA is not going to take over the DCF methods, but it is a tool to complete the limits of the DCF models (section 5.2.2). In addition, ROA models are usually more sophisticated. As for the DCF method, several ROA models exist, and the investor can chose the most appropriate for his/her case (paragraph 5.3).

In this study we analyse the profitability of the plants proposed, with Real Options Approach, using two algorithms based on Monte Carlo Simulations (MCS). It allows to consider also the volatility and the stochastic nature of prices and costs.

1.6 RESEARCH OBJECTIVES AND QUESTIONS

From the operational point of view, SMRs are very flexible, thanks to their modularity. They are also feasible for many applications:

- Provide electricity in remote locations without advanced infrastructures and for small electric grids
- Provide electricity in large site, being clustered together in a single multi-module, large capacity site
- Supply auxiliary thermal end electric energy for non-electric applications, replacing the existing fossil fuels boiler.

Therefore, the aim of this work is to suggest an alternative to work with a nuclear power plant in load following mode. For not wasting fuel, for not creating thermo-mechanical stresses in the primary circuit and especially for taking full advantage of the huge invested capital costs, the reactor core always works and supplies full nominal designed power. The idea is to supply energy also to auxiliary non-electric plants, when there is an excess of produced thermal energy in comparison with the electric demand of the grid. Having n-nuclear modules, it is possible to modulate the electricity produced, switching the steam between the turbines of the secondary loops of the SMRs and the auxiliary plants.

Here we want to plan a daily production of electrical energy, based on different power level. For example, regarding the Figure 8, the yellow area is the energy that the SMR has to supply to the grid, whereas the red area is the energy produced "in excess". Consequently during the night the steam is switched to the auxiliary plants (biorefinery and/or desalination) in order to follow the grid, producing worthwhile co-products. Hence we are going to reply to the following research questions:

- Is it possible to couple a biorefinery in a flexible layout in order to work in load following?
- Is it possible to couple a desalination plant in a flexible layout in order to work in load following?
- Do these cogenerations add an extra value to the investment?
- What is the more profitable level of energy (or utilization factor) to address to the auxiliary plant?
- Does Real Option Approach capture the flexibility of the markets and of such a layout?

CHAPTER 2.

MICROALGAE AS BIOFUEL FEEDSTOCK

In the second chapter, the technology for the production of biofuels is introduced. Such a plant, called Biorefinery, is one of the auxiliary plant that is investigated as a non-electric application to connect to the nuclear site, with the goal of following the grid. Between all the possible biomass, the attention was moved on microalgae. They represent an innovative primary feedstock, because of their many biological and environmental advantages, that could lead to a proper significant development of biofuels in the near future. To understand all the advantages, limitations and scenarios evaluated, a deep section is inserted in this chapter about the characteristics of this raw material from the biological point of view. Subsequently, the main technologies and steps to obtain biofuels are investigated. Because of the modernity of this kind of technology, currently there is not a well-defined unique procedure to produce biodiesel and/or bioethanol from microalgae. Simultaneously, there is not a defined development of commercial-scale facilities. For these reasons many possibilities, technologies and processes to obtain the final products are deeply analysed in this chapter and in the Appendix A. In the last paragraph studied scenarios in matter of different layouts/parameters for the implementation of the model of a biorefinery are introduced.

2.1 WHY ALGAE?

In 2012 the annual world primary energy consumption was estimated at 12479 million tonnes of oil equivalent (mtoe). Fossil fuels accounted for 86,9% of the primary energy consumption, with oil (33,1% of share), coal (29,9%) and natural gas (23,9%) as the major fuels, while nuclear energy, hydroelectricity and other renewable accounted for 4,5%, 6,7%, and 4,7% respectively of the total primary energy consumption [55]. Hydrocarbons are being used for the production of fuel, electricity and other goods [75]. Excessive consumption of fossil fuels, particularly in large urban areas, has resulted in generation of high levels of pollution during the last decades. Fossil fuels are the largest contributor of greenhouse gases (GHGs) to the biosphere [56]. Furthermore, it is well known that their combustion release in the air a sizable quantity of particulate, i.e. PM 5 or PM 10, and nitrate oxides (NO_x) [57]. These factors contribute to damage the life of populations, especially in that developing countries in which legislation is still not clear and focused on these modern issues. In Figure 10 a glaring example of not monitored situation, in China. Moreover, with the expansion of human population and increase of industrial prosperity, especially in new world-powers (i.e. China and India), global energy consumption also has increased gradually, even if the recent economic crisis has slow down the growth. By 2030, CO_2 emission from road transport in China are expected to have increased by a factor of 3.4, whilst in India they are predicted to have raised by a factor of 5.8 [58]. Additionally, transport fuels, such as diesel and gasoline, are affected by limited reserves of fossil fuels, and the annual global oil production will begin to decline within the near future [59].

In this scenario, renewable sources might serve as an alternative. Nuclear, wind, water, sun, biomass and geothermal heat can be the renewable sources for the energy industry, whereas fuel production and the chemical industry may depend on biomass as an alternative source in the nearby future [76]. All petroleum based fuels can be replaced by renewable biomass fuels such as bioethanol, biodiesel, biomethane, biohydrogen [77]; they are derived for example from sugarcane, corn, switchgrass, algae [34], canola, forestry wastes [78], potato [79], etc. Countries across the globe have considered and directed state policies toward the increased and economic utilization of biomass, for meeting their future energy demands in order to meet carbon dioxide 5.2% reduction from

FIGURE 9. SHARING OF THE WORLDWIDE PRIMARY ENERGY CONSUMPTION IN 2012.

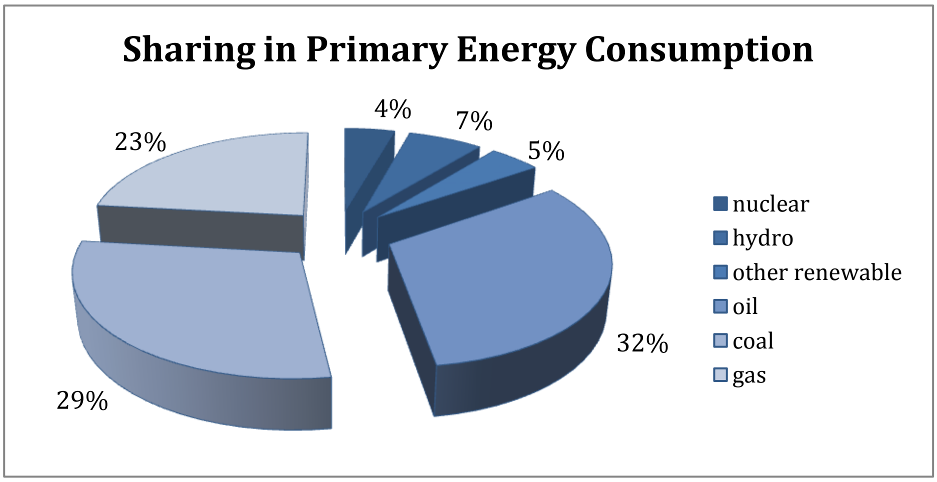


FIGURE 10. POLLUTION IN BEIJING. PICTURE FROM BBC.NEWS WEBSITE "BEIJING SMOG: WHEN GROWTH TRUMPS LIFE IN CHINA" BY MARTIN PATIENCE, PUBLISHED THE 27TH JANUARY 2013 [239].

1990 values targets, as specified in the Kyoto Protocol of 1997 [60] as well as to decrease reliance and dependence on the supply of fossil fuels.

The focus on Climate Change in Europe that led to the 20-20-20 targets is an example. These targets require countries to decrease GHG emissions, increase electricity generation from renewable, and decrease whole consumption by 20%. All EU-15 countries intend to include biomass in their portfolios to meet these targets by the year 2020 [22].

Three main paths are been suggested to reach this ambitious target: increasing energy efficiency, increasing use of clean fossil energy (i.e. use of fossil fuels coupled with CO_2 separation from flue gases and injection into underground reservoir for gradual release) and increasing use of renewable energy [80].

For all these reasons, in the last few decades we have assisted to a growing interest for biofuels, i.e. fuels that can be produced or directly extracted from biomass. Looking at the environment, the main advantage of biofuels is to fix the CO2 of the air during the cultivation phase (known in literature as *well to pump*) of the biomass and to release a smaller amount of GHGs during the combustion phase in the engine (*pump to wheel*), thanks to a different quality and composition of the biofuels themselves in comparison with the conventional fossil fuels used so far. First generation biofuels which have now attained economic levels of production, have been mainly extracted from food and oil crops including rapeseed oil, sugarcane, sugar beet and corn [81], as well as vegetable oils and animal fats using conventional technology [82]. It is projected that the growth in production and consumption of liquid biofuels will continue [83], but their impacts towards meeting the overall energy demands in the transport sector will remain limited due to: competition with food and fibre production for the use of arable land, regionally constrained market structures, lack of well managed agricultural practices in emerging economies, high water and fertiliser requirements, and a need for conservation of bio-diversity [84]. Other main constraints of this kind of crops are the slow growth and low photosynthetic efficiency [12]. Typically, the use of first generation biofuels has generated a lot of controversy, mainly due to their impact on global food markets and on food security, especially with regards to the most vulnerable regions of the world economy. This has raised pertinent questions on their potential to replace fossil fuels and sustainability of their production [64]. For example, apart from the risk that higher food prices may have severe negative implications on food security, the demand for biofuels could place substantial additional pressure on the natural resource base, with potentially harmful environmental and social consequences. Currently, about 1%, 17 million hectares according to [63], of the world's available arable land is used for the production of biofuels, providing 1% of global transport fuels. Clearly, increasing that share to anywhere near 100% is impractical owing to the severe impact on the world's food supply and the large areas of production land required [85]. The advent of second generation biofuels was intended to produce fuels from the whole plant matter of dedicated crops or agricultural residues, forest harvesting residues or wood processing waste [64], rather than from food crops. However, the technology for conversion in the most part has not reached the scales for commercial exploitation which has so far inhibited any significant exploitation [81]. Furthermore, technologies using straws or agricultural and forestry waste, called lignocellulosic feedstock, seem don't reaching an economical suitability, because of the complexity of the energy intensive pretreatment processes that such a raw materials require [34]. Indeed, the lignocellulosic biomass is made up from complex carbohydrate polymers: cellulose, hemicellulose and lignin [65]. Even if these carbohydrates do not have use as food, lignin require a large amount of thermo/mechanical energy to destroy its bonds and make simple sugars accessible to the fermentation reaction for the production of biofuel, bioethanol in such a case [66].

Conditions for a technically and economically viable biofuel resource are that [86]: it should be competitive or cost less than petroleum fuels; should require low to no additional land use; should enable air quality improvement (e.g. CO_2 sequestration) and should require minimal water use. Judicious exploitation of microalgae could meet these conditions and therefore make a significant contribution to meet the primary energy demand, while simultaneously providing environmental benefits [60].

In addition, because of their variety of high-value products, microalgae can become an important resource for a large number of biotechnology areas, including cosmetics, pharmaceuticals, nutrition and food additives.

Nowadays, large-scale microalgal culture as source for renewable energy is getting more attention [15] because of its advantages compared to terrestrial oil crops with respect to its high growing rate, low land requirement, possible high oil content (30–60 wt.%), and the opportunity to develop a completely closed algae-to-biofuel cycle [20]. In Figure 11 there is a comparison of oil yields between some typical biomasses. As a consequence, there has been a recent resurgence of interest and a proliferation of algal biofuel projects. At the same time, studies are published in literature discussing the energy balance for these processes to determine whether more energy is required for the growing and harvesting of algae than the algae can release or not. In Table 4 it is possible to find a list of advantages of this organism, even in comparison with crops of first generation biofuels and waste of second generation: the benefits can be gathered in two groups, environmental and/or industrial.

In particular, from the environmental point of view, microalgae have many further advantages: in addition to the already mentioned conveniences for low land and water requirements, they also have a high photosynthetic efficiency and a great capability with regards to CO_2 fixation. In fact, like all the plants growing in autotrophic condition, microalgae use the CO_2 in the air like a source of carbon; but in contrast with the usual crops, the optimum level of carbon dioxide concentration in the environment is highly superior for algae. This permits to absorb a higher quantity of GHG and for this reason algae have an important

Crop	Oil yield (lit/ha)
Rubber seed	80–120
Corn	172
Soybean	446
Safflower	779
Chinese tallow	907
Camelina	915
Sunflower	952
Peanut	1059
Canola	1190
Rapeseed	1190
Castor	1413
Jatropha	1892
Karanj	2590
Coconut	2689
Oil palm	5950
Microalgae (30% oil by wt.)	58,700
Microalgae (70% oil by wt.)	136,900

FIGURE 11. MICROALGAE HAVE THE POTENTIAL TO PRODUCE BIOFUELS WITH HIGH YIELDS [86]

potential in the entire GHG balance of the overall cycle, from the cultivation of the feedstock to the burning of biofuels in engines. Furthermore, many authors suggest to connect an algae cultivation system with the flue gas coming from a conventional power plant (i.e. coal or gas). With regards to CO_2 market, this create an ulterior economic benefit both to the owner of the power plant and to the owner of the algae cultivation system. Moreover, as explained better in paragraph 2.2, the main nutrients of such plants are simply phosphorus (P) and nitrogen (N), as well as CO_2.

Consequently, thanks to the great adaptability of microalgae to many different and incongruous environments, several studies have confirmed the possibility of growing in wastewater of urban area or factories, with the advantage to clean and treat the wastewater itself, thanks to the peculiarity of feeding, and therefore reducing, the quantity of the main pollutants. In the matter of industrial advantages, the biological composition itself represent a notable feature. The high lipid and carbohydrate contents permit to reach high yield of biofuels even with relative small amount of land (see Figure 11). Additionally, algae can reach higher level of concentration in the batch, especially in which are called “fermenter tanks”. High yields can be obtained from fast growth and a short cultivation-harvesting cycle, as well. Finally the biodiversity within this organisms and, more in general, the existence of a wide number of algal strains, offer the opportunity to choose the microalga and the technology more suitable for any specific climate and market conditions.

On the other hand, the current technology of microalgae results in a more expensive final product than the biofuel produced with the cycles obtained with first generation biomass. In [77] the price of the oil from microalgae is reported to be 2.4 US$ against 0.6-0.8 US$ of the oil from more straight plants. A more pessimistic scenario is drawn by [26]: in that report, some layouts of cultivation were investigated, in two diverse Canadian weather conditions and with two different production periods during the year. The result of the report is a set of prices that range from 24.6 US$ per litre of oil in Photo Bio Reactors (PBRs) cultivation system in base case scenario, to 1.54 US$ per litre of oil in fermenter tanks cultivation system in best case scenario (higher concentration and higher lipid yield in the batch).

TABLE 4. LIST OF ADVANTAGES OF USING MICROALGAE AS FEEDSTOCK TO PRODUCE BIOFUELS.

Advantage	Kind of benefit[5]	Compare[6]	References[7]
High photosynthetic efficiency	E, I	I	[12], [13]
Fast growth	I	I	[14], [15], [16]
Short cultivation-harvesting cycle (only 3-10 days)	I	I	[17]
High starch content	I	general	[18], [19]
High lipid content	I	I, II	[15], [20], [13]
Possibility to produce many products and co-products[8]	I	I, II	[12], [21], [17]
Wastewater treatment	E	I, II	[22], [11]
CO_2 fixation[9]	E	I, II	[12], [23], [13]
Not compete with food	E	I	[24], [25], [26]
Not influence food prices	E	I	[24], [25]
Low water requirement	E, I	I	[27], [25]
Low land requirement (high biofuel yield per ha)[10]	E, I	I	[13], [25], [28]
Not arable land requirement	E, I	I	[29], [16], [21]
Easy adaptability to climate	I	I	[26], [30]
Growing in many different environment (water condition)	I	I	[21], [31]
Conventional pretreatment	I	II	[32], [33], [34]

As explained in the next paragraphs, there are many factors that affect the growing of microalgae and for this reason it is difficult to make a fair comparison with the previous reference. Despite that, even considering the more optimistic price of 1.54 US$ per litre, it is still more expensive than 0.88 US$ per liter of oil, that is the worth presumed for oil from canola, reported in the same reference. Although this could represent an insurmountable issue for the development of this selected biomass, in literature it is underlined that the main item cost is the energy supply, both thermal and electric. The energy requirements for the production phases are still too high to make this resource economically competitive. Consequently the prices of cultivated algae are high, as well as the prices of the final biofuels and intermediate product (such as oil). Therefore it is reasonable to think that in a layout in which energy is not bought from outside but it is directly produced in the connected nuclear plant, the prices would be drastically lower. It is an aim of this study try to find an acceptable price.

[5] It means if the advantage represents a benefit from the industrial (I) point of view (i.e. technical or economical advantage) or from the environmental (E) point of view (i.e. reducing global warming or more sustainability).

[6] This column refers in comparison to which other biofuel generation the benefits represent an advantages. "I" indicates the first generation of biomass, i.e. corn, rice, soy, etc; while "II" indicates the second generation, i.e. agricultural or forestry waste.

[7] In literature many other works about this features are published. For brevity, only some of them are included in the table.

[8] As detailed in 2.2, the composition of these organisms is very suitable for the production of biofuels. Actually the three main components of such plants are proteins, carbohydrates and lipids. This permits to produce bioethanol and/or biodiesel and/or biomethane. From the industrial point of view, it is a great benefit because it makes possible to choose the more convenient technology to produce the more convenient biofuel. Furthermore, the co-products of the processes, like protein-rich cakes or glycerine, add an additional worth to this production chain.

[9] Although also the cultivation of conventional crops involve CO_2 sequestration, microalgae require more carbon dioxide. The atmospheric CO_2 levels (about 0.0387% v/v) are not sufficient to support the high microalgae growth rates and productivities needed for industrial scale biofuel production. A solution to this issue could be using flue gas from close conventional power plants that supply waste gases from combustion processes with CO_2 levels above 15% (v/v) [24].

[10] Microalgae have the highest oil yield among various plant oils. They can produce up to 100000 l of oil per ha per year, whereas palm, coconut, castor and sunflower produce up to 5950, 2689, 1413 and 952 l per ha per year, respectively [236].

2.2 BIOLOGY OF MICROALGAE: WHAT ARE THEY AND WHAT AFFECTS THEIR GROWTH?

In this paragraph a brief presentation of the main biological features of microalgae is presented, together to the main nutrients that they require to grow and some of algal strains studied by scientific community. Algae are recognised as one of the oldest life-forms [87]. They are primitive plants (*thallophytes*), i.e. lacking roots, stems and leaves, have no sterile covering of cells around the reproductive cells and have chlorophyll a as their primary photosynthetic pigment [88]. Their simplicity results in a restricted diameter, from 2 to 20 μm [89]. Algae structures are primarily for energy conversion without any development beyond cells. Their simple structure allows them to adapt to prevailing environmental conditions and prosper in the long term [87]. Prokaryotic cells (cyanobacteria) lack membrane-bound organelles (plastids, mitochondria, nuclei, Golgi bodies, and flagella) and are more akin to bacteria rather than algae. Eukaryotic cells, which comprise of many different types of common algae, do have these organelles that control the functions of the cell, allowing it to survive and reproduce. Eukaryotes are categorised into a variety of classes mainly defined by their pigmentation, life cycle and basic cellular structure [90]. The most important classes are: green algae (*Chlorophyta*), red algae (*Rhodophyta*) and diatoms (*Bacillariophyta*). Algae can either be autotrophic or heterotrophic; the former require only inorganic compounds such as CO_2, salts and a light energy source for growth; while the latter is non-photosynthetic, therefore requires an external source of organic compounds as well as nutrients as an energy source. Some photosynthetic algae are mixotrophic, i.e. they have the ability to both perform photosynthesis and acquire exogenous organic nutrients [88]. For autotrophic algae, photosynthesis, summarized in equation (1), is a key process of their survival, whereby they convert solar radiation and CO_2, absorbed by chloroplasts, into adenosine triphosphate (ATP) and O_2 the usable energy currency at cellular level, which is then used in respiration to produce energy to support growth [91]:

$$6CO_2 + 12H_2O + (photons) \rightarrow C_6H_{12}O_6 + 6O_2 + 6H_2 \qquad (1)$$

Heterotrophic production has also been successfully used for algal biomass [92]. In this process microalgae are grown on organic carbon substrates, such as glucose, in stirred tank bioreactors or fermenters. In this case algae growth is independent of light energy.

The most decisive cultivation factors determining algal growth and product formation rates are quality and quantity of nutrients, the temperature of the environment, PH of the broth, and, for autotrophic growth, light supply (spectral range and photoperiod are crucial factors and have to be optimized for all microalgal species) and light intensity. Nutrients added to algal cultures must provide the inorganic elements that make up the algal cell and include also macronutrients, vitamins and trace elements. While there is very little published work on the optimal levels of nutrients required for mass algal cultures, the macronutrients required are generally considered to be nitrogen and phosphorus [93], at a ratio of 16N:1P [94]. Typical trace metals used include chelated salts of iron, zinc, cobalt, manganese, selenium and nickel [26]. Furthermore, in literature there is evidence that a medium characterized by starvation (or lower than optimal, however) condition of some nutrients, especially nitrogen, highly influences the yield of algae and their composition. In particular, nitrogen starvation has been studied intensely and general results confirm that the final yield of the biomass decreases, but the percentages of lipids and carbohydrates in the share of the whole composition increase. These nutrients also provide the best chemical condition of the cultivation batch, in matter of PH and salinity of the cultivation medium. In matter of temperature, in general algae growth increases exponentially with rising temperatures until an optimum level is reached, after which growth declines. This is particularly important for outdoor cultures, where the ability to control temperatures is often limited (especially in open systems) and is determined by atmospheric temperature, solar irradiance and humidity. While temperatures below the optimal range will generally not kill algae (except for freezing conditions) sustained temperatures above the optimal range will. Furthermore, higher temperatures during dark periods have been shown to increase biomass losses [95]. Thus, basically it is important for the culture to reach optimal temperatures quickly in the morning and to rapidly decrease temperatures after darkness, thereby maintaining high productivity during the day and minimizing biomass loss at night. Regarding the light intensity, it is of importance to avoid as well light limitation that results in so-called “dark reactions” of the cells by utilization of molecular oxygen, as photo-inhibition by excessive irradiation with photons that might even cause severe cell damage [96]. Therefore, salinity (ion strength and ionic composition of the cultivation medium), pH-value, turbulence and temperature are decisive for cellular growth and product formation. Typical values found in literature report temperature ranges of 16-27 °C, pH-values of 4-11, salinities of 12-40 g/L, and light intensities of 1000-10,000 lux [97].

Thus, the quantities and qualities of nutrients, temperature and light affect also the quantity and the quality of the cultivated microalgae. In particular, quantity means the yield of the cultivation batch, usually measured in metric tons per hectare per year, or in grams per square meter per day (for open systems) or in grams per litre per day (for closed systems). Instead, quality means the composition of microalgae. Indeed, their main components are lipids, carbohydrates and proteins. To produce biofuels, lipids are used to extract oil to process in biodiesel refinery, whereas carbohydrates are important as source of sugars (glucose) to ferment in bioethanol. Consequently, for the purpose of this work, it is important to reach as high as possible yields in lipids and/or carbohydrates. It is important to note that every single algal strain, growing in several different conditions, has a different yield and is characterised by a different composition. Therefore to work out a reasonable simulation many variables was taken into account: variables considered in this study are:

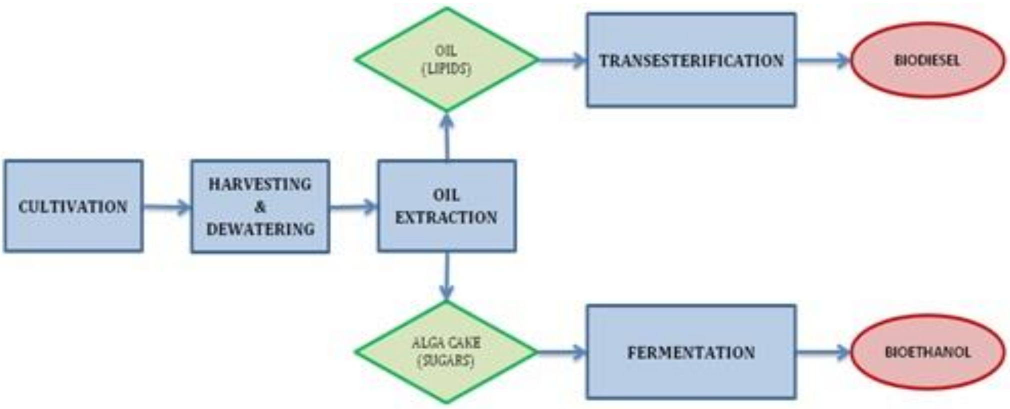
1. Algal strains
2. Weather conditions
3. Quantity of nutrients in the batch
4. Type of cultivation layout

All this factors, only one per time, were changed in comparison to a base case scenario, to confront if there is any optimal condition and how the change of the parameters could edit the final results. In the following paragraph, the entire route to produce biofuels will be investigated and the selection of the layout of the biorefinery used in this study will be integrated. While in the paragraph 2.4, it is explained which alga has been selected, in which weather condition, in which cultivation medium and why. Five scenarios has been evaluated and for every of them the main features, i.e. yield and composition of biomass, have been inserted in the model implemented, to simulate the energetic consumptions and yields of the main products of the biorefinery.

2.3 ROUTE TO BIOFUELS

The first question that has to be faced before a new technology is proposed to the market is whether the LCA of the whole process is positive or negative. A Life-Cycle Assessment (LCA) consists of making a comparison between all energy inputs, i.e. energy requirements and energy included in all material used (for raw materials, nutrients, buildings, etc.), and energy outputs, i.e. energy extracted from the final products. In this case, a positive LCA means that it's possible to obtain and extract more energy from the biofuels in comparison to the whole energy needed to produce them. Many authors have already delved into a complete LCA for production of biodiesel or bioethanol from algal feedstock, allocating also worth to the co-products of the chemical and technical processes. The opinion broadly is that the energetic output is hardly larger than the required fossil fuel inputs for the production processes from the microalgae [33]. Thanks to this assumption, in this study a LCA has not be implemented, but the attention has been focused only on all the energy requirements for all the production phases of the cycle, starting from the cultivation of microalgae up the production of biodiesel and bioethanol. This is a totally new approach, since a major conclusion from the literature review is that a complete energy balance for a closed microalgae to biofuel concept is not available [33]. Actually, some authors have proposed a study for the production of biodiesel only [89], [24], [98], without taking care of the algal cake (rich of carbohydrates) that results from the oil extraction phase; others have done the vice versa, producing bioethanol without using the lipid content of the residual biomass [99]; some others have studied only the cultivation phase [100]. On the contrary, as already discussed in the previous paragraph, the innovation of this study is to obtain both biodiesel (from lipids) and bioethanol (from carbohydrates), to maximize the exploitation of the biomass, describing every single process involved in the production chain (see Figure 12). Firstly the biomass is cultivated in one of the possible layouts. Then it is harvested, passing through different steps to lower the water content of the biomass itself. Subsequently, algae enter into an oil extraction process to recover the lipid content. Therefore the oil obtained enters the biodiesel production process, via a chemical reaction called transesterification; while on the other side the residual algal cake is fermented to obtain bioethanol. The feasibility to produce both the biofuels is already suggested and confirmed by [26], [9], [98].

FIGURE 12. THE SCHEME WITH ALL THE PROCESSES STUDIED AND SIMULATED IN THIS WORK FOR THE BIOREFINERY.



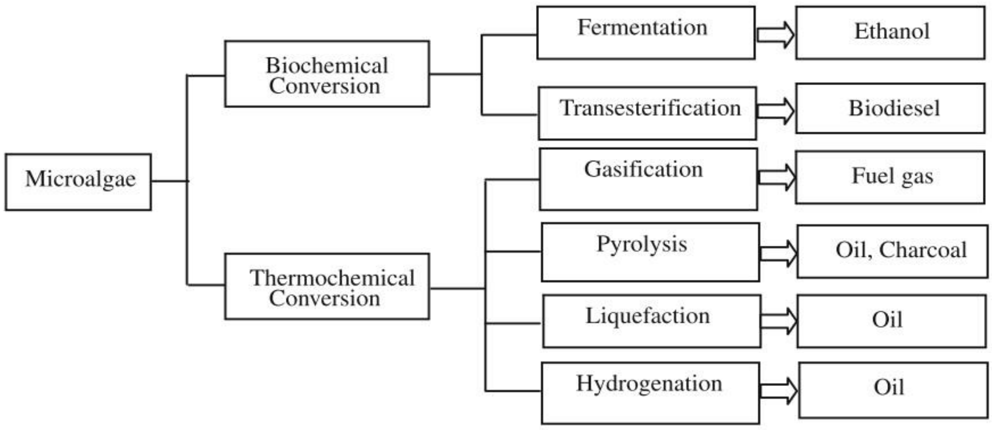


FIGURE 13. IN THE DIAGRAM REPORTED IN [12] IT IS POSSIBLE TO SEE THE TWO GROUPS THAT GATHER THE DIFFERENT PROCESSES TO OBTAIN BIOFUELS. ALGAE CAN BE USED TO PRODUCE DIFFERENT KINDS OF BIOFUELS, SUCH AS ETHANOL, BIODIESEL AND GAS.

For this purpose, data available from literature, laboratory experiments, current technologies and consultation with companies have been adopted for the model here developed. Since there is no current existing commercial scale facility, a model was developed using potentially scalable systems. This model, discussed in the following sections, consists in cultivation, harvesting, drying, oil extraction, chemical reactions, and product purification stages.

In general, the main chemical processes to produce biofuel from algae are gathered in two groups: thermochemical processes and biochemical processes (see Figure 13). Thermochemical processes involve very extreme temperature and/or pressure conditions. The main characteristics of these processes are:

- Gasification: temperature 800-900 °C and pressure 24 MPa [101]
- Pyrolysis: temperature 500 °C, absence of air [102]
- Liquefaction: temperature 340 °C and pressure 20 MPa [103]
- Hydrogenation: temperature 400-430 °C and pressure 7-14 MPa [104]

The steam of a conventional light water reactor can't reach these thermodynamic conditions. Then a SMR can't provide the heat required by such technologies. For this reason, here the thermochemical conversion processes will not be investigated further. Instead, the attention will be focused on biochemical methods. Therefore, in the following, transesterification and fermentation processes will be deeply investigated. The main issue has been faced is that this technology applied to microalgae is a very modern challenge and, consequently, commercial plants for biofuels production from microalgae do not exist currently [26], [9]. Nowadays, the only commercial facilities existing regarding algae use this biomass for other applications, like pharmaceutics, cosmetic or food. For this reason, most of available data are in matter of the cultivation phase, especially in open system layout. To tackle this problem, present technologies of production of biodiesel from soybean and of bioethanol from corn have been used as general guides. In fact, some points have to be clear:

- The production of biodiesel simply involve a chemical transformation (transesterification) of the oil extracted by the raw material. Once the oil has been extracted the method is identical for all the biomasses [89]
- Also, the oil extraction phase for microalgae is identical to the oil extraction from soybean [105]

- Therefore, the biodiesel production from soy technology can be used, simply adjusting the lipids content and yields of algal biomass
- The production of bioethanol simply involve a chemical transformation (fermentation) of the pre-treated sugars of the biomass. Once the sugars have been split in monomers of glucose, the method is identical for all the biomasses
- Also, starch is a very important component in microalgae composition and to divide their carbohydrates can be used the same enzymatic pre-treatment process employed for corn feedstock [9]. The only difference is the right enzymatic cocktail (and hence chemical conditions) that has to be applied to attack also the cellulose, that shares an important percentage in the microalgal carbohydrate composition, and not only the starch
- Therefore, the bioethanol production from corn technology can be used, simply adjusting the glucose content of biomass and yields of algal biomass itself.

In the Appendix A., the route describing all the involved processes to produce biofuels from algae is inserted. In that appendix, the processes, the energy requirements, the schemas, the technical choices, the yields and all the necessary to fully comprehend the modelled biorefinery are reported and deeply analysed. For brevity, all these explanations are omitted in this chapter.

2.4 SCENARIOS ESTIMATED

Except the cultivation, the route to biofuels modelled is only one. It is composed by the phases just discussed. The scenarios created have the aim to study the dependence of the final results from the cultivation phase, that could be subjected to the main fluctuations of values. As usual, considering a production chain, the variability of the outputs depends largely by the fluctuations of the beginning inputs, in comparison to the fluctuations of values of a process close to the end of the chain. Therefore, scenarios created take into account variable inputs.

Here are illustrated the main features chosen for each scenario to investigate the dependence of the biorefinery to some defined parameters. They are: the algal strain, the weather condition, the quantity of nutrients supplied in the medium and cultivation system. The influence of these parameters on the final outputs of the cultivation phase regards the yields of the biomass and its composition (i.e. mainly the percentage of lipids and carbohydrate) that in turn affect the yields of biofuels. For these variables, the following scenarios have been evaluated:

- Scenario 1 – *base case*
- Scenario 2 – *fermenter*
- Scenario 3 – *unfavourable climate*
- Scenario 4 – *lipid rich algal strain*
- Scenario 5 – *low N*

2.4.1 Algal strain and weather conditions

Chlorella is a green alga, growing in freshwater. It is often used as a representative specie to study the life-cycle assessments, the GHG emission balances, or the water footprint of microalgae biofuel production [31]. Because of the extent of data available in literature, the easy adaptability to most of weather conditions, good yield of biomass and useful composition of alga itself [106], in this study *Chlorella* strain has been selected as the benchmark used in scenario 1 (base case). In particular, *Chlorella Vulgaris* has been chosen for the open ponds cultivation systems (autotrophic growth condition), while *Chlorella Protothecoides* for the fermenter tanks system (heterotrophic growth condition) [80], analysed in scenario 2. Therefore, first of all the features of the base case have been determined. In scenario 1 *Chlorella Vulgaris* is grown in an open pond system, in a standard weather condition (not excessively favourable or unfavourable) and with standard quantities and qualities of nutrients supplied in the pond. Apart the selection of the strain, the most significant decisions to be taken are related to the values that have to be used in the model to represent the biology of the alga. Table 5 contains the values regarding the *Chlorella* alga cultivated in open ponds. Even if the correct average of the yield is 20 grams per square meter per day, here it has been assumed a more optimistic yield of 24.75 grams per square meter per day [89], gained in Mediterranean area. The choice is motivated by three main factors:

1. In Table 5 there are several data referred to yield obtained in countries with unfavourable data, like Canada and Denmark, that lower the average
2. Although this study consider an European scenario, a reflection has to be done: in this historical period, some countries has reached a saturation of the nuclear market and some others are planning to enter in it for the first time. Among these, it's reasonable to focus on rich growing countries willing to improve their power plants portfolio. For these motivations it is logical to think to consider all the Mediterranean areas and Arabic countries, such as Saudi Arabia and UAE. Actually they have recently shown a strong interest on the way to nuclear energy
3. Even if in some cases low yields are supposed, it's realistic to believe in an improvement of this very young application with some tricks. For example, genetically adjusting the strain or using industrial devices, i.e. heating the pond

For all these causes, a slightly positive value has been assumed for scenario 1. Whereas for scenario 3 (unfavourable weather) a yield closer to that reached in Canada or Denmark (with heated pond or anyway with a temperature not dramatically low) has been selected: 15.47 grams per square meter per day will be used for the analysis. In matter of composition, the same share between proteins, carbohydrates and lipids has been considered for scenarios 1 and 3. This share is very close to the average of the data. Only the amount of carbohydrates is rather positive because has been assumed that a careful selection of the inoculum[11] and the possibility of a future genetic modification of algae could help to reach the 50% in carbohydrate hypothesized in papers looking the Mediterranean area [33] and [89].

To make an accurate simulation and give an example of what has been assumed for favourable and unfavourable weather conditions, some cities are indicated. For favourable climate, weather of cities like Bordeaux and Bucharest

[11] The inoculum is the colony, the stock of the microalgae. In a cultivation plant, a laboratory of microbiology must be installed. Its aim is to take care of the original colony of microorganism, permit their breeding and to make inoculation into the cultivation systems. The inoculum must be inserted periodically in the cultivation medium.

are taken as benchmark. The average yearly temperature is 13°C, while the average yearly solar irradiance is 3,65 kWh/m^2d. Data were calculated very easily, taking the coordination (latitude and longitude) of the cities, and using a tool furnished by NASA website. [107]. Similarly, for unfavourable climate, weather of cities like Glasgow or Stockholm are taken as example. There, the average yearly temperature and average yearly solar irradiance are 7°C and 2,8 kWh/m^2d, respectively.

2.4.2 A different cultivation layout

As discussed in paragraph 2.3, there are only two interesting layouts where growing algae: raceway ponds and fermenter tanks. In scenario 2 typical data of a cultivation of algae in tanks, toward a heterotrophic process, are used. In fermentors is not possible to use an alga that grown autotrophically; for this reason a heterotrophic alga of the same specie was chosen, like *Chlorella Protothecoides*. In contrast to raceways and PBRs, the yield from fermenters is better expressed as grams per litre, per day of cultivation. The average of whole amount of data is 40 g/L, but excluding numbers belonging to study characterized by very small fermenters (laboratory size) or inefficient fermenter (simple batch), the average rise to 52,7 g/L. However, algae cell densities of up to 51g/L have been obtained in seven-day culture, and up to 116 g/L have been reported by other groups. While even higher cell concentrations of 302 g/L have been achieved by [108] in mixotrophic cultures, for the fermentation scenario discussed here, a yield of 51,2 g/L is conservatively assumed. It was calculated in [109] and hypothesized also in [26].

TABLE 5. A LIST OF DATA CONCERNING TO *CHLORELLA* OR *CHLORELLA VULGARIS*, CULTIVATED IN OPEN PONDS.

ALGAL STRAIN	PLACE	SOLAR IRRADIANCE [kWh/m²*d]	AVERAGE TEMPERATURE [°C]	YIELD [g/m2*d]	PROTEINS (%)	CARBOHYDRATES (%)	LIPIDS (%)	REFERENCE
Chlorella	Ballen, Denmark	2,91	8,6	9,5	-	-	20%	[22]
Chlorella	Ballen, Denmark (April-October)	4,177	13,9	15,3	-	-	20%	[22]
Chlorella	Trebon (Czech Republic)	-	-	-	-	55%	-	[18]
Chlorella	Prince George, Canada	3,28	1,5	9,38	25%	25%	15%	[26]
Chlorella (heated pond)	Prince George, Canada	3,28	1,5	15,47	25%	25%	15%	[26]
Chlorella	Prince George, Can.(April-September)	4,96	8,9	21,6	25%	25%	15%	[26]
Chlorella	Nanaimo, Canada	3,45	6,27	11,4	25%	25%	15%	[26]
Chlorella (heated pond)	Nanaimo, Canada	3,45	6,27	16	25%	25%	15%	[26]
Chlorella	Nanaimo, Canada (April-September)	5,12	12,06	22,89	25%	25%	15%	[26]
Chlorella	Asia	-	-	8,5-21	-	-	-	[110]
Chlorella	-	-	-	-	52,64%	10,62%	14,57%	[92]
STANDARD microalga	-	-	-	-	25,00%	25,00%	15,00%	[26]
Chlorella	Czech Republic (September)	-	-	11,1	-	-	-	[111]
Chlorella	Czech Republic (July)	-	-	23,5	-	-	-	[111]
Chlorella	Czech Republic (October)	-	-	18,1	-	-	-	[111]
Chlorella	Czech Republic (July)	-	-	32,2	-	-	-	[111]

Chlorella	New Mexico	-	-	21	-	-	-	[95]
Chlorella	-	-	-	-	51-58%	12-17 %	14-22 %	[112]
Chlorella	Mediterranean Area	-	-	13,7	29%	49,50%	19,70%	[33]
Chlorella	Mediterranean Area	-	-	-	-	-	-	[33]
Chlorella	Mediterranean Area	-	-	24,75	28,20%	49,50%	17,50%	[89]
Chlorella vulgaris	-	-	-	19	-	-	25%	[113]
STANDARD microalga	-	-	-	25	-	-	20,00%	[114]
OPTIMAL microalga	-	-	-	40	-	-	40,00%	[114]
Chlorella vulgaris	-	-	-	-	29%	51%	18%	[115]
Chlorella vulgaris	-	-	-	-	60,38%	12,16%	11,61%	[106]
Chlorella sp.	-	-	-	25	-	-	-	[116]
Standard	-	-	-	25	-	-	25%	[27]
				20	*36%*	*35%*	*20%*	*AVERAGE*[12]
				24,75	**29%**	**50%**	**20%**	**This study (base case)**
				15,47	**29%**	**50%**	**20%**	**This study (unfavourable climate)**

In this table it is possible to confront the yield obtained in the references and the composition of the alga selected. At the end of the table a weighted average of the values and the choice for this study for scenario 1 (base case) and scenario 3 (unfavourable climate).

[12] The average for the percentages of proteins, carbohydrates and lipids is done considering only once the values belonging to the same reference.

Table 6. List of data reported in literature concerning the cultivation of Chlorella Protothecoides in heterotrophic culture.

ALGAL STRAIN	FINAL CONCENTRATION (g/L)	CULTIVATION TIME	PROTEINS (%)	CARBOHYDRATES (%)	LIPIDS (%)	REFERENCE
Chlorella Protothecoides	15,5	184 h (7,7 days)	10,28%	15,43%	55,20%	[117]
Standard case	50	1 day	10%	20%	50%	[26]
Better case	-	-	-	-	60%	[26]
Chlorella Protothecoides (11000 L culture)	14,2	200 h (8,3 days)	-	-	44,3%	[118]
Chlorella Protothecoides (5 L culture)	15,5	-	-	-	46,1%	[118]
Chlorella Protothecoides (750 L culture)	12,8	-	-	-	48,7%	[118]
Chlorella Protothecoides (improved fed-batch)	51,2	167 h (7 days)	-	-	50,3%	[109]
Chlorella Protothecoides (primary fed-batch)	16,8	184 h (7,7 days)	-	-	55,2%	[109]
Chlorella Protothecoides (batch)	3,2	-	-	-	57,8%	[109]
Chlorella	116,2	-	-	-	-	[119]
Chlorella	104,9	-	-	-	-	[119]
	40	-	10,1%	17,7%	52%	AVERAGE
Chlorella Protothecoides	**51,2**	**167 h (7 days)**	**10,28%**	**15,43%**	**55,20%**	**This study**

Looking at this table, most of papers reports a lipid concentration of about 50%, while the yield in biomass is more spread. The latter strongly depend from the batch used, nutrients and size. As discussed in the text, a reasonable yield could be fixed at about 50 g/L in a 7 days culture.

In matter of composition, percentage of lipids is the most important number in a heterotrophic culture. Most of authors reports a concentration of about 50% in lipids, confirmed by the average of 52%. In this study we will use the more detailed composition supplied in [117] and showed in Figure 14.

2.4.3 A different algal strain

In the paragraph 2.4.1 *Chlorella* specie has been selected how benchmark for this study. *Chlorella* is reported to be a highly carbohydrate rich microalgae (50% in composition has been assumed), therefore useful for bioethanol production. An interesting investigation could be how the strain could modify the yield of final products and, as a result, the economic analysis. For scenario 4 a highly lipid rich microalga has been used to support the biodiesel production. In particular, *Botryococcus Braunii* is usually reported like a lipid rich microalgae in comparison with other strains [113] and for this reason has been inserted in our model. As it is confirmed by Table 7, the content in lipids has been fixed at 57%, while proteins have the 22% of the share and carbohydrate the 14% [120]. To make a fair comparison, for the *Botryococcus Braunii* has been assumed the same yield of the *Chlorella Vulgaris* in scenario 1 (24,75 grams per square meter per day).

Figure 14. The composition of Chlorella Protothecoides, grown in autotrophic cultivation (AC) and heterotrophic cultivation (HC) [188].

Component (%)	AC	HC
Protein	52.64 ± 0.26	10.28 ± 0.10
Lipid	14.57 ± 0.16	55.20 ± 0.28
Carbohydrate	10.62 ± 0.14	15.43 ± 0.17
Ash	6.36 ± 0.05	5.93 ± 0.04
Moisture	5.39 ± 0.04	1.96 ± 0.02
Others	10.42 ± 0.65	11.20 ± 0.61

Table 7. Lipid content in Botryococcus Braunii reported by some authors. An average of 57% is calculated and used in this study.

LIPID CONTENT	**REFERENCE**
64%	[11]
25-75%	[16]
64%	[121]
44,5%	[120]
25-75%	[122]
70%	[15]
57%	AVERAGE
57%	**This study**

2.4.4 A different nutrients quantity

The last scenario identified is the case in which the alga is grown in a condition of nitrogen starvation. Though also phosphorous starvation has been reported in literature, nitrogen starvation is definitely more studied by many authors. For this reason in this thesis the latter has been investigated. In general, in case of starvation of nutrients, the organisms slow down the reproduction, i.e. decrease the yield, but enhance the content of energy in their cells, i.e. increase the contents of lipids and carbohydrates. This results in a higher quality of alga for our purpose, because of lipids and carbohydrates increase, while proteins decrease. Thanks to Table 8, similarly to previous cases, has been evaluated the yield and the composition of *Chlorella Vulgaris* for scenario 5. According to [89] the yield has been fixed at 19,25 g/m^2d, while according to [115] the composition closer to the average chosen is: 7% of proteins, 55% of carbohydrates and 40% of lipids. Furthermore, also the net calorific value of algae increases. Actually lipids are the most energetic component of an organism. In Table 9 it's possible to make an assessment of the lower heating values (LHVs) of algae cultivated in normal condition and in nitrogen starvation condition. In the former case, with a lower lipid content, the LHV is 17,5 MJ/kg, while in the latter, with a higher quality, it increase to 22,6 MJ/kg. LHVs are calculated with data available in [89], to supply an example.

Finally, summary tables with numbers, values and parameters used for all the five scenarios evaluated are included in section 4.1.1. In that paragraph are inserted also the results of the analysis in term of energetic demand and yields of products. Also comments and discussion of results are in 4.1.2.

TABLE 8. IN CASE OF NUTRIENTS STARVATION THE WHOLE YIELD OF THE BIOMASS DECREASES, BUT THE PERCENTAGES OF ENERGETIC COMPONENTS, LIKE CARBOHYDRATES AND LIPIDS, INCREASE.

ALGAL STRAIN	STANDARD NITROGEN				STARVATION NITROGEN				REFERENCE
	YIELD [g/m2*d]	PROTEINS (%)	CARBOHYDRATES (%)	LIPIDS (%)	YIELD [g/m2*d]	PROTEINS (%)	CARBOHYDRATES (%)	LIPIDS (%)	
Chlorella	13,7	29%	49,50%	19,70%	10,5	6%	51%	43%	[33]
Chlorella	-	-	-	-	19,2	6%	51%	43%	[33]
Chlorella	24,75	28,2%	49,5%	17,5%	19,25	6,7%	52,9%	38,5%	[89]
Chlorella vulgaris	19	-	-	25%	-	-	-	42%	[123]
STANDARD microalga	25	-	-	20,00%	-	-	-	40%	[114]
OPTIMAL microalga	40	-	-	40,00%	-	-	-	60%	[114]
Chlorella vulgaris	-	29%	51%	18%		7%	55%	40%	[115]
Chlorella vulgaris	-	-	-	-	-	-	44%	-	[106]
Chlorella vulgaris	-	60,38%	12,16%	11,61%	-	21,09%	51,3%	19,03%	[106]
	24,49	*37%*	*41%*	*22%*	*16,32*	*9%*	*51%*	*41%*	*AVERAGE*
					19,25	**7%**	**55%**	**40%**	**This study**

TABLE 9. LIPIDS HAVE GOT THE HIGHEST NET CALORIFIC VALUE. THEREFORE THE QUALITY OF MICROALGA INCREASES IN THE CASE OF STARVATION OF NITROGEN [89].

BIOLOGICAL COMPONENT	NET CALORIFIC VALUE [MJ/kg]			LOWER HEATING VALUE [MJ/kg]
Protein	15,5		Standard N	17,5
Carbohydrate	13		Starvation N	22,6
Lipid	38,3			

CHAPTER 3.

DESALINATION TECHNOLOGY

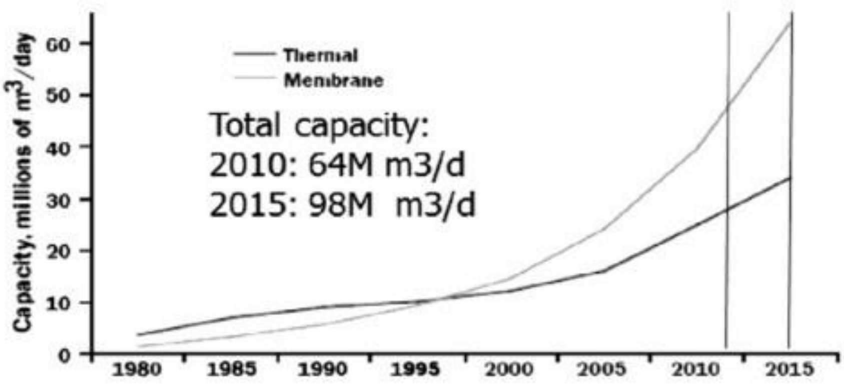
In the third chapter, the technology of second auxiliary plant to couple with the NPP is introduced. In particular at the beginning, the chapter contains a brief historical review of the desalination technology. In the following the main processes to produced desalinated water are discussed, together with the chosen plant for this dissertation. Then a literature review is presented in order to choose the values to use in this work in matter of costs and power requirements.

3.1 DESALINATION IN NUMBERS: HISTORICAL REVIEW

Together with pollution and depletion of hydrocarbon resources, water scarcity is one of the most serious global challenges of our time. Presently, over one-third of the world's population lives in water-stressed countries and by 2025, this figure is predicted to rise to nearly two-thirds [124]. The challenge of providing ample and safe drinking water is further complicated by population growth, industrialization, contamination of available freshwater resources, and climate change. These motivations suggested the birth of desalination technology in 1960s and 1970s, and moreover caused a strong development of such a technology in the last few decades. In some countries, desalination is no longer a marginal or supplemental water resource. For example, Qatar and Kuwait rely 100% on desalinated water for domestic and industrial supplies [68]. Kuwait was the first state to adopt seawater desalination, linking electricity generation to desalination plants. Kuwait began desalinated water production in 1957, when 3.1 million m^3 were produced per year. Saudi Arabia entered the desalinated water field much later than Kuwait. The first plant was commissioned in 1970. It has, however, gone in for an ambitious program of desalination plants construction on both the Red Sea and Gulf coasts. The "Saline Water Conversion Corporation" promoted 30 desalination plant projects by the end of the 1980s [125]. Saudi Arabia is currently the world leader for production of desalinated water. Presently, the total global desalination capacity is around 66,5 million m^3/d [69] and it is expected to further increase in the next years, reaching about 100 million m^3/d by 2015 [70], that corresponds twice the rate of global water production by desalination in 2008. The growth of desalination capacity worldwide is shown in Figure 15. Two main considerations can be done. Firstly, the growth rate is increasing very rapidly, especially in the last decade. It is currently about 55% per year [35]. Secondly, in late 1990s the membrane processes overtook the thermal one as the main technology to produce desalinated water. An explanations of such a processes will be done in paragraph 3.2.

The increase of desalination capacity is caused primarily not only by increases in water demand but also by a significant reduction in desalination cost as a result of substantial technological advances that contributed to make desalinated water cost-competitive with other water sources (see Figure 16). In some specific areas, desalination has now been able to successfully compete with conventional water resources and water transfers for potable water supply (e.g., construction of dams and reservoirs or canal transfers) [71].

FIGURE 15. THE GROWTH RATE OF INSTALLED CAPACITY WORLDWIDE [48].



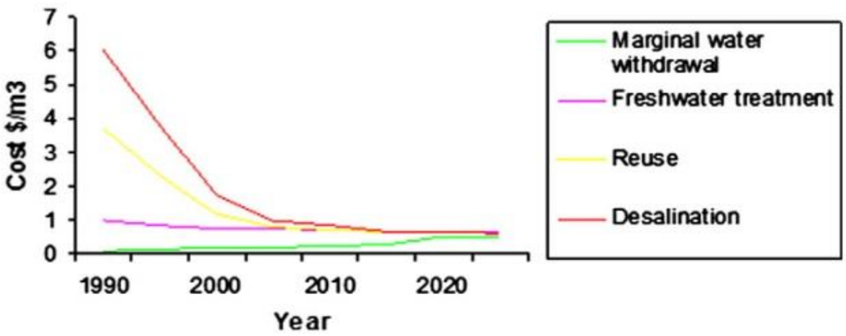


FIGURE 16. THE GRAPH SHOWS THAT THE TREND OF COSTS OF PRODUCING A DESALINATION WATER IS DECREASING. IT IS REACHING THE SAME LEVEL OF COST OF OTHER SOURCE OF WATER [48].

Desalination has great development potential on a global scale. This is attributed to the fact that, out of 71 largest cities that do not have local access to new fresh water sources, 42 are located along coasts [68]. Out of the entire world population, 2400 million inhabitants representing 39% of the total, live at a distance of less than 100 km from the sea [71].

Other than the fact desalination may be the only option for some countries, there are some driving forces behind its development potential, making it more favourable than conventional resource development. Being independent of climatic conditions, rainfall and so on, a primary force is its identification as a secure source of supply. In addition, desalinated seawater has an essentially unlimited capacity, not subject to sustainability criteria, although perhaps limited only by energy requirements [70]. Indeed, desalination capacity is continuously increasing worldwide, not only in the Middle East and North Africa (MENA) region where water demand is high and other sources of supply are limited, but also in countries where desalination was unthinkable in the past, such as in Spain and Australia (both contain arid and semiarid lands).

In Table 10 and in Figure 17 are listed the top 10 countries for already installed capacity and for future developing of the market, respectively. Making a comparison it is possible to recognize that probably Kuwait has already saturated his market, while countries like Israel and India have showed a new interest in this market. As a confirm to these graphs and statistics, it is expected that the total desalination market will reach over US$31 billion by 2015 [126]. About 50% of the total desalination investments are for Seawater Reverse Osmosis (SWRO) projects due mainly to its lower overnight costs and total water costs compared to other conventional processes. Thermal processes will also continue to be utilized especially where energy is available at low-cost, but the tendency is that Multi-Effect Desalination (MED) will replace Multi-Stage Flash desalination (MSF) in future projects and could even compete with SWRO where raw water is highly polluted or of very high salinity (like Arabian Gulf seawater). Thermal processes will remain in the market because they have been widely accepted in the Arabian Gulf area. They have a proven record of reliability, are dependable and have the potential for cogeneration of power and water (hybrid systems). Moreover, many efforts have been done to prove and confirm a good suitability of a cogeneration between a nuclear power plant and a desalination facility [2] [43] [72].

In 2011, there were 15988 plants operating in 150 countries worldwide. Currently, the capacity of many commissioned plants exceeds 400000 m^3/d; the largest plant of the world (Ras Azzour) is going to produce 1,034 million m^3/d, with a hybrid layout [69]. The Figure 18 list the largest plants built or contracted.

TABLE 10. LEFT. TOP 10 DESALINATION COUNTRIES AT JUNE 30, 2008. SOURCE: "GLOBAL WATER INTELLIGENCE" AND "IDA".
FIGURE 17. RIGHT. TOP 10 DESALINATION MARKETS [69].

COUNTRY	TOTAL CAPACITY [m^3/d]	SHARE IN WORLD CAPACITY
1) Saudi Arabia	10,759,693	17%
2) UAE	8,428,456	13%
3) USA	8,133,415	13%
4) Spain	5,249,536	8%
5) Kuwait	2,876,625	5%
6) Algeria	2,675,958	4%
7) China	2,259,741	4%
8) Qatar	1,712,886	3%
9) Japan	1,493,158	2%
10) Australia	1,184,812	2%

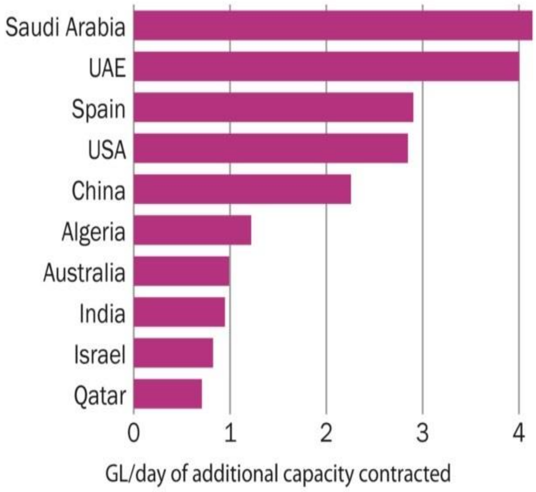
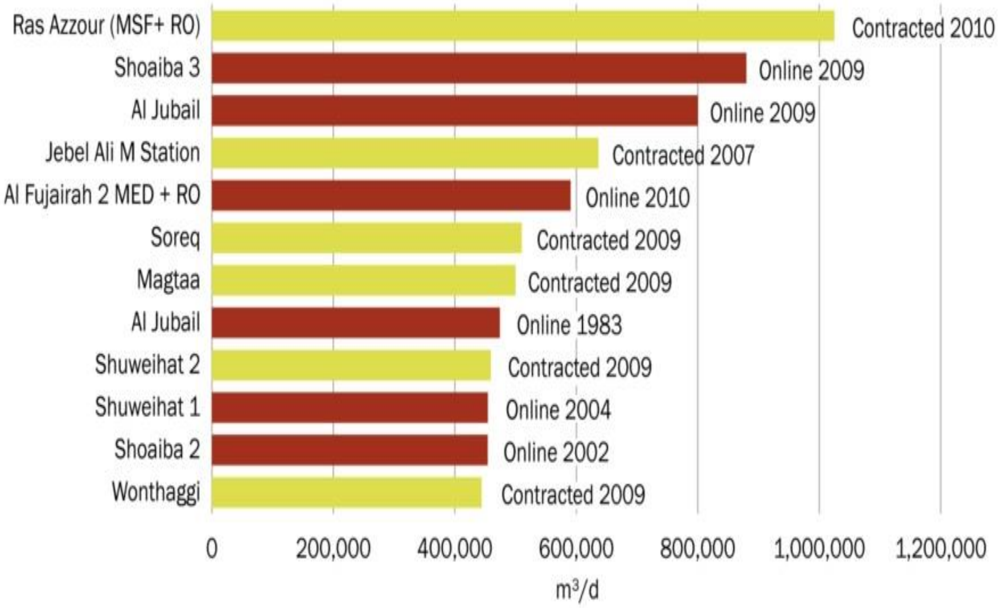


FIGURE 18. THE TREND OF THE LAST YEAR HAS BEEN TO INCREASE THE SIZE OF THE PLANTS DESIGNED TO TAKE ADVANTAGE OF ECONOMIES OF SCALE. BROADLY THE BIGGEST PLANTS USE MEMBRANE PROCESS OR HAVE A HYBRID LAYOUT.



3.2 DESALINATION PROCESSES

Of the global desalted water, 63,7% of the total capacity is produced by membrane processes and 34,2% by thermal processes. The desalination source water is split with about 58,9% from seawater and 21,2% from brackish groundwater sources, and the remaining percentage from surface water and saline wastewater [69]. These figures are constantly changing because the desalination market is growing very rapidly. However it is possible to assert that RO is prevalent outside of Middle East, and that nowadays there is a big attention for hybrid plants, combining thermal and membrane processes. In this paragraph the main processes for a desalination plant are explained, together with an economical and technical comparison, with attention to modern issues and improvements.

The aim of a desalination facility is the separation of salt water in two streams: one with a low concentration of dissolves salts (called "fresh water stream"), the other one containing the remaining part of dissolved (called "brine stream"). The thermodynamic efficiency of a distillation plant is expressed by the ratio:

$$GOR = \frac{kg\ of\ H_2O\ produced}{kg\ of\ steam\ used}$$

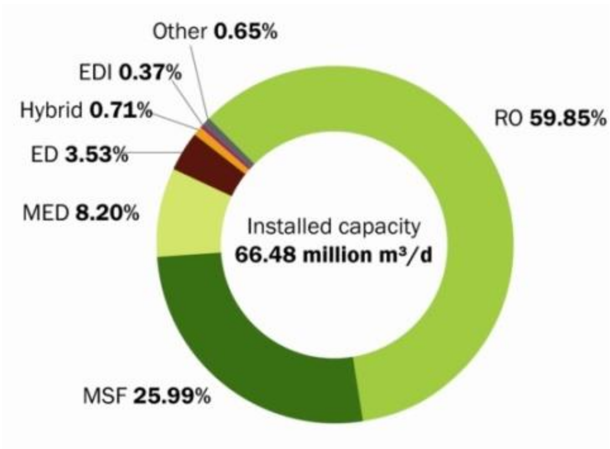
where GOR means "Gain-Output Ratio". Hence GOR is a measure of how much thermal energy is consumed in the process. Typically the value of GOR ranges from 1 to 10 kg/kg, even if recent improved technologies often reach 12 or above. For multiple effect (ME) systems the GOR is directly related to the number of effects (e.g. GOR= 0.8 n, where n is the number of effects). More effects directly increases GOR and for systems using thermo-compressors the GOR is also impacted by the pressure of the steam. Higher pressure steam will recycle more process vapour within the ME part of the process thereby improving the GOR and reducing external enthalpy requirements. These and other details are discussed here in the following.

Several desalination technologies exist and are industrially used but, in every case, the process requires energy. As revealed in advance, desalination processes can be divided in two sets: the thermal and the membrane processes.

3.2.1 Thermal Processes

Thermal processes are distillation processes, where water is heated to the boiling point to produce the maximum amount of water vapour. Steam is generated through distillation where the condensate is "fresh water". To do this economically in a desalination plant, the pressure of water that has to be boiled is adjusted to control the boiling point: a pressure decrease means a lower boiling

FIGURE 19. MEMBRANE PROCESSES ARE THE MAIN TECHNOLOGY TO PRODUCE FRESHWATER CURRENTLY.

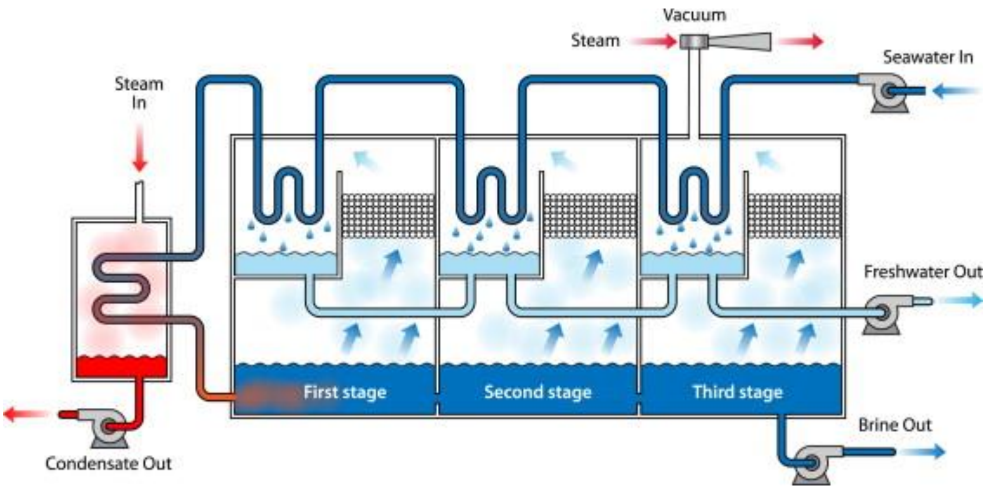


temperature. When water is heated to its boiling point and then the heat is turned off, water will continue to boil only for a short time because additional energy (the heat of vaporization) is needed to permit boiling. Once the water stops boiling, boiling can be renewed by either adding more heat or by reducing the pressure of the ambient in which water is located. Actually, if the ambient pressure were reduced, the water would be at a temperature above its boiling point and would flash to produce vapour. If more vapour can be produced and then condensed into fresh water with the same amount of heat, the process tends to be more efficient. To significantly reduce the amount of energy needed for vaporization, the distillation desalting process usually uses multiple boiling in successive vessels, each operating at a lower temperature and pressure. Thus, for example 8 tons of distillate can be produced from 1 ton of steam. Aside from multiple boiling, the other important factor is scale control. Although most substances dissolve more readily in warmer water, some dissolve more readily in cooler water. Some of these substances, like carbonates and sulphates, are found in seawater. One of the most important is calcium sulphate ($CaSO_4$), which begins to leave solution when sea water approaches about 115 °C. This material forms a hard scale that coats any tubes or surfaces present. Scale creates thermal and mechanical problems and, once formed, is difficult to remove. To avoid the formation of this scale, besides the addition of special chemicals to the sea water that reduce scale precipitation, maximum temperature of the process must be limited.

MULTI-STAGE FLASH DESALINATION (MSF)

In the MSF process, seawater is heated in a vessel called the brine heater (the red vessel in Figure 20). This is generally done by condensing steam on a bank of tubes that carry seawater which passes through the vessel. This heated seawater then flows into another vessel, called a stage, where the ambient pressure is lower, causing the water to immediately boil. The portion of introduced total water that evaporates depends on the stage pressure. The vapour steam generated by flashing is converted to fresh water by being condensed on tubes of heat exchangers that run through each stage. The tubes are cooled by the incoming feed water going to the brine heater. This, in turn, warms up the feed water so that the amount of thermal energy needed in the brine heater to raise the temperature of the seawater is reduced. Typically, an MSF plant can contain from 15 to 25 stages. Adding stages increases the total surface area, thus increases the capital cost in addition to the complexity of operation.

FIGURE 20. SCHEMA OF A MSF DESALINATION PLANT.

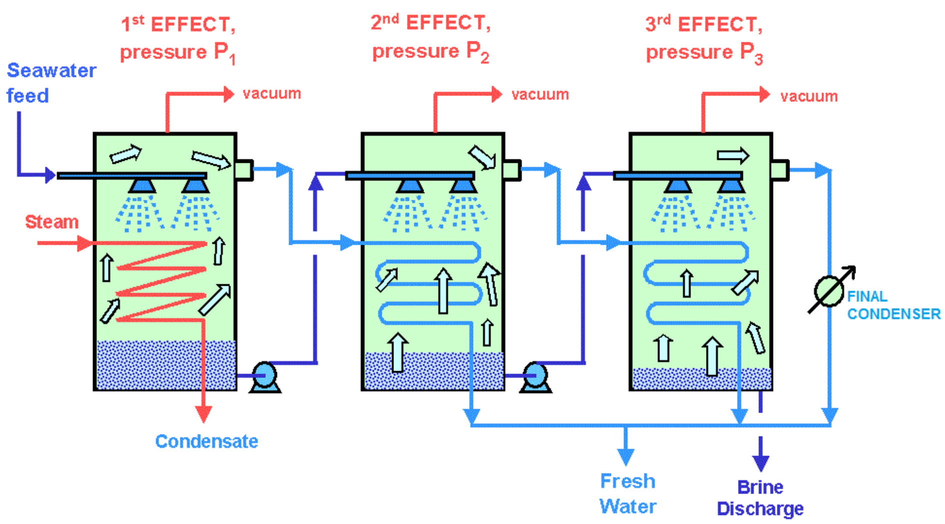


The MSF plants usually operate at the top brine temperatures (TBT) after the brine heater of 90-110°C. One of the factors that affect the thermal efficiency of the plant is the difference between the temperature of the brine heater exit and the temperature in the last stage on the cold end of the plant. Operating with a high TBT (110-130°C) increases the efficiency, but it also increases the potential for detrimental scale formation and accelerated corrosion of metal surfaces. Many countries on the Arabian Peninsula, such as Saudi Arabia, the United Arab Emirates, and Kuwait, are highly dependent on MSF facilities to supply water to their urban areas. This dependence, combined with a large installed capacity, has encouraged them to take measures to protect this investment. The water authorities in these countries have invested funds to increase plant reliability and this leaded to improvements in scale control, materials of construction, automation and controls. In addition, increases in the size of the basic unit have produced economies of scale in capital costs.

MULTI-EFFECT DESALINATION (MED)

MED process has been used for industrial distillation for a long time. Traditional uses for this process are the evaporation of juice from sugar cane in the production of sugar and the production of salt with the evaporative process. MED, like MSF, takes place in a series of vessels (stages or effects) and uses the principles of condensation and evaporation at reduced ambient pressure in the various effects. This permits the seawater feed to undergo boiling without the need to supply additional heat after the first effect. In general, an effect consists of a vessel, a heat exchanger, and devices for transporting the various fluids between the effects. Diverse designs have been or are being used for the heat exchanger area, such as vertical tubes with falling brine film or rising liquids, horizontal tubes with falling film, or plates with a falling brine film. By far the most common heat exchanger consists of horizontal tubes with a falling film. There are several methods of adding the feed water to the system. Adding feed water in equal portions to the various effects is the most common. The feed water is sprayed or otherwise distributed onto the surface of the evaporator surface (usually tubes) in a thin film to promote rapid boiling and evaporation after it has been preheated to the boiling temperature on the upper section. The surfaces in the first effect, are heated by steam from steam turbines of the power plants or a boiler. The steam is then condensed on the colder heat transfer surface inside the effect.

FIGURE 21. SCHEMA OF A MED DESALINATION PLANT.

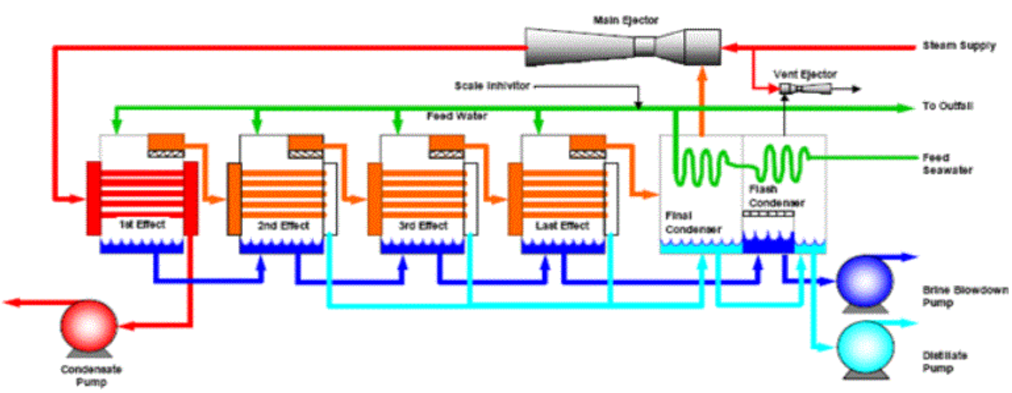


The condensate is recycled to the boiler for reuse. The surfaces of all the other effects are heated by the steam produced in each preceding effect. The steam produced in the last effect is condensed in a separate heat exchanger called the final condenser (on the right of Figure 21), which is cooled by the incoming sea water, thus preheating the feed water. Only a portion of the seawater applied to the heat transfer surfaces is evaporated. The remaining feed water, of each effect, now concentrated and called brine, is often fed to the brine pool of the next effect, where some of it flashes into steam. This steam is also part of the heating process. All steam condensed inside the effects is the source of the fresh water product. The ambient pressure in the various effects in the MED process is maintained by a separate vacuum system. The thermal efficiency of the process depends on the number of effects with 8 to 16 effects being found in a typical plant. MED plants are typically built in units of 2,000 to 30,000 m^3/d. Two different configurations are industrially used: the LT-HTFE (Low Temperature Horizontal Tube Film Evaporator) and the HT-VTE (High Temperature Vertical Tube Evaporator). In LT-HTME the tubes are arranged horizontally and the evaporation of the brine occurs outside the tube bundles by spraying the brine over them creating a thin film from which steam evaporates, while in HT-VTE evaporation takes place inside vertical tubes. LT-HTME plants are more recent and their TBT is limited to 70°C, and this significantly reduces the risk of scaling. At the same time, larger heat exchange surfaces are required. Most of the more recent applications for the MED plants have been in India, the Caribbean, the Canary Islands and the United Arab Emirates.

VAPOUR COMPRESSION (VC)

The vapour compression (VC) distillation process is generally used in combination with other processes (like the MED) and by itself for small and medium scale seawater desalting applications. The heat for evaporating the water comes from the compression of vapour rather than the direct exchange of heat from steam produced in a boiler. Steam ejectors (thermal vapour compression, TVC) and mechanical compressors (mechanical vapour compression, MVC) are used in the compression cycle to run the process. All steam is removed by the compressor/ejector from the last effect and introduced as heating steam into the first effect after compression where it condenses on the cold side of the heat transfer surface. Seawater is sprayed, or otherwise distributed on the other side of the heat transfer surface where it boils. In order to use low cost compressors, the pressure increase is limited, and therefore, most smaller plants only have one stage. In newer and larger plants, several stages are used. The mechanical VC units are produced in capacities ranging from a few litres up to 3000 m^3/d.

FIGURE 22. SCHEMA OF MED-TVC DESALINATION PLANT.



MSF	MED	VC
-Process and individual equipment designs optimization -Typical unit size increased from 19,000 to 90,000 m^3/d -Increase of brine chamber load -Improved plant reliability, thermal performance (Performance Ratio above 9) -TBT[a] gradually increased from 90 to 112 °C and now to 120 °C -Materials of construction and structural aspect improved (low cost materials) -Reduction in the installation costs despite the increased cost of raw materials and labor costs -Use of hybrid systems (thermal–thermal and thermal–other technologies such as MEE-TVC, MSF-RO, MSF-MEE[b] and air cooling systems) and addition of membrane softening of feed water	-Typical unit size increased from 3800 to 22,700 m^3/d. Scaling problems reduced by proper condenser tube bundle design and evaporating saline water distribution -Improved heat transfer coefficients with proper heat transfer surfaces -Stabilization of low temperature operation and use of Aluminum for heat transfer surfaces -TBT[a] is maintained at 70 °C -Use of TVC[c] with MED	-VC is generally integrated with MED plants -They are generally used for small or medium scale applications -Two types of VC (TVC[c] and MVC[d]) -Typical unit size increased by 5 tim now about 3800 m^3/d. Energy consumption is reduced; it is less th 6 kWh/m^3

[a] TBT: top brine temperature. [b] MEE: Multi Effect Evaporation. [c] TVC: thermal vapor compression. [d] MVC: mechanical vapor compression.

FIGURE 23. IMPROVEMENTS, CURRENT STATE-OF-THE-ART, COMPARISON BETWEEN THERMAL PROCESSES FOR PRODUCTION OF DESALTED WATER.

They generally have an energy consumption of about 7 to 12 kWh/m^3. With the steam-jet type VC unit, also called a thermo-compressor, an ejector operated using 3 to 20 bar motive steam removes part of the water vapour from the vessel. In the ejector, the removed vapour is compressed to the necessary heating steam pressure to be introduced into the first effect. VC units are often used for resorts, industries, and drilling sites where fresh water is not readily available. Their simplicity and reliability of operation make them an attractive unit for small installations where these factors are desired.

VC is often coupled with MED plants. In this case steam ejectors are used, that remove vapour from the last effect and compress it. Steam coming from a turbine or a boiler is first used to compress part of the last effect vapour and then the two flows are sent to the first effect as heating source. Recycling low pressure steam from the last effect through thermo-compression allows a decrease of external steam required. Almost all recent MED plants are coupled with TVC. This combination is used when motive steam is available at higher pressure and temperature than the ones required in the first effect. Besides the increase of efficiency, TVC processes are inexpensive and durable as they do not have any moving parts. Finally, many other advances were recently done to improve thermal processes. They are summarized in Figure 23 [70].

3.2.2 Membrane Processes

ELECTRODIALYSIS (ED)

ED was commercially introduced in the early 1960s, about 10 years before reverse osmosis. ED depends on the following general principles:

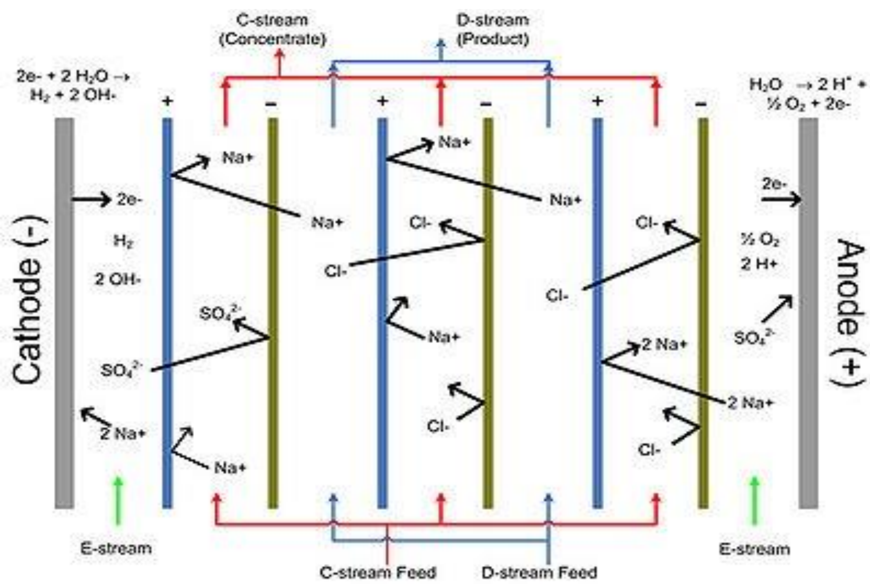
- Most salts dissolved in water are ionic. They could be both positively (cations) or negatively (anions) charged.
- These ions migrate toward the electrodes with an opposite electric charge.
- Membranes can be constructed to permit selective passage of either anions or cations.

The dissolved ionic constituents in a saline solution, such as chloride (-) sodium (+), calcium (2+), and carbonate (–), are dispersed in water, effectively neutralizing their individual charges. In Figure 24 it is drawn a schema of the process. When electrodes are connected to an outside source of direct current like a battery and placed in a container of saline water, electrical current is carried through the solution, with the ions tending to migrate to the electrode with the opposite charge.

Membranes are placed between the two electrodes. These membranes are arranged alternately, with an anion-selective membrane (only anions can pass through it) followed by a cation-selective membrane (only cations can pass through it). A spacer sheet (called a cell) that permits water to flow along the face of the membrane is placed between each pair of. One cell is fed with saline water while the next one is fed with recycled brine. As electrodes are charged and saline water is fed the anions (such as chloride or carbonate) in the water are attracted and diverted through the membrane towards the positive electrode. This dilutes the salt content of the water in the product water channel. The anions (-) pass through the anion-selective membrane, but cannot pass any farther than the cation-selective membrane, which blocks their path and traps the anions in the brine stream. Similarly, cations (such as sodium and calcium) under the influence of the negative electrode move in the opposite direction in comparison to anions, through the cation-selective membrane to the concentrate channel on the other side. Here, the cations (+) are trapped because the next membrane is anion-selective and prevents further movement towards the electrode. Therefore, there will be cells from which the ions have migrated (the dilute cells for product water) and others in which the ions concentrate (the concentrate cell for the brine stream).

The basic ED unit consists of several hundred cells. The raw feed water must be pre-treated to prevent materials that could harm the membranes or clog the narrow channels in the cells from entering the membrane stack. A post treatment process is also required for stabilization of the water and preparing it for distribution. This post-treatment consists mainly of removing gases such as hydrogen sulphide and adjusting the pH.

FIGURE 24. THE SCHEMA OF ELECTRODIALYSIS PROCESS.



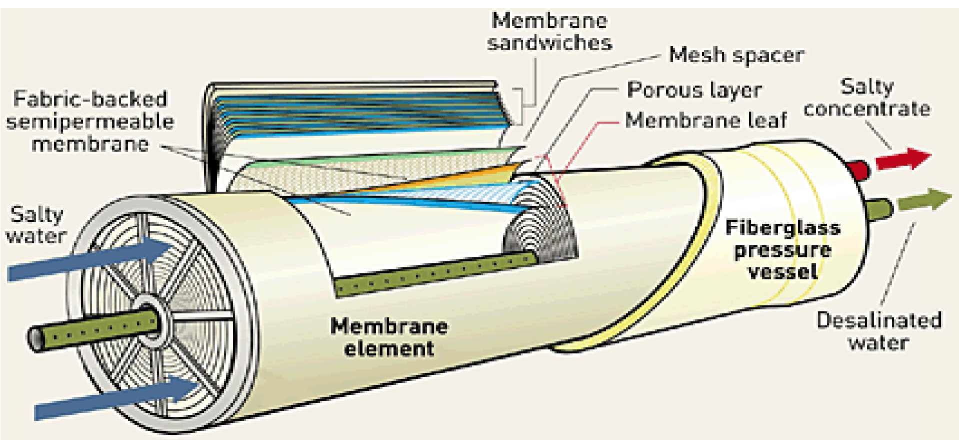
REVERSE OSMOSIS (RO)

In comparison to distillation and electrodialysis, RO is relatively new, with successful commercialization occurring in the early 1970s. RO is a membrane separation process in which the water from a pressurized saline solution is separated from the solutes (the dissolved material) by flowing through a membrane. No heating or phase change is necessary for this separation. The major energy required to desalt is for pressurizing the feed water. The saline feed water is pumped into a closed vessel where it is pressurized against the membrane. As a portion of the water passes through the membrane, the remaining feed water increases in salt content. At the same time, a portion of this feed water is discharged without passing through the membrane. Without this controller discharge, the pressurized feed water would continue to increase in salt concentration creating problems such as precipitation of super-saturated salts and increased osmotic pressure across the membranes, and this would limit the water flow through the membrane. The process is driven by the following equation:

$$Q_a = K_a * (P - \Pi) * \frac{A}{e} \qquad (2)$$

where Q_a is the water mass flow through the membrane; K_a is the membrane permeability coefficient that depends on membrane material, thickness and temperature; P is pressure difference between the two sides of membrane; Π is the osmotic pressure; A is the membrane surface; finally e is its thickness. As it is clear from equation (2), water must be pumped to a pressure higher than the osmotic pressure to pass through the membrane. These pressure values can reach the 80 bar for sea water. The membrane must be able to withstand the entire pressure drop across it. The two most commercially successful membrane configurations are spiral-wound and hollow fibre modules. A spiral-wound module element consists of two membrane sheets supported by a grooved or porous sheet that provides the pressure support for the membrane as well as providing the flow path for the product water. Each sheet is sealed along three of its edges, while the fourth is attached to a central product discharge tube. A plastic spacer sheet is located on each side of the membrane assembly sheets in order to provide the flow channels for the feed flow. The entire assembly is then spirally wrapped around the central discharge tube forming a compact RO module element. The recovery ratio (permeate flow rate divided by the feed flow rate) of

FIGURE 25. REPRESENTATION OF RO PROCESS. CASE OF SPIRAL-WOUND MEMBRANE.



this membrane configuration is very low so that up to 7 elements are arranged in one module to get a higher overall recovery ratio. Spiral-wound membranes have simple design and reasonable production costs and, despite this, offer a relatively high resistance to fouling. Hollow fibre membranes are made of hair-like fibres (external diameter 85-200 μm) which are united in U-tube bundles and arranged in pressure vessels. The feed is introduced along a central tube and flows outward on the outside of the fibres, in a radial way. The pure water permeates the fibre membranes and flows axially along the inside of the fibres to a header at the end of the bundle.

RO processes require both pre and post treatment. The first ones are important because the membrane surfaces must remain clean: suspended solids must be removed and the water pre-treated so that salt precipitation or microbial growth does not occur on the membranes. Usually, the pretreatment consists of fine filtration and the addition of acid or other chemicals to inhibit precipitation and the growth of microorganisms. Post-treatment consists of stabilizing the water and preparing it for distribution and usually consists of the removing gases such as hydrogen sulphide and adjusting the pH.

3.2.3 Hybrid systems

Two or more desalination processes can be combined or coupled with a power plant in a hybrid configuration to produce water at low cost [70]. Where there is considerable fluctuation in water and power demands, it is very suitable to use hybrid desalination systems (co-generation power-MSF plant with SWRO plant). In general, MSF or MED is combined with VC and RO or nanofiltration (NF). Combination of processes and power production can more efficiently utilize fuel energy as well as the produced power. For utilization of idle power to produce water via RO or MVC, the additional produced water can be stored in aquifers and recovered when demand is higher, thereby increasing overall system efficiency and reducing cost [127].

3.2.4 Economical and Technical Comparison

When desalination started in the late 1950s, the cost was not as important since the main challenge was to produce fresh water from seawater for boilers and drinking purposes in ships. Later in the 1960s and early 1970s, desalination technologies (thermal processes) were widely available for commercial production but the cost was still too high. Membrane processes began to compete in the 1970s and started the trend toward costs reduction. As late as 1975, seawater desalination costs were quoted in planning documents as being about US\$ 2.10/m^3 (Southwest Florida Regional Planning Council, 1980). The expansion of the desalination market has attracted many organizations and companies to improve desalination technologies to reduce the costs. Remarkable decreases in desalination costs were continuously achieved in the last decades causing the water price to reach US\$ 0.50/m^3 [128] for large scale SWRO plants and for specific local conditions, and below US\$ 1.00/m^3 for MSF.

In addition, although desalination is expensive compared to conventional treatment of fresh water, the cost of desalination is decreasing, while the costs for developing new fresh water sources of potable supply are increasing or no longer possible. For example, prices of thermal processes are falling due to material improvements, process innovation, and increasing competition. Also, as technological developments cause a reduction in the cost of equipment, the overall relative plant costs are expected to decline. This trend has made desalination, once a costly alternative to the provision of potable water, a viable solution and economically competitive with other options for water supply. The main factors that contributed to cost reduction in all desalination processes are the significant improvements in the performance of these technologies during the last recent years. Increasing in plant capacity have also contributed to a reduction in unit water cost [129]. The magnitude of the respective costs due to improvements in the membranes and increases in plant capacity are difficult to measure since they have both taken place simultaneously. Plant capacities increased by a factor of 10 between 1995 and 2010.

The investment cost of different commercial desalination technologies differs widely between thermal-based and membrane-based technologies (see Figure 26). For a similar plant capacity, thermal processes require larger footprints and use more costly materials and equipment than the SWRO process. Similarly, thermal processes consume higher amounts of specific energy (electrical and thermal) than RO (only electric requirements) and more chemicals are needed to control scaling, corrosion and foam. However, on the other hand, thermal distillate is of higher quality than the RO product. Also, thermal processes function using nearly any quality (read salinity) of feed water without extensive pretreatment. Therefore, in this study a thermal desalination technology has been chosen for the following reasons:

1. This work has the aim to direct the steam (thermal energy) to another process to work in load following mode, therefore without wasting energy trough a conversion with a net efficiency of about 33%
2. The “not-requirement” of expensive pre-treatment, the ease of the process, and the application of a proven working technology can guarantee a strong reliability, that could represent a key issue in a coupling with a NPP
3. The high quality of the distilled water is what is needed to refill the fermentors of the algae cultivation process, in a possible scenario of coupling both the biorefinery and the desalination plant to the NPP
4. Modern thermal distillation plants, like plant with a MED-TVC layout, have already given the proof to be highly competitive with SWRO plants, especially if coupled with a power plant. Indeed costs between these two technologies are comparable [70].

The thermal process can be chosen between MED (or MED-TVC) and MSF. VC alone is not suitable for large industrial desalination. MED desalination processes are superior to MSF processes from a thermodynamic point of view because of lower total energy consumption (as explained in the next paragraphs). This can be shown comparing the GOR of the two kind of plants with the same heat transfer area and the same temperature difference between the heat source and the cooling water sink: the GOR of MED plants is much higher. Furthermore, recent MED plants work with TBT lower than MSF and this is a positive factor to limit scaling. Advantages of MED process, compared to MSF process, can be summarized as following:

1. Lower electric needs
2. Higher efficiency (GOR), therefore lower thermal needs required
3. Lower top brine temperature (TBT), thus less potential scaling problems
4. Pretreatment of sea water easier and less expensive because of lower TBT (for example cheaper and less harmful anti-scale additives)
5. Cost of material is lower due to low corrosion and less harmful additives

6. Lower cost per cubic meter of desalinated water

A more detailed explanation, with some examples, will be supplied in the next two paragraphs. In addition, MED process efficiency rises if coupled with a TVC process. TVC has always been provided in recently constructed MED plants, when motive steam is available at higher pressure than the one required in the first effect. For the reasons just discussed, it is possible to say that nowadays a MED process seems to be more attractive than MSF.

Therefore, in this work a MED desalination plant will be considered for the coupling with the nuclear plant. More exactly, a TVC-MED LT- HTFE (Horizontal Tube Film Evaporators) would be the best choice.

3.3 COSTS IN DESALINATION

In this section, firstly the main cost items for a desalination plant will be introduced, together with a comparison between values for the different technologies. Secondly, some examples for recent construction plants and the current general trend will be discussed. Finally data adopted here for the economic analysis will be supplied.

The capital cost includes all expenditures associated with the implementation of a given desalination project from the time of its conception, through design, permitting, financing, construction, commissioning and acceptance testing for normal operation [130]. Capital cost is often referred as CAPEX or Capital Expenditure [131]. Capital cost includes direct and indirect costs. Direct capital costs represent the installed process equipment, auxiliary equipment (MSF/ MED equipment is generally more costly than RO), and the associated piping and instrumentation, site civil works, intake (may include wells, open or sub-surface intakes) and brine discharge (may include outfall, injection wells, evaporation ponds) infrastructures, buildings, land, roads and laboratories. Construction costs are typically 50–85% of the total capital cost. Indirect capital costs represent interest during construction (overheads), working capital, freight and insurance, contingencies, import duties (in some cases waived), project management, and architectural and engineering (A&E) fees. These costs are usually calculated as a percentage of the direct capital costs with an average of 40% [132], 15–50% [130] or 30–45% [133], but are very project specific.

O&M costs are site-specific and consist of fixed costs and variable costs. Fixed costs include insurance and amortization (annual interest for direct and indirect

FIGURE 26. COMPARISON BETWEEN 4 MAIN TECHNOLOGIES TO PRODUCE DESALINATED WATER IN MATTER OF INVESTMENT COST AND ENERGY REQUIREMENTS [48].

Process	Thermal energy kWh/m^3	Electrical energy kWh/m^3	Total energy kWh/m^3	Investment cost \$/m^3/d	Total water cost US\$/m^3
MSF	7.5–12	2.5–4	10–16	1200–2500	(0.8–1.5)[a]
MED	4–7	1.5–2	5.5–9	900–2000	0.7–1.2
SWRO	–	(3–4)[b]	3–4	900–2500	0.5–1.2
BWRO	–	0.5–2.5	0.5–2.5	300–1200	0.2–0.4

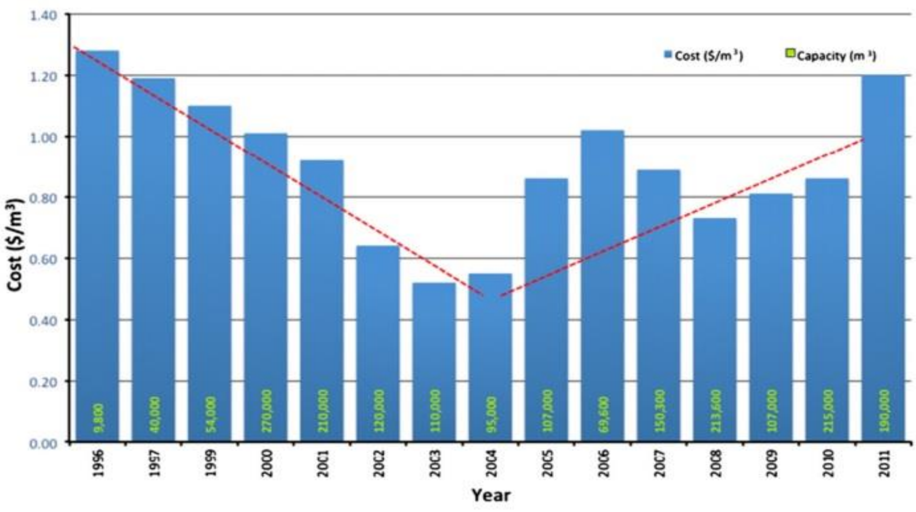
[a] Including subsidies (price of fuel).
[b] Including energy recovery system.

costs) costs. The primary variable operating costs (OPEX) include the cost of labour, energy, consumables (chemicals, membrane replacement, pump replacement), maintenance, and spare part costs, which are dependent on the relationship of facility location to manufacturing and distribution centres.

The total water cost (TWC) is the sum of capital cost and operating cost for the contract period. The cost is calculated by dividing the sum of the annualized capital costs and the annual O&M costs by the average annual potable water production volume. In general, TWC excludes the distribution costs. TWC and investment costs of the new Barcelona brackish water EDR plant, having a total capacity of 200,000 m/d^3 and feed salinity of 2 g/L are US$ 0.26/m^3 and US$ 79.56 million, respectively. TWC for large-scale thermal process facilities ranges from US$ 0.80 to 1.50, US$ 0.70 to 1.20, and US $0.60 to 1.00/ m^3 for MSF, MED and VC, respectively. Some recent TWCs of various desalination plants using different technology are summarized in Table 11. It shows costs with significant differences in total price. In some projects, water price is high due to specific conditions like the necessity to install complex pretreatment systems or because of severe environmental regulations that increase permitting and construction costs. Actually, in comparison to prices of Figure 26, the Table 11 shows higher costs for desalination technology. This could be explained looking at the Figure 27: it shows that the TWC decreased significantly from 1991 to 2003 mainly due to the technological developments described. But, when TWC is extended a little further into the future, the projected cost curve begins to turn upwards (dotted line).

Therefore, after extensive investigation of the TWC evolution over the last decades and for recent projects, the authors join the expectation of many desalination experts that the desalinated water cost will not reduce further at the same rate of decline for several reasons. The primary reasons are the instable prices of crude oil that affects energy costs, currency fluctuations, and increases in membrane prices. Membrane prices were kept stable in the last years by technology improvements and competition causing many companies to operate at low profit margins. Many membrane manufacturers predict that there is no way to avoid membrane price increases in the near future [126]. However both cost of crude oil and rising in price of membranes don't affect the MED-TVC plant in this study. Another important issue is instead the increased costs of shipping, raw

FIGURE 27. TREND IN DESALINATION COSTS IN THE LAST TWO DECADES.



materials (particularly specialty metals), equipment and chemical prices along with more restrictive environmental regulations

Furthermore, in some projects, intake and reject disposal systems costs are higher than the cost of the entire capital cost of a similar plant in another country where restrictions are not severe. Short supply of highly skilled man power for plant construction and O&M could also be a factor in cost increases [134]. Expected improvements in existing technologies, such as optimization of chemical dosing, post-treatment [135], and new cleaning methods (without shutting down the desalination unit) may not significantly reduce the cost of desalination in the future. Existing desalination technologies are mature and desalinated water cost reductions will be marginal due to further developments in technologies.

Making a first raw approximation, for a nuclear site composed by SMRs it is possible to consider a installed thermal power of about 4000 MWt. Considering a thermal power requirement of 50 kWh/m^3 (see par. 3.4), and having such an amount of power supply, the capacity of the MED-TVC plant could be about 800000 m^3/d. Hence, in such a project, there could be some important economies of scale, that could reduce costs. Then, Figure 26 and considerations about economies of scale for large capacity plants could suggest to use a cost of about 1000 \$/(m^3/d). But to follow the trend showed in Figure 27, a more precautionary cost for MED-TVC plant has been assumed in this study, also because the modularity of such huge plant could decrease the advantage of economies of scale itself. Indeed, after a continuous strong decreasing in prices until 2003, nowadays costs are predicted to slightly increase, due to fluctuations in market and increasing of prices for raw material and energy in general. Cost of energy don't affect directly our desalination facility, thanks to the coupling with the nuclear power plant. But it could affect costs indirectly, for example making more expensive other surrounding requirements like shipping, etc. Instead, in matter of O&M, the fact that energy has not to be purchased in a cogenerative layout, can strongly reduce the costs; for this reason, O&M used in this work is considerably lower than data offered in literature:

- Capital cost is considered to be 1300 \$/(m^3/d)
- O&M is assumed to be 0,10 \$/m^3 [125] [136]

TABLE 11. DATA IN MATTER OF COSTS FOR SOME RECENT DESALINATION PLANTS.

Site	Technology	Start-up	Capacity [m^3/d]	Capital Cost [M$]	Product [million m^3/y]	Capital Cost [$/($m^3/d$)]	Annual Capital Cost [M$]	TWC [$/$m^3$]	Annual Capital Cost [$/$m^3$]	O&M [$/$m^3$]
Fujairah 2	MED	2011	460000	616	151,1	1339	24,6		0,16	
Fujairah 2	SWRO	2011	136000	190	44,7	1397	7,6		0,17	
Adelaide	SWRO 2-pass	2010	273000	1790	89,7	6557	71,6		0,80	
Sydney	SWRO 2-pass	2010	250000	933	82,1	3732	37,3		0,45	
Hadera	SWRO	2010	347900	425	114,3	1222	17,0	0,63	0,15	0,48
Shuaiba	MSF	2010	880000	2400	289,1	2727	96,0	0,95	0,33	0,62
Barka 2	SWRO 2-pass	2009	123500	800	40,6	6478	32,0		0,79	
Marafiq	MED-TVC	2009	800000	3400	262,8	4250	136,0	0,83	0,52	0,31
Skikda	SWRO	2008	100000	110	32,9	1100	4,4	0,73	0,13	0,60
Ras Laffan B	MSF	2008	272500	900	89,5	3303	36,0	0,8	0,40	0,40
Oxnard	BWRO	2008	28400	25	9,3	880	1,0	0,31	0,11	0,20
Alicante 2	SWRO	2008	65000	89	21,4	1369	3,6		0,17	
Hamma	SWRO	2008	200000	250	65,7	1250	10,0	0,82	0,15	0,67
El Paso	BWRO	2007	55670	87	18,3	1563	3,5	0,41	0,19	0,22
Perth	SWRO 2-pass	2007	143700	347	47,2	2415	13,9	1,2	0,29	0,91
Palmachim	SWRO	2007	110000	110	36,1	1000	4,4	0,78	0,12	0,66
Rabigh	MED-TVC	2005	25000		8,2			1,15		
[125][13]	MED-TVC		220000	220	72,3	1000	8,8	1,98/0,2	0,12	1,86/0,07
This Study	**MED-TVC**					**1300**				**0,10**

In the table the kind of technology and the stat-up date are indicated. Table has been elaborated by data of Table 4 in [70]. To calculate the annual productivity of water a availability of 90% has been assumed. The annual productivity permitted to calculate the capital cost referred to cubic metres of water produced (columns 10 and 11). To calculate data of column 10, also a life of 25 years has been presumed for the plants. Then, to calculate O&M, a simple subtraction of annual capital cost from the TWC values has been done.

[13] For O&M column there are two values. The first one consider also the purchase of energy (both natural gas and electricity). The second one refers to a case of coupling with a NPP. Therefore O&M, and consequently TWC, become lower.

3.4 ENERGY CONSUMPTIONS IN DESALINATION

The minimum energy consumption required for separating a saline solution into pure water and concentrated brine under ideal conditions is dependent only on the salt content of the saline solution, regardless of the technology and configuration of the desalination system in question. In other words, all desalination systems, which may be based on different technologies and may have different configurations, share a common minimum energy requirement for driving the separation process, regardless of the system. In practice, however, the energy requirements in all desalination processes are considerably higher than those computed for the reversible ideal separation. This is because a certain process irreversibility occurs due to friction losses, non-equilibrium and other thermal losses, including boiling point elevation, flow resistance through membranes and pump efficiencies. Hence, the deviation of the actual energy required in any given desalination system depends on the system's design and engineering characteristics and its principle of operation in the quantity and type of losses encountered during separation. The actual energy consumption is four times higher than the minimum energy required to produce fresh water from seawater.

The typical electric energy consumption for a SWRO plant is reported to be 3–4 kWh/ m^3 excluding water distribution systems. In MED the specific electric power consumption is below 2 kWh/m^3 of distillate, which is significantly lower than MSF which is typically 4 kWh/ m^3 [70]. The major advantage of the MED process, however, is the ability to produce significantly a higher performance ratio (GOR). If the corrosion and scaling potentials are reduced through some means such as the application of the NF membrane softening process, then these plants can be operated at higher temperatures with a corresponding increase in efficiency and significant decrease in matter of costs [137]. The described improvements helped in reducing the unit water cost produced by MSF to less than US$ 1.00/m^3 for large-scale plants. The investment cost of large MSF plants is about US$ 1500/(m^3/d), based on the average value of recently contracted plants installed in the Middle East.

However, MED is similar to MSF in that it requires two types of energy, namely, low temperature heat and electricity. The low-temperature heat is the main portion of the total energy input to the system regardless of whether it is supplied by the extracted steam from a power plant, waste heat recovery boiler, or fuel-fired boiler. Electricity instead is mainly required to drive the system's pumps. Concerning the thermal needs of a MED plant, they strongly depend on the GOR of the system. Moreover, it makes a significant difference whether it is coupled or not with a TVC module. For MED without TVC thermal needs range from 60 to 80 kWh/m^3 [138] [136] [139]. Whereas the combining with a TVC, with a typical GOR value of 12, can lower these needs to 40 kWh/m^3 [140]. In this work it was chosen to couple the MED plant with a TVC. Hence, the average energy needs estimated are the following:

- Thermal energy needs: 50 kWht/m^3 (even if TVC-MED plants with lower thermal energy needs than 50 kWht/m^3 do exist);
- Electric energy needs: 2 kWhe/m^3 (almost every MED plant features an electric consumption of 1,5-2 kWhe/m^3, according to Figure 26);

where the cubic meter refers to the fresh produced water.

CHAPTER 4.

TECHNICAL CONSIDERATIONS: POWER REQUIRED, SIZING AND LAYOUTS OF THE PLANTS

In the fourth chapter, first technical results of this dissertation are discussed. In particular the calculations in matter of power needed by the biorefinery are inserted. Then even a comparison between scenarios' results is introduced (always for the biorefinery). This comparison allows to choose the best scenario to investigate in the further economic analysis. Once the technology is chosen, even a discussion regarding the sizing of the plants is inserted in the following of the chapter. Instead, at the end of the chapter, other technical considerations and specific comments to finalize the layout of the plant are inserted. In particular, all the power requirements and technical values to insert in the economic balance sheets are listed.

4.1 BIOREFINERY

4.1.1 Model and Scenarios' results

As explained in paragraph 2.3, the biorefinery is subdivided by 5 main phases: cultivation, harvesting & dewatering, oil extraction, biodiesel production and finally bioethanol production. The numbers, yields, power requirements per unit of algae produced or per biofuel obtained, efficiencies, etc. that characterize the entire production chain are already discussed in chapter 2. Here they are combined in a work-sheet that represents the model of the biorefinery. By inserting the inputs parameters, it restores the yields of biofuels and the power required to produce them.

In particular the input parameters are different for each scenario evaluated, regarding the variability of values in matter of the cultivation phase. On the contrary all the subsequent phases are considered the same for each scenario.

Firstly, let's introduce the inputs parameters of 5 scenarios, that are listed in Table 12. The size of cultivation's plants is preliminary chosen only with the aim to make a comparison between scenarios. As said many times within the thesis, nowadays there is a lack of available data for plants of a commercial-scale, because this approach to produce biofuels is new and therefore still under study and in a development stage. Regarding open pond systems, broadly in literature studies are done on a size in the range of 100-400 ha [26], [110]. Even data reported concerning pilot scale facilities match this range. In this context, the project "Green Crude Farm" of Sapphire Energy (headquarter in San Diego, California, USA) is a perfect example: currently its pilot scale is a 40 ha facility, with the goal to expand it to 120 ha in the next few years, producing about 4 million litres per year (MLPY) of biofuels. As technology is proven and economies of scale are achieved the project considers to start the construction of the first commercial biorefinery in 2015, and by 2018, Sapphire aims to produce about 200 MLPY [10], therefore expanding the ponds to thousands of hectares.

Hence to make a preliminary analysis we set the size of the open ponds biorefinery to 400 ha. Setting the size of the fermenter biorefinery to 400 ha is not reasonable, because it should be composed by 250000 units, producing about 24 billion litres per year (indeed a very high yield of biofuels per land occupied is expected in case of fermenters). To have a fair reasonable contrast between open ponds scenarios and fermenter scenario, we chose as a parameter the biomass grown per day (metric tonnes per day). Since this data is variable and it depends not only on the size of the plant, but even by the yield hypothesized in the scenario, we haven't a punctual precise value, but a range of values: for a 400 ha facility the yield of algae harvested per day was in the region of 62-99 ton/day (data calculated with the model). Typically the size of fermenters is measured by volume. Therefore to obtain a yield of biomass within this range we set the size of fermenter biorefinery to 16 million litres (composed by 160 units, occupying a total land of 0,3 ha). This size permits to have 94 tonnes of algae harvested per day, as shown in Table 15.

The model starts with the cultivation phase. This receives as input the yield of algae grown, assumed in a particular scenario. For example in case of scenario 1, it is 24,75 g per m^2 per day. In case of fermenter we assume that the harvesting and dewatering phase is done "in continuous". Actually, considered data report a cultivation time of about 167 hours (7 days).

TABLE 12. INPUTS PARAMETERS FOR 5 SCENARIOS EVALUATED.

ITEM		SCENARIO 1. BASE CASE	SCENARIO 2. FERMENTER	SCENARIO 3. UNFAVORABLE CLIMATE	SCENARIO 4. LIPID RICH ALGAL STRAIN	SCENARIO 5. LOW N
Cultivation type		Open pond raceway	Fermenter	Open pond raceway	Open pond raceway	Open pond raceway
Algal strain		*Chlorella Vulgaris*	*Chlorella Protothecoides*	*Chlorella Vulgaris*	*Botrycoccus Braunii*	*Chlorella Vulgaris*
Dimension of a single pond (Length x Width x Depth) [m x m x m]		100 x 10 x 0,3	-	100 x 10 x 0,3	100 x 10 x 0,3	100 x 10 x 0,3
Dimension of a single fermenter (High x Diameter) [m x m]		-	10,5 x 3,5	-	-	-
High-to-diameter ratio		-	3	-	-	-
Single Unit Area [m^2]		1000	9,6	1000	1000	1000
Single Unit Area Occupied [m^2]		1000	17,3	1000	1000	1000
Single Unit Volume [m^3]		300	100	300	300	300
Capacity utilized [%]		100%	80%	100%	100%	100%
Number of Units		4000	160	4000	4000	4000
Total Land occupied [ha]		400	0,3	400	400	400
Total Volume [million of L]		1200	16	1200	1200	1200
Climate Condition		Favourable	Not important	Unfavourable	Favourable	Favourable
Average Yearly Temperature [°C]		13	-	7	13	13
Average Yearly Solar Irradiance [kWh/m^2d]		3,65	-	2,8	3,65	3,65
Optimal temperature [°C]		26	28	26	22	26
Final Cell Concentration [g/L]		0,33	51,2	0,21	0,33	0,26
Time needed for the growth		Continuous	167 hours	Continuous	Continuous	Continuous
Volume Harvested per day [%]		25%	-	25%	25%	25%
Volume Harvested per day [m^3/d]		300000	-	300000	300000	300000
Yield [g/m^2d]		24,75	-	15,47	24,75	19,25
Composition of alga	Protein	29,00%	10,28%	29%	22%	7%
	Lipid	20,00%	55,20%	20%	60%	40%
	Carbohydrate	50%	15,43%	50%	14%	55%
	(of which glucose)	90,4%	90,4%	90,4%	90,4%	90,4%

A fermentation process has to be constantly monitored, sometimes it could require more hours, other times less. In fermenters it is possible to have a batch system or a continuous system, that is a fermenter with a hole in the bottom for a continuous discharge of the medium which is contained inside. Because of the large multitude of fermenters involved in the system, we can hypothesize that even if we have a batch system, it is possible to organize the growth shifting the starting of the operations of fermenters during a week. Therefore in this work we assume that constantly, every day there is the same quantity of algae harvested, assuming that cultivation was in continuous. Knowing the tons of biomass grown and cubic meters of water involved, we can calculate the power requirements for the cultivation phase: for the 400ha raceway scenario it is about 4,6 MW, while for 16 million litres fermenters it is 16 MW. They represent almost 90% of whole electricity required in open pond case, whereas almost the 98% in case of scenario 2.

After, the biomass is harvested and starts the dewatering process. In case of an open pond system, it begins with a flocculation. Power required for flocculation is already included in water pumping of the previous phase. The biomass is taken out from flocculation pool still containing a liquid percentage of 98%. Because of the microalgae harvested from fermenters have a higher "dry-weight" content (final concentration level of about 50 g/L, that corresponds to 5% w/w[14]), the flocculation step is not needed in this scenario. From here, both scenarios follow exactly the same route.

According to literature, we consider that for every dewatering step, a loss of biomass occurs. This is the meaning of efficiency for the dewatering processes. Actually the literature doesn't consider a loss in the thermal drying step (efficiency 100%), but conservatively we consider some eventual loss and an efficiency of 95%. The overall dewatering phase has a not negligible efficiency of 70%. The remaining 30% of grown algae doesn't enter the oil extraction process because it is physically lost during the dewatering steps or because the biomass putrefies and it is no longer available. Therefore, considering scenario 1, of 99 tons grown every day, only 69,3 tons go through the following process of the chain. The power requirements are also important: only 0,2 MWe are required, but 6 MWt are needed, representing about the 70% of the whole thermal energy required in the entire production cycle.

Also the oil extraction phase is very delicate and complicated. The solvent extraction process is a well-established technique. Despite this, the achievable efficiency using algae is still not clear. Literature usually suggests pretreatment process to increase the amount of lipids that can be extracted: from about 70% [89] without any pretreatments this figure can be raised to approximately 95% [33], [141] with a previous pretreatment (i.e. grinding or using ultrasounds). We assume that 92,5% of lipids could be extracted by hexane extraction, forerun by a grinding step. Therefore, for scenario 1 (20% lipids content) 69,3 tons of algae enter the extraction phase: 12,8 tons is extracted in oil to send to the biodiesel branch, while the rest (56,5 tons, called alga cake) is sent to the bioethanol branch. Energy requirements are very low for the extraction phase and, broadly, even for the successive production phases. This is explained because data refers to the power required per tons of biofuel produced. A 400 ha algae facility permits to obtain at the end only about 10 MLPY of biofuels. Instead, for example in the USA conventional (first generation) biorefineries usually reach a capacity of 150-200 MLPY. Therefore we can immediately understand that a 400 ha algae facility

[14] Assuming the density of algae equal to the density of water.

is small in comparison with a common biorefinery and the energy necessities are consequently low. This aspect will be further discussed in section 4.1.2. Therefore, although we would choose a more sophisticated pretreatment process for this phase (e.g. ultrasounds), they are even more demanding from the energetic point of view, we expect that the whole electric requirement wouldn't change considerably. This consideration allows to adopt an optimistic efficiency of 92,5%.

Now the description is split between the two branches. Firstly biodiesel is analysed.

About 12,8 tons of oil enter the biodiesel conversion process. As described in Appendix A. the oil starts with a refining step, during which 4% is lost. Then, 12,3 tons of degummed refined oil process with the transesterification reactions to be converted in biodiesel. We assume that 99,4% of oil is transformed into biodiesel, whereas 9% into glycerol. The sum is greater than 100%, because the chemical reaction entails a significant amount of methanol to take place (usually the quantity of methanol is 10% in weight of the quantity of oil). Even in this phase power requirements are low: only 42 kWe and 0,3 MWt. The yield of biodiesel obtained is 12,2 tons per day, with a co-production of 1,1 tons per day of glycerol (80% pure grade). Considering a plant availability of 90%, it corresponds to 4,54 MLPY of biodiesel.

Similarly we can calculate the yield in ethanol. The alga cake produced as a result of the oil extraction phase is composed mainly by carbohydrates, proteins and a slight percentage of not-extracted lipids. To calculate the yield of ethanol we need to know the following data (composition is referred to scenario 1, Table 12):

- Alga cake 56,5 tons
- Lipid content 20%
- Carbohydrate content 50% (of which 90,4% is glucose, mainly starch and cellulose)
- Oil extraction efficiency 92,5%
- Ethanol yield in fermentation reaction 50% (the other 50% is composed by CO_2)

Therefore, remembering that only the amount of glucose is fermentable, the yield is calculated as follow

$$\begin{aligned} Yield_{ethanol} &= 56{,}5\ tons \cdot \left(\frac{1}{1 - 20\% \cdot 92{,}5\%}\right) \cdot 50\% \cdot 90{,}4\% \cdot 50\% \\ &= 56{,}5\ tons \cdot 27{,}7\% = 15{,}7\ tons \end{aligned}$$

Assuming an availability of 90%, the biorefinery would produce 6,52 MLPY of ethanol. The corresponding co-products are the DDGS (Dried Distillers Grains with Soluble), sealable like animal feeds or generally a product with a high protein content or simply a burnable biomass. Yield of DDGS is calculated taking the amount of alga cake and subtracting the products of the fermentation reaction: ethanol (15,7 tons) and CO_2 (15,7 tons). Energy needed is 1,6 thermal MW and 229 electric kW.

Therefore in scenario 1, to produce 11 MLPY of biofuels we calculate a power needed of 5,22 MWe and 8,76 MWt.

Similarly it is possible to calculate yields and power requirements for all the other scenarios, changing inputs parameter. The results are displayed in Table 14 and in Table 15.

TABLE 13. THE MODEL FOR THE BIOREFINERY. DATA REPORTED REGARD THE FIRST SCENARIO, IN CASE OF A RACEWAY OF A 400 HA AREA.

PRODUCTION PHASE	PROCESS	RESULT	EFFICIENCY	WATER INLET	WATER OUTLET	WATER REMOVED	QUANTITY	LITERATURE DATA	THERMAL POWER	ELECTRIC POWER
Cultivation	Puddle wheel mixing + CO_2 circulation	-	-	Pond volume 1200000 m^3			99 ton/d	3,72 W/m^3	-	4,464 MW
	Pumping water	-	-	-	-	-	99 ton/d	71,2 kW/ton	-	0,206 MW
	SUBTOTAL	-	-	-	-	-	*99 ton/d*	-	-	*4,67 MW*
Harvesting & Dewatering	Flocculation	From 0,033% to 2%	91%	299901 m^3/d	4414 m^3/d	295487 m^3/d	90,09 ton/d	See text	-	0
	Centrifugation	From 2% to 12%	90%	4414 m^3/d	594 m^3/d	3820 m^3/d	81,08 ton/d	1 kWh/m^3	-	0,184 MW
	Filtration	From 12% to 27%	90%	594 m^3/d	197,3 m^3/d	397 m^3/d	72,97 ton/d	0,88 kWh/m^3	-	0,021 MW
	Thermal drying	From 27% to 90%	95%	197,3 m^3/d	7,7 m^3/d	189,6 m^3/d	69,32 ton/d	628 kW/m^3	6,009 MW[15]	-
	SUBTOTAL	-	*70%*	-	-	-	*69,32 ton/d*	-	*6,009 MW*	*0,206 MW*
Oil Extraction	Grinding	-	-	-	-	-	-	4,14 kWh/ton E.E.	-	0,012 MW
	Oil extraction	-	-	-	-	-	-	3,6 kWh/ton E.E.	-	0,01 MW
	Meal processing	-	-	-	-	-	-	14,56 kWh/ton E.E. 109 kWh/ton t.	0,315 MW	0,042 MW
	Solvent recovery	-	-	-	-	-	-	0,52 kWh/ton E.E. 122 kWh/ton t.	0,352 MW	0,002 MW
	Oil recovery	-	-	-	-	-	-	0,38 kWh/ton E.E. 24,2 kWh/ton t.	0,07 MW	0,001 MW
	Oil degumming	-	-	-	-	-	-	1,69 kWh/ton E.E. 19,02 kWh/ton t.	0,055MW	0,005 MW
	Waste treatment	-	-	-	-	-	-	0,57 kWh/ton E.E. 10 kWh/ton t.	0,029 MW	0,002 MW
	SUBTOTAL	-	*92,5%*	*Crude degummed oil: 12,83 ton/d. Alga cake: 56,5 ton/d*				-	*0,821 MW*	*0,074 MW*
Biodiesel Production	Oil refining	-	96%	-	-	-	-	100 MJ/ton E.E. 600MJ/ton t.	0,085 MW	0,014 MW
	Two-steps transesterification	-	99%	-	-	-	-	200 MJ/ton E.E. 1600 MJ/ton t.	0,227 MW	0,028 MW
	Biodiesel purification	-	99,4%	-	-	-	-			
	Glycerine purification	-	9%	-	-	-	-			
	Methanol recovery	-	-	-	-	-	-			
	SUBTOTAL	-	-	*Biodiesel: 12,24 ton/d. Glycerol: 1,11 ton/d*				-	*0,311 MW*	*0,042 MW*
Bioethanol Production	*SUBTOTAL*[16]	-	*27,7%*	*Bioethanol: 15,67 ton/d. DDGS: 25,17 ton/d*				*1 MJ/L E.E.* *7,05 MJ/L t.*	*1,62 MW*	*0,23 MW*
Total Power Requirements				**Biodiesel: 12,24 ton/d. Ethanol: 15,67 ton/d**					**8,76 MW**	**5,22 MW**

[15] In the thermal power calculated we add also the power necessary to bring the algal slurry to the temperature needed for the evaporation of water.
[16] The production of ethanol is primarily divided in liquefaction, sterilization, saccharification, fermentation, distillation, dehydration, centrifuging, evaporation and drying. Because of respective used data are for the whole phase, they are not split in this table.

TABLE 14. RESULTS OF 5 SCENARIOS STUDIED FOR THE BIOREFINERY, REGARDING THE POWER REQUIREMENTS. SIZE OF OPEN PONDS: 400 HA. VOLUME OF FERMENTERS 16 MILLION OF LITRES.

PRODUCTION PHASE	PRODUCTION PROCESS	SCENARIO 1		SCENARIO 2		SCENARIO 3		SCENARIO 4		SCENARIO 5	
		EL. POWER [kW]	TH. POWER [kW]	EL. POWER [kW]	TH. POWER [kW]	EL. POWER [kW]	TH. POWER [kW]	EL. POWER [kW]	TH. POWER [kW]	EL. POWER [kW]	TH. POWER [kW]
CULTIVATION	puddle wheel mixing + CO_2 circulation	4464,00	-	-	-	4464,00	-	4464,00	-	4464,00	-
	pumping water	205,66	-	-	-	128,55	-	205,66	-	159,96	-
	Subtotal	*4,67 MW*	-	*16,00 MW*	-	*4,59 MW*	-	*4,67 MW*	-	*4,62 MW*	-
HARVESTING & DEWATERING	flocculation	0,00	-	-	-	0,00	-	0,00	-	0,00	-
	disk stack centrifuge	183,93	-	72,72	-	114,97	-	183,93	-	143,06	-
	chamber filter press	21,80	-	22,79	-	13,63	-	21,80	-	16,96	-
	thermal drying	-	6008,60	-	6281,62	-	3755,68	-	6008,60		4673,36
	Subtotal	*0,21 MW*	*6,01 MW*	*0,10 MW*	*6,28 MW*	*0,13 MW*	*3,76 MW*	*0,21 MW*	*6,01 MW*	*0,16 MW*	*4,67 MW*
OIL EXTRACTION	grinding	11,96	-	12,50	-	7,47	-	11,96	-	9,30	-
	oil extraction	10,40	-	10,87	-	6,50	-	10,40	-	8,09	-
	meal processing	42,06	314,85	43,97	329,15	26,29	196,80	42,06	314,85	32,71	244,88
	solvent recovery	1,50	352,40	1,57	368,41	0,94	220,27	1,50	352,40	1,17	274,09
	oil recovery	1,10	69,90	1,15	73,08	0,69	43,69	1,10	69,90	0,85	54,37
	oil degumming	4,88	54,94	5,10	57,44	3,05	34,34	4,88	54,94	3,80	42,73
	waste treatment	1,65	28,89	1,72	30,20	1,03	18,05	1,65	28,89	1,28	22,47
	Subtotal	*0,07 MW*	*0,82 MW*	*0,08 MW*	*0,86 MW*	*0,05 MW*	*0,51 MW*	*0,07 MW*	*0,82 MW*	*0,06 MW*	*0,64 MW*
BIODIESEL PRODUCTION	oil refining	14,16	84,99	40,87	245,22	8,85	53,12	42,49	254,96	22,03	132,20
	transesterification, and purification and recovery of products	28,33	226,63	81,74	653,93	17,71	141,66	84,99	679,90	44,07	352,54
	Subtotal	*0,04 MW*	*0,31 MW*	*0,12 MW*	*0,90 MW*	*0,03 MW*	*0,19 MW*	*0,13 MW*	*0,93 MW*	*0,07 MW*	*0,48 MW*
BIOETHANOL PRODUCTION	*Subtotal*	*0,23 MW*	*1,62 MW*	*0,08 MW*	*0,54 MW*	*0,14 MW*	*1,01 MW*	*0,06 MW*	*0,45 MW*	*0,20 MW*	*1,39 MW*
	TOTAL [MW]	**5,22**	**8,76**	**16,37**	**8,58**	**4,94**	**5,48**	**5,14**	**8,22**	**5,10**	**7,18**

TABLE 15. COMPARISON BETWEEN SCENARIOS.

ITEM	SCENARIO 1	SCENARIO 2	SCENARIO 3	SCENARIO 4	SCENARIO 5
Area [ha]	400	0,3	400	400	400
Volume [ML]	1200	16	1200	1200	1200
Biomass harvested [ton/d]	99	94	62	99	77
Electric power [MWe]	5,22	16,37	4,94	5,14	5,10
Thermal power [MWt]	8,76	8,58	5,48	8,22	7,18
Total power [MWt]	24	57	20	24	22
Biodiesel [MLPY]	4,54	13,11	2,84	13,63	7,07
Ethanol [MLYP]	6,52	2,19	4,08	1,83	5,58
Total biofuels [MLPY]	11,07	15,3	6,92	15,45	12,65
Specific power [MWt/LY]	2,2	3,76	2,92	1,52	1,77
Specific land requirements [ha/LY]	36,15	0,02	57,83	25,88	31,63
Productivity of biofuel per alga harvested [L/kg]	0,340	0,495	0,340	0,475	0,500
Hypothetical revenues[17] [M$]	10,5	18,6	6,5	19	13,1
Revenues/power required [$/W]	0,43	0,32	0,32	0,81	0,58

[17] Assuming a biodiesel price of 1,3 US$/L and an ethanol price of 0,7 US$/L.

FIGURE 28. SPECIFIC POWER REQUIRED FOR EVERY "LITRE PER YEAR" INSTALLED. LITRE PER YEAR IS THE MEASURE OF THE CAPACITY OF THE BIOREFINERY.

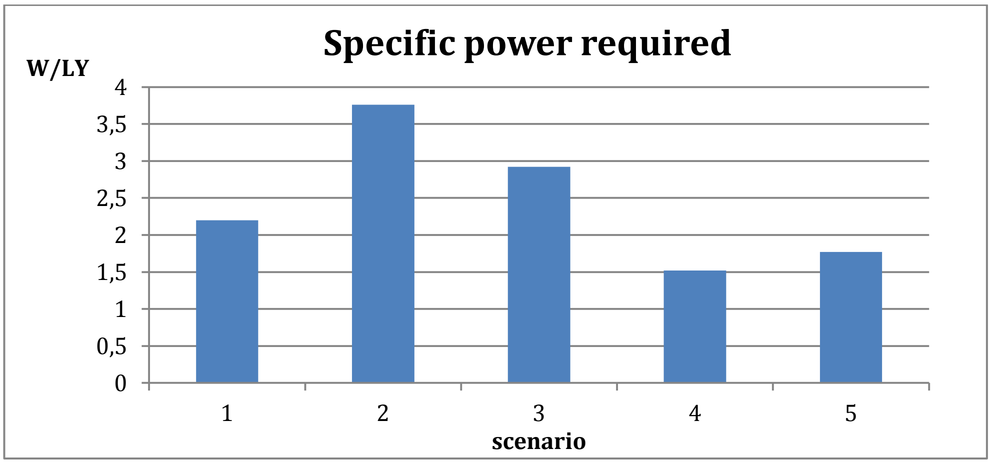


FIGURE 29. SPECIFIC LAND REQUIRED RELATED TO THE SIZE OF THE BIOREFINERY.

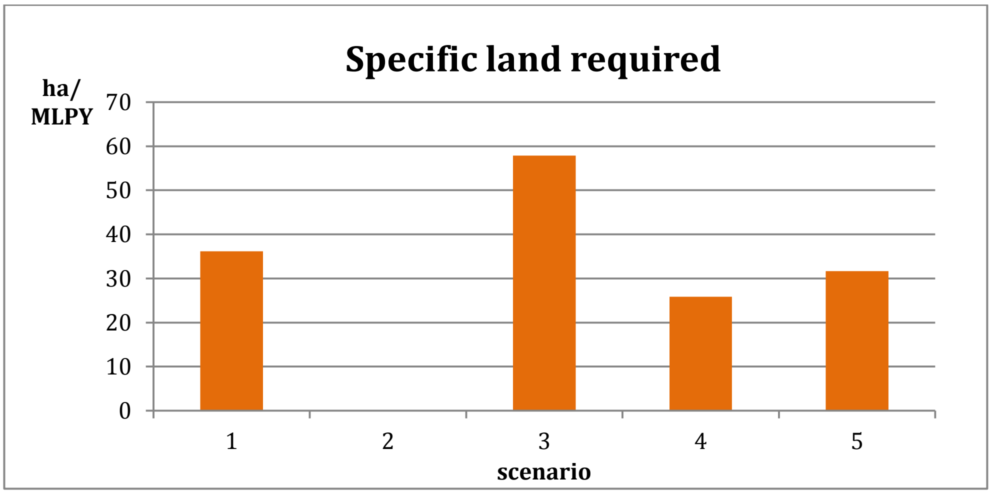
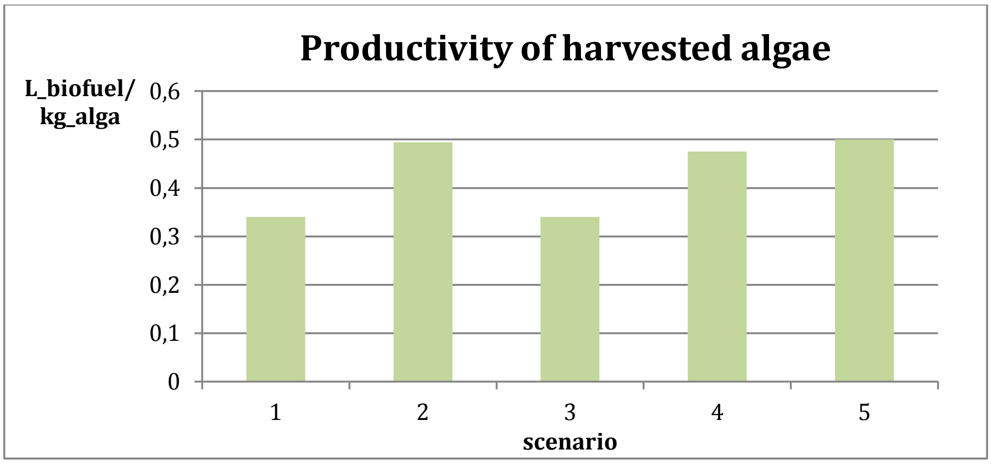


FIGURE 30. LITRES OBTAINED PER EVERY KILO OF ALGA HARVESTED.. IT IS A MEASURE OF THE QUALITY OF THE ALGA CULTIVATED.



4.1.2 Discussion and Comments

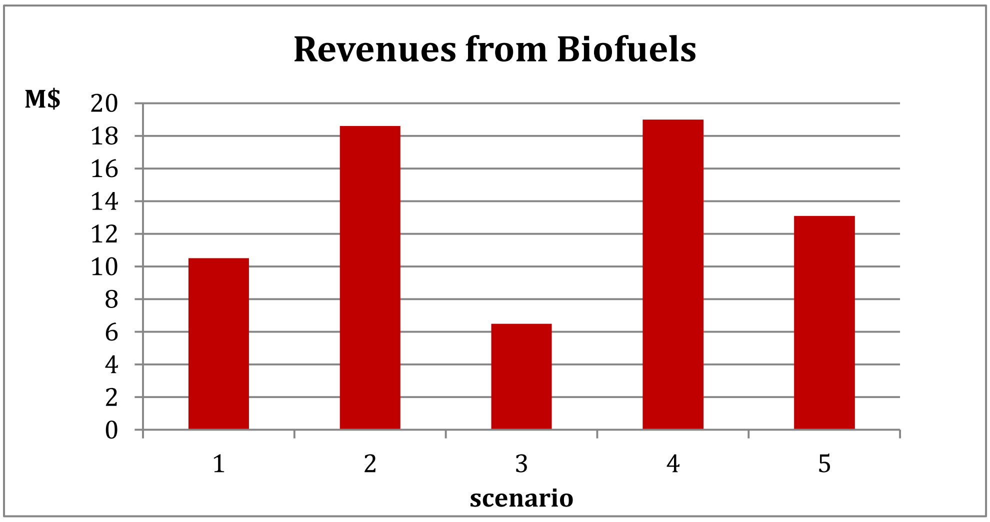
This section will start with a discussion regarding Figure 28, Figure 29 and Figure 30. The productivity of harvested algae is a direct measure of the quality of the alga in the scenario. Indeed the specifiic productivity (litres of biofuels per kilogram of alga) depends only by the algae's composition. The lower the protein content in the alga, the higher the whole productivity. Therefore the more interesting scenarios from this point of view are numbers 2, 4 and 5. Moreover, observing only open pond scenarios, the numbers 4 and 5 offer a higher performance even in the case of the specific power required and the specific land required. Indeed, having the same production chain, the fact to have a higher productivity enables the process to have even a lower specific demand of energy and land, as well. In addition scenario 4 seems to be slightly better than scenario 5, because the general reduced biomass yield in the latter case invalidates the fact to have a higher quality.

Finally, making even an economic consideration, scenario 4 produces mainly biodiesel, whereas in number 5 this distiction is not so evident. Considering that biodiesel has a higher price than ethanol (in the United States a fair, average price for biodiesel is 1,3 $/L whereas for ethanol it is 0,7 $/L) [142], scenarios characterized by a higher produciton of biodiesel, could enjoy higher revenues. This is underlined in Figure 31.

Therefore scenario 4 is the better case in a layout characterized by the cultivation in open ponds. Now it is interesting to have a contrast between scenario 2 and scenario 4. Productivity and revenues are very similar. The main differences are in matter of power requirements and land occupied: scenario 4 requires 59,4% less of the specific power required by scenario 2. On the other hand, scenario 4 needs 1450% more land than scenario 2. From these points of view, data are comparable with difficulty. Therefore a different kind of reflection is suggested.

From the energetic point of view, perhaps scenario 2 is not the best. Looking at the last line of Table 15 can help in the valuation. That line is obtained dividing the revenues [M$] per the power required in the scenario [MWt]. Indeed, we want to produce biofuels, subtracting a part of the thermal power produced by the nuclear

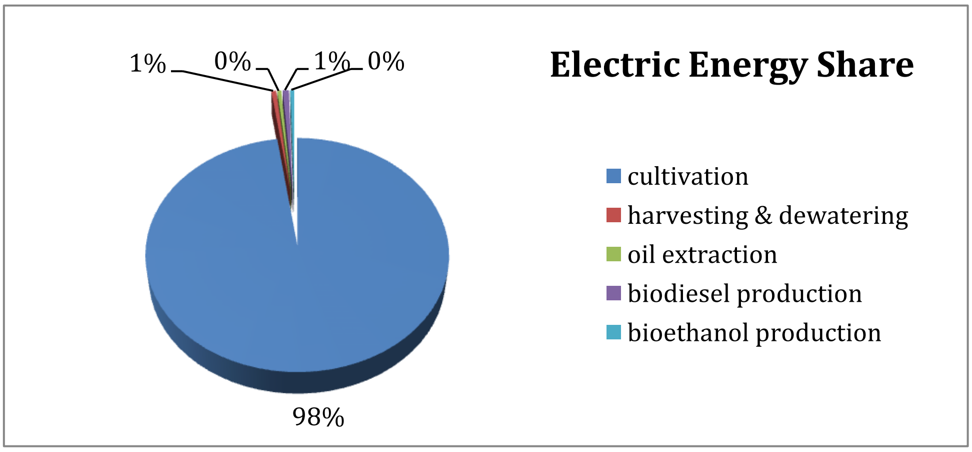
FIGURE 31. POSSIBLE REVENUES FROM THE SALE OF ONLY BIOFUELS IN EACH CASE. SCENARIOS HAVING A HIGHER PRODUCTION OF BIODIESEL COULD REACH A HIGHER PROFITABILITY.



power plant. Reducing thermal power to the turbine means producing less electrical energy, and therefore having less revenues from the sale of electricty. The last line tell how many dollars could be gained, selling biofuels, for every thermal MW subtracted to the turbine. Hence, once again, scenario 4 and 5 seems to be the most profitable. However it is important to introduce a commercial and practical consideration, observing Figure 29: in scenario 4, we need 26 ha for every million litres installed (capacity of biorefinery). That corrisponds to a power requirment of 1,52 MWt. Consequently to have a reasonable power requirement to couple with a nuclear power plant, in order to work in load following mode, the land occupied by the raceway ponds became immeasurable. For example, even for a small nuclear site (say 1000 MWt), it would be needed a thermal requirement biorefinery-side of about 500 MWt, that corrisponds to 8550 ha (and consequently a productivity of 329 MLPY of biofuels). Therefore, even if scenarios 4 and 5 seem to be characterized by higher performance coefficients, they are not usable for our purpose.

In this study, we are not considering further open ponds scenarios, but we focus only on scenario number 2, i.e. fermenter case. It is important to remember that, from the environmental point of view, this represents a lost in matter of sustainability. Actually, cultivating algae in fermenters, the biomass can't absorb the CO_2. In addition, a small amount of CO_2 is in contrast produced during the fermentation of the bioethanl production phase. However, before proceeding with other comments, it is important to clarify even one other aspect. The reader could think to consider the option to buy the biomass from an algae producer. With this possiblity, the issue of land requirement would not be a problem any longer. An investor in SMRs that would couple a biorefinery could simply focus on the capacity of biorafinery in term of biofuels produced, and not in term of hectares. But nowadays cultivating algae is very expensive, because, as said, the high energy requirements needed for the cultivation phase (see Figure 32). Therefore is not economical profitable buying algae to transform in biofuels, because their price is very high. This is confirmed by the fact that companies are investing in this technology providing both the cultivation and the conversion in biofuels phases. Moreover this is even the reason that makes the algae of Figure 77 so expensive, mostly in comparison with conventional biomasses. Then currently buying algae from the market is too expensive. So from the economic point of view it seems mandatory to introduce even a cultivation phase and consequently the land requirements became an important restriction for the purpose of this thesis.

FIGURE 32. ELECTRIC ENERGY SHARE BETWEEN PRODUCTION PHASES, FOR A BIOREFINERY IN CASE OF FERMENTER CULTIVATION SYSTEM.



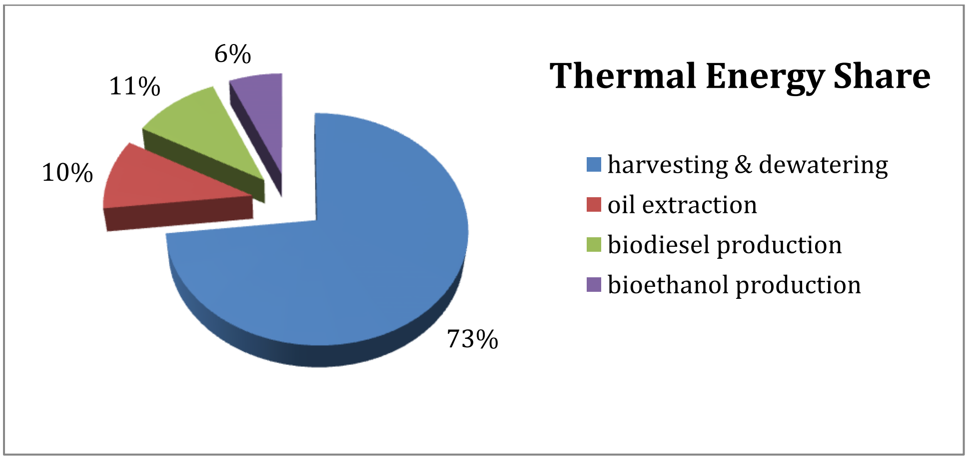


FIGURE 33. THERMAL ENERGY SHARE BETWEEN PRODUCTION PHASES, FOR A BIOREFINERY IN CASE OF FERMENTER CULTIVATION SYSTEM.

In the following we concentrate the study on a biorefinery in which algae grows heterotrophically inside stirred tank fermenters (for more details see section 1.1A and paragraph 2.2).

Looking at the relative column in Table 14, it is possible to represent in a graph the sharing of energy required in the entire cycle between all the process involved, Figure 32 and Figure 33. In particular it is possible to see how the largest part of energy is used to grow the algae (electric energy) and to dry them (thermal energy).

Because of the period that algae takes to grow in fermenters (7 days), the energy in the cultivation phase has to be supplied continuously It is not possible to switch off the fermenters, otherwise the biomass dies, since fermentation process is very delicate from chemical and biological point of view: the tanks must be always monitored and internal stirred by impellers.

When algae are grown they are harvested. They are initially inside a very liquid medium. (5% w/w solid content). Therefore the logistic to store such amount of algal slurry is very delicate, it would require additional pipes, pumps and vessels for the storage and, mostly, it risks to comprise the organic cells of microalgae. Indeed, if algae are not treated within few hours they risk to putrefy, and their worthwhile components could be no longer available. Therefore even the dewatering phase has to be done in continuous. This means that less than 30% of thermal energy would be available to work in load following mode.

Consequently we have to assert that to work in load following mode with a algal biorefinery seems not possible. This is because the largest part of energy has to be supply to the auxiliary not-electric plant continuously during the day. Even switching on the subsequent phases (oil extraction, biodiesel and bioethanol production) would not be sufficient to guarantee the requirement of 50%, imposed by [7] in matter of load following.

However, the algal biorefinery remains an interesting application to couple to the NPP in a static cogenerative configuration. For the biorefinery scenario we explore the economic suitability with the conventional Discounted Cash Flow method and with the option to build (Real Options Approach).

4.1.3 Sizing of the plant

Coupling a biorefinery with a NPP, without the possibility to work in load following mode, is a "stationary" cogeneration. For stationary, we mean that there is not the possibility to vary the production within a time period, and therefore flows, powers and yields are stable and fixed.

In this configuration, varying the size of the nuclear site wouldn't vary the final economic results. Varying the size of the NPP letting a static size of biorefinery, doesn't mean to vary the quantity of energy to provide to the biorefinery. The not-sold electricity (because of the energy directed to the auxiliary plant) depends only from the capacity of the biorefinery.

However a parameter has to be chosen in order to make a reasonable comparison for example also with the case of coupling with a desalination plant, discussed in 4.2.2. Some parameters can be investigated:

- Total investment cost
- Percentage of electric load lost in order to drive the auxiliary plant
- Other …

In case of coupling with desalination and working in load following mode, about 20-30% of the theoretical electric energy produced is sacrificed to produce water. Therefore only 70%-80% of the theoretical electric capacity is effectively sent to the grid. Hence the same approach is used to decide the size of the biorefinery.

Hypothesizing a site designed with 4 IRIS reactors, 4000 MWt and 1340 MWe could be produced in a "stand-alone" layout. Therefore the maximum capacity chosen for the biorefinery is a total volume of 330000 m^3 for the fermenters. Considering that they are 100 m^3 each, fill for the 80% of their capacity, the cultivation system needs 3300 units. Coupling such a plant with the nuclear site permits to connect to the grid an electric plant of 943 MWe, representing the 70% of the original stand-alone production, as detailed in Table 16. Because of economies of scale in the construction of the biorefinery, it is also interesting to investigate how the profitability of the investment change, varying the capacity of the auxiliary plant. Hence the capacity of the biorefinery is changed in the following range: 110 – 330 million of litres (ML). It corresponds to an electric capacity usage of the NPP in the range 90-70%.

TABLE 16. POWERS USAGE IN CASE OF MAXIMUM CAPACITY BIOREFINERY.

Volume BioRefinery	330000 m^3
NPP – Electric Power	335
NPP - Number of units	4
NPP – eta	33,5%
Thermal Power per unit	1000 MWt
Total Thermal Power of site	4000 MWt
Thermal power to biorefinery	177 MWt
Total Electric Power	1281 MWe
Electric power to biorefinery	338 MWe
Electric power to grid	943 MWe
Theoretical Electric power	1340 MWe
Net difference	-397 MWe
Electric capacity used of the NPP	70%

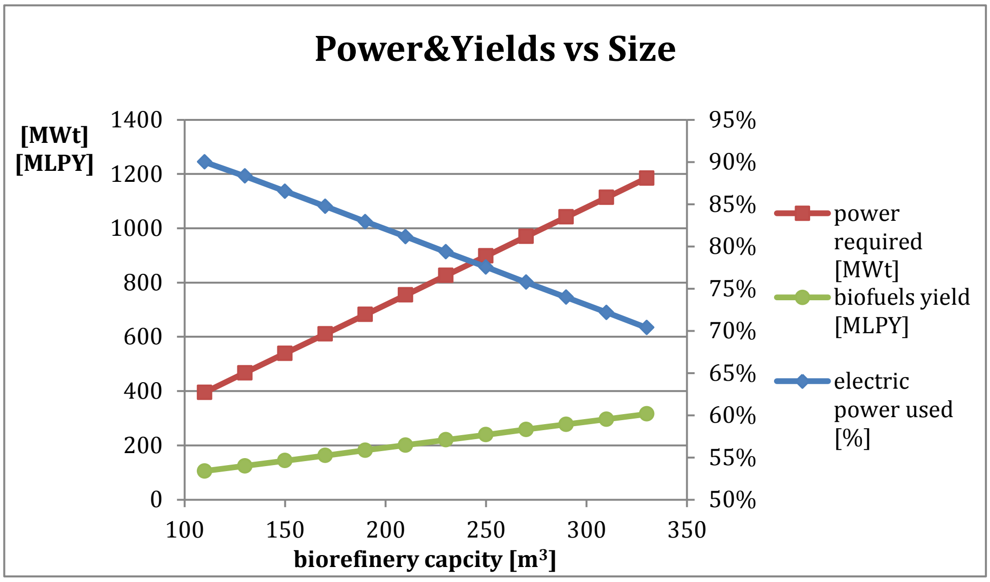


FIGURE 34. HOW POWER REQUIRED, BIOFUELS YIELDS AND USAGE OF ELECTRIC POWER OF THE NPP VARY IN FUNCTION OF THE SIZE OF THE BIOREFINERY.

4.2 DESALINATION

Thermal desalination plants are composed by several modular units. For example, the largest worldwide operating MED-TVC plant produces 800000 m^3/d with 27 units of 30000 m^3/d each (Marafiq, Saudi Arabia). This permits to have large flexibility in the layout of the plant. Moreover to manage the warming up of a smaller unit is reasonably simpler and quicker than a situation in which the plant is composed by a single huge unit. Thermal desalination process is simple from the physical point of view: it consists in an evaporation of water when it reaches the thermodynamic boiling point. It means that if within the stages of the unit, thermal and pressure conditions are above the boiling point of water, the plant starts to produce desalinated water. Therefore, considering the simplicity of the MED-TVC operation and the flexibilities both of the SMRs and the desalination plant, we have assumed the possibility to give to the MED-TVC a variable energetic input within the day. In particular during the low-load hours, MED-TVC is assumed to work at 100% of nominal level, whereas during the high-load hours is planned to work at minimum level.

4.2.1 Load Following: what modifications within the plants

To design the productivity of such a plant, some variables are investigated. In particular we focus on the following questions:

1. What is the minimum electric power level that the NPP has to reach?
2. How much energy has to be directed from the NPP to the desalination plant?
3. Which is the operation range of a steam turbine within a Rankine cycle?
4. Which is the operation range of a MED-TVC plant?

For the first question, the answer is given directly by the requirements established by [7], that declares a nuclear power plant should operate even to the 50% of its nominal electric power level (see paragraph 1.2).

For the second point, Figure 4 represents some current examples of load following made with 6 NPPs in Germany. The most representative graph (KWG plant, yellow line) has been drawn in Microsoft Excel and subsequently imported in Matlab to calculate the area under the graph. Results are shown in Figure 35.

Characteristics of the plants are the following:

- Nominal electric power: 1360 MWe
- Theoretical electric energy produced in one day: 1360 MWe * 24 h/d = 32,640 GWh/d
- Minimum electric power level: 700 MWe (51,5% of nominal power at 2.30 a.m.)
- Effective electricity produced in one day: 25,968 GWh/d (79,6% of maximum theoretical energy)

The questions 3 and 4 are more delicate. When a steam turbine is designed for a Rankine cycle, the stators, rotors and velocity triangles and designed to maximize the efficiency of the machine, given a "by-design" steam flow. In daily operations, the steam flow is not stationary, but can change in order to make load regulation. When the steam flow is adjusted within the turbine, the efficiency of the machine change. In particular the link between the variation of the steam flow rate ṁ (kg/h) with the turbine load during so called "throttle governing" is linear and is given by the "Willan's line", which is a straight line between no load and most economic load [143]. The equation for the Willan's line is given

$$\dot{m} = K + a \cdot P$$

Where a is the steam rate in kg/kWh, P is the load on turbine in kW and K is no load steam consumption.

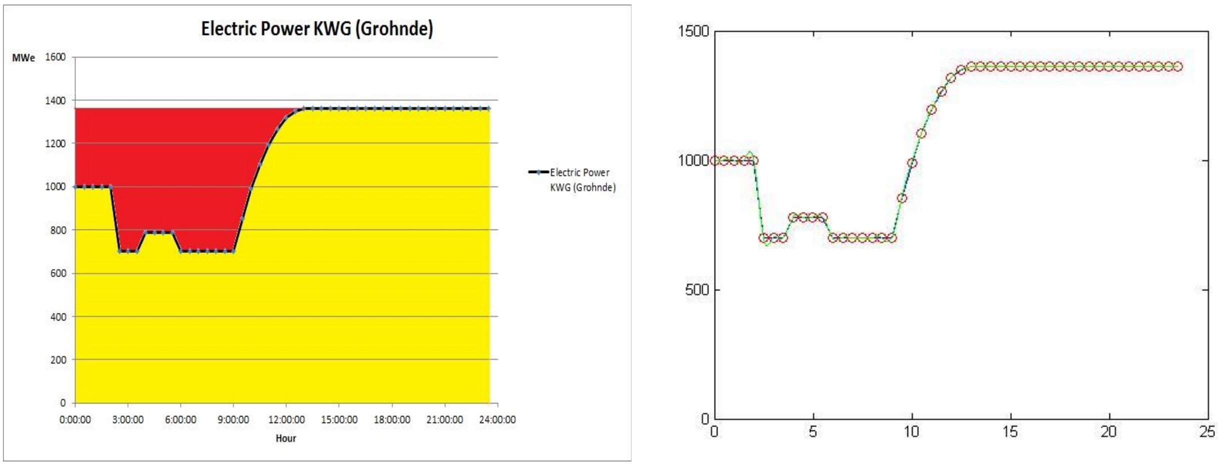
It is interesting to note the presence of the term K into the equation. K is a constant that represent the percentage of steam (say also power) that has to be kept inside the turbine even without producing any energy. Therefore this no load power level must be always provided to avoid overheating of the machine, even if the turbine is not producing electricity.

Typical values of no load constant are within the range 3-10% of the full load consumption of steam. In particular in [144], for a turbine of nominal power of 210 MW, a specific steam consumption of 3,28 kg/kWh is reported, with a no load consumption rate of 15 kg/s. It correspond to a percentage of 7,8%. Even if exact number should be obtained from manufacturer for every specific installation, we assume that the no load power level for the NPP in "off-design" operation (during water production period) is 7,8%.

Similarly, considerations are done to evaluate the eventual minimum energy that has to be provided to the MED-TVC even if the plant is not producing water (during the high-load level hours). Thanks to a literature review and discussions with experts, we can assert that switching-off the plant doesn't comport any technical issue to the plant itself (i.e. no problems of corrosion, overheating, etc.) [145].

Moreover, any MED-TVC can be started in a very short time from the cold situation, let's say about one hour. Most of the time is necessary to drive the vacuum within the stages.

FIGURE 35. ON THE LEFT THE POWER TREND DURING A DAY OF GROHNDE (KWG) PRESSURIZED WATER REACTOR. ON THE RIGHT THE IMPORTED GRAPH IN MATLAB USED TO CALCULATE THE INTEGRAL.



Therefore it is sufficient to keep vacuum during the day for reducing the start-up time down to a few minutes. The vacuum can be kept either by steam ejectors or by a vacuum pump, both sized to vent the air leaking from the gaskets (very small duty to be predicted according to the number of flanges and to the accuracy of fabrication). A reasonable value for air consumption is a flow rate of 400 m^3/h at 170mbar. Vacuum pumps to provide such condition of air can consume 200 kW. Instead, with a two-stage steam ejector, steam consumption is about 1000 kg/h. Both the quantities and the time needed to start-up the plant are negligible in comparison with other amounts of power involved in our calculations [145].

The last technical consideration is about the quality of water. Even if it is possible to keep the MED-TVC readily available with one of the two systems named before, the plant start to produce standard quality water when the power load is in the range 20-30%. It means that the plant produce water, but with high salinity content. The worth of this water is difficultly predictable. Therefore, starting and shutting down the plant in the range 0-100% of the nominal load inserts important losses. Hence in this study we assume to use the desalination plant in the range 25-100% of the nominal capacity. In this way MED-TVC constantly produce an amount of sealable quality water and when it is asked to increment the production it is readily available and no losses have to be considered. This is an important assumption, since the desalination plant is characterized by a high investment cost. For this motivation, like in the nuclear plant, enhancing the availability and the whole production of the plant permits to take full advantage of this rather expensive technology. The minimum level of 25% is only slightly optimistic but it is assumed that further improvements can be introduced in the nearby future.

4.2.2 Sizing of the plant

The size of the desalination plant is dictated by technical limitations, load following requirements and nuclear site's installed power. Technical limitations and load following requirements have been investigated in the previous section. Let's start to consider a site composed by 4 IRIS reactor, i.e. 4000 MWt. If roughly the aim is to direct 50% to the auxiliary plants, the layout of the site is designed in the following way:

- Two IRIS always produce only electricity. They are always connected to the grid, working at full power load
- Other two IRIS instead are connected to the desalination units, as well. As said, desalination plant is composed by several small modular units. As such, the flexibility of the layout is enhanced.

Therefore, considering also the minimum no load level for the steam turbine, 1844 MWt must directed to MED-TVC. Since we assume a thermal energy need of 50 kWh/m^3, having a plant availability of 90%, the size has to be fixed to 885120 m^3/d, a similar size of MED-TVC in Jubail (Saudi Arabia). Therefore even the electric energy required is fixed. In addition knowing the minimum level to provide to the desalination plant (25%), minimum and maximum power levels supplied to the plants and to the grid during the day are found (see Table 17).

For MED_max it is meant the maximum load for the desalination plant, and in turn the production called "off-design", since this way of production is that one designed for the night (low-load hours). The opposite way of production (MED-min) is instead called "by-design".

With these power distributions, during the day 461 MWt and 18 MWe are always supplied to MED-TVC. The electric power to the grid subsequently is 1167 MWe. On the contrary during the low-load hours the layout designed permits to connect to the grid only 596 MWe, that represents the 51% of the nominal power.

Actually with 4 SMRs, 2 of which are connected to auxiliary units, 3 production modes are possible:

- 4 SMRs connected to the grid
- 3 SMRs connected to the grid, 1 to the MED-TVC units
- 2 SMRs connected to the grid, 2 to the MED-TVC units.

This give the opportunity to operate with 3 different power levels. Despite this the general results of the analysis don't vary, because what is important is the whole output of the plants. The latter is studied varying how many hours per day the plant operate in by-design configuration or in off-design configuration. However the target is initially proposed by the red area calculated with Matlab (Figure 35): 80% of the energy is available for the desalination plant. Obtaining this value with 2, 3, 4, n different power levels don't change the final production.

On the contrary installing more smaller nuclear modules (but for example leaving the same thermal power installed), enhances the available power levels, making possible smoother load variations. In general, with n SMRs installed, there are $\left(\frac{n}{2}+1\right)$ power levels.

TABLE 17. THERMAL AND ELECTRIC ENERGY REQUIREMENTS FOR MED PLANT IN BOTH BY-DESIGN AND OFF-DESIGN PRODUCTION WAYS.

ITEM	POWER
P_MED (th)_max	1844 MWt
P_MED (th)_min	461 MWt
P_MED (el)_max	74 MWe
P_MED (el)_min	18 MWe

4.2.3 Daily production

To calculate the revenues of the economic analysis, the yields of production must be known. To complete this goal, the daily production of the cogenerative plant is modelled. As said, just two production ways are considered: by-design and off-design. They are descripted in Table 18.

TABLE 18. POWER DISTRIBUTIONS BETWEEN PLANTS IN BY-DESIGN AND OFF-DESIGN ARRANGEMENTS.

PRODUCTION PHASE	ITEM	POWER	
IRIS REACTOR stand-alone	total thermal power	4000 MWt	
	net electric power	1340 MWe	
	net electric efficiency	33,5%	
	thermal power to condenser	2660 MWt	
		By-design way	***Off-design way***
DESALINATION	thermal power	461 MWt	1844 MWt
	electric power	18 MWe	74 MWe
COGENERATION POWER DISTRIBUTION	thermal power to turbine	3539 MWt	2000 MWt
	gross electric power	1186 MWe	670 MWe
	electric power to the grid	1167 MWe	596 MWe
	electric power sold	87%	44%
	thermal power to tower	2353 MWt	1486 MWt
	thermal power to security condenser	461 MWt	1844 MWt

Now, to know the effective yields of products, the respective working hours must be introduced. In particular, the final aim is to have the used electric capacity of the nuclear plant equals to 80%. To obtain this value, we have calculated with the model that the plant should work for 20 hours per day in the by-design arrangement. With even this parameter fixed, the daily production of water is established, as in Table 19.

TABLE 19. PRODUCTION OF WATER AND ELECTRICITY DURING A DAY.

		By-design way	***Off-design way***
DATA	percentage of day	83%	17%
	hours	20 h	4 h
YIELDS	theoretical electric energy	26805 MWh	5355 MWh
	electric energy to the grid	23347 MWh	2383 MWh
	electric energy lost	3458 MWh	2972 MWh
	water produced	184436 m^3	147372 m^3
DAILY RESULTS	*theoretical electric energy*	*32160 MWh/d*	
	electric energy to the grid	*25730 MWh/d*	
	electric energy lost	*6430 MWh/d*	
	water produced	*331809 m^3/d*	
	used electric capacity_NPP	*80%*	
	used capacity_Desalination	*42%*	

From the tables it is possible to note that the times needed to change power levels are not considered, because the plants are assumed to respond quickly to load variations. Therefore times required are negligible.

Instead, it is also interesting to investigate how the production changes varying some parameters: minimum percentage for the desalination plant in by-design hours, no load percentage for steam turbines and distributions of daily hours between two production arrangements. In particular the direct consequence of the last point is that the used electric capacity of the NPP is varied. Even if nowadays the not sold electricity is about the 80%, this figure is predicted to decrease in the future. Therefore, we investigate the possibility to vary this parameter in the range of 70-80%. Consequently the hours in by-design arrangement range within the interval 60-83%.

FIGURE 36. DEPENDENCE OF USED PLANTS CAPACITY, WITH THE MINIMUM CONSTANT LOAD LEVEL TO SUPPLY TO THE MED-TVC DURING THE BY-DESIGN HOURS. IN BRACKET WITHIN THE LEGEND THE PERCENTAGE OF MINIMUM LOAD, CONSTANTLY PROVIDED TO THE DESALINATION PLANT.

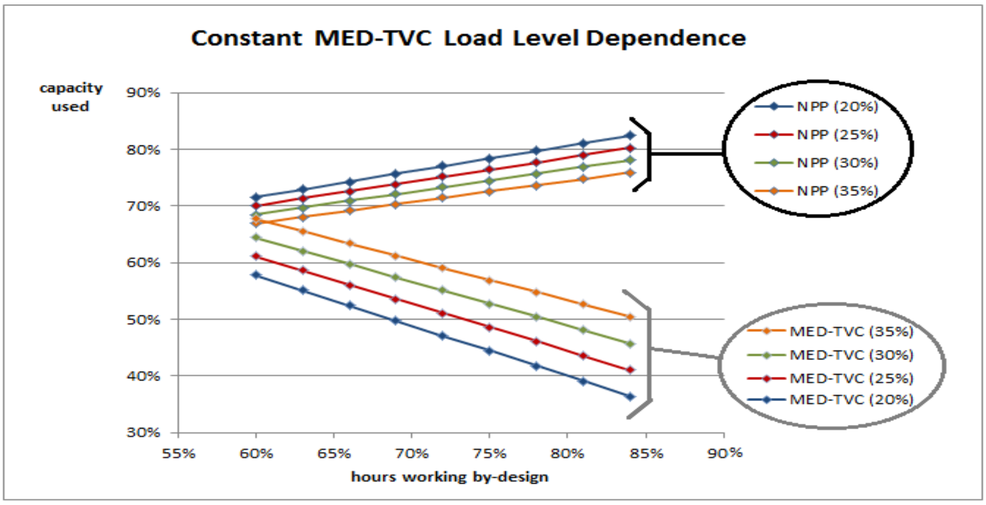
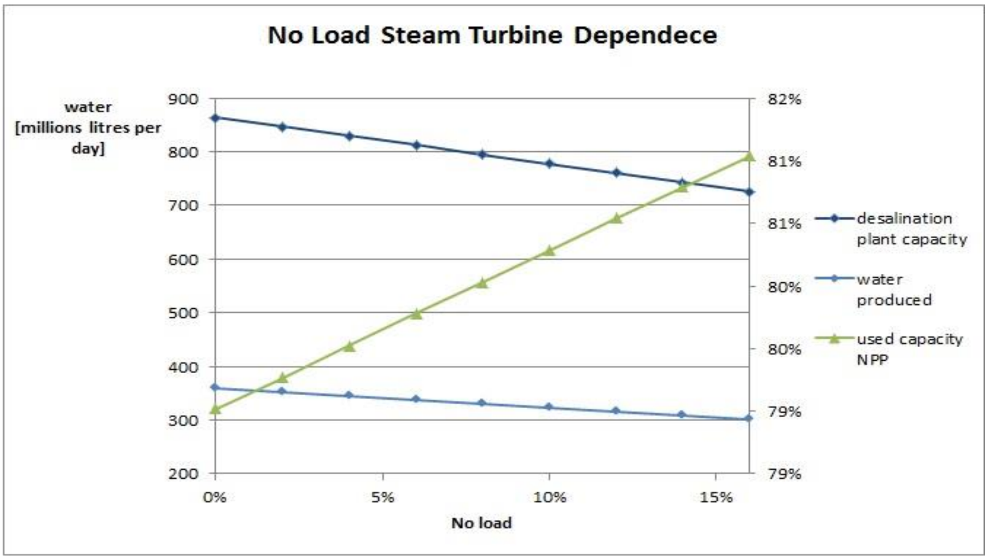


FIGURE 37. DEPENDENCE OF DESALINATION INSTALLED CAPACITY, WATER EFFECTIVELY PRODUCED AND ELECTRIC CAPACITY USED FOR THE NPP WITH THE NO LOAD POWER LEVEL FOR THE STEAM TURBINE.



4.2.4 Comments

To vary the minimum constant level of the desalination plant don't change the size of the plant itself, since the thermal power available remain constant. Instead the total used capacity is affected by this parameter. Figure 36 underlines the dependence of usage's percentage for both the plants. Moreover, the lines that refer to the MED-TVC are wider than that ones that refer to the NPP. It means that this parameter affect most the desalination plant. In this study the minimum level selected is 25% (red lines). Looking at that lines on this figure, it is possible also to see how the used capacities of the plants change in relation to the numbers of hours in which the plant works in by-design arrangement.

Figure 37 instead says that modifying the minimum load to provide to the turbine varies even the plant capacity installed of MED-TVC. Indeed, the thermal power available for the latter is lower. Consequently also the total water produced varies accordingly. In addition the ratio between water produced and installed capacity (i.e. capacity used) remains constant. Similarly the minimum load following level stays on the same value of 51%. This is because the no load level for the turbine directly affect the sizing o the plants and not their operations. Moreover, because of the smaller capacity of MED-TVC, the used capacity of NPP is slightly higher. Indeed, the fact to have a smaller auxiliary plant is largely compensated by the higher degree of power "wasted" to run the turbine without producing energy. For this figure, trends are found with a percentage of minimum load for MED-TVC of 25%, and planned operation of 20 hours of by-design arrangement.

Therefore, with all parameters fixed (minimum load levels for plants) and static planned operations, it is possible to make a fixed, rigid load following. The aim of such fixed operation mode is to satisfy the grid requirements (minimum power level, about 51%, and used NPP's electricity capacity used, about 80%). This rigid load following is a non-static operation mode, but is fixed and therefore not flexible. It is analysed with an option to build, because the manager has only to opportunity to decide if build or not the plant. When the plant is built, it is going to operate every day, constantly, with its fixed, non-static operation mode.

Instead, if the freedom to change the operation mode is given to the manager, he can capture the best market prices every day. In this case we refer to a flexible load following. The minimum load following level (51%) remains the same, because it is determined by the discussed sizing of the plant. Instead presumably the level of electricity produced is going to vary, because the operation of the plant are not fixed, but can change accordingly to the prices of the market during the day. Indeed, if water has a fixed price during the day, the same thing is not true for the electricity. Electricity has a flexible daily price. Some days it could be more profitable takes advantage of MED-TVC, some others of NPP. This possibility to switch the production between two products, it is analysed with an option to switch.

4.3 COUPLING WITH A NPP

In comparison to a NPP in a stand-alone layout, the coupling with auxiliary plants in a cogenerative layout requires some modifications.

1. First of all, the auxiliary plants do not require a boiler or any other kind of source of heat, because the steam comes directly from the nuclear plant
2. Consequently, as discussed in paragraph 3.3, O&M costs don't include the purchase of energy, like electricity and natural gas to burn into the boiler
3. On the contrary, the coupling with a power plant requires a heat exchanger and additional protective barriers to avoid and prevent potential carry-over of radioactivity to the products [146]
4. Finally, since a percentage of the produced steam is constantly directed to the auxiliary plant, both the low pressure turbine and the condenser of the secondary cycle are smaller.

Points 1, 2 and 4 lead to a decrease of overnight capital costs and O&M of the plants, whereas point 3 lead to an increase.

To give a value of the saving of capital and operative costs, regarding points 1 and 2, the analysis in made by a literature review in section 5.4.

Point 3 refers both to the heat exchangers that have to be installed to provide heat to the auxiliary plants and the introduction of an additional condenser that would allow to condensate the steam sent to the cogenerative heat exchangers in case of failure of the biorefinery and/or desalination plant. Indeed, if a fault or unexpected maintenance in one of the cogenerative plants occurred and its operations were stopped, the heat exchangers would not allow to condensate steam drawn from the turbines, since no cooling water would enter the heat exchanger. Thus, a "security condenser" that works only in the case of an "emergency" is needed.

Cost of standard IRIS cooling tower condenser, which has to exchange 665 MW, is assumed 12,5 M$ [147]. Referring to this value, the cost of the "security condenser" has been estimated (assuming an exponent of 0.9 to account for the economies of scale). For example, for the coupling of 4 IRIS with the desalination plant, the security condenser would be sized for a thermal power of about 1860 MWt, and its cost would be 30,3 M$.

Cost of heat exchangers for the coupling between plants were calculated with proportional method, considering that the value of heat exchanger in a similar work has been assumed 3-5% of the total overnight capital costs of the corresponding plant [125].

Finally to estimate the decrease of costs due to the smaller low pressure turbine, and a smaller condenser for the secondary cycle, we need to know the quantity of power directed to the auxiliary plants. The saving of cost for the turbine is been assumed of 320 $/kW [148]. Thus, if the power of the turbine was 100 MW less than the standard one, the overnight cost would decrease 32.0 M$. Amount of power saved is easily calculated knowing the size of the plants.

To assess the cost reduction for the smaller IRIS cooling tower system, as already discussed for the case of "additional condenser", economies of scale are considered (assuming an exponent of 0.9). For example, if thermal power to exchange in the condenser was 200 MW less than the standard value (665 MW), the overnight cost would decrease of 4 M$.

CHAPTER 5.
THE ECONOMICAL ANALYSIS

In the fifth chapter, the main items involved for the economic analysis are introduced. At the beginning of the chapter, a brief literature review of economic analysis methods is presented. In the following a comparison between methods and the innovative features that characterized the Real Options Approach are discussed. Subsequently, the chapter lists all involved data regarding costs, that are used to sort the economic analysis. At the end of the chapter the algorithms developed in this dissertation in order to study the profitability of the investments are deeply explained.

5.1 THE DCF METHOD

In capital budgeting the traditional methods are based on the Discounted Cash Flow (DCF) analysis framework, that is an investment valuation methodology focused on the time value of money. There are literally hundreds of DCF models, but they all based on the same foundation that simply involves calculation of the Net Present Value (NPV) of a project over the entire life cycle (say T, e.g. 40 years), accounting for the investment costs and the production phase free cash flows.

The method that financial analysts use to value projects may range from the simple to the sophisticated, but the building blocks are the same. Any valuation starts with estimation of costs and revenues over the project life. Because of the time value of money, each cash flow from the future is converted into today's value, using the formula:

$$PV_t = \frac{FV_t}{(1+r)^t} \qquad (3)$$

where FV is the future value of the cash flow, PV is the present value, r is the discount rate per time period, and t is the number of the time period.

The project NPV is simply the summation of the PVs of all the cash inflows and cash outflows:

$$NPV = \sum_{t=0}^{T} PV_t = \sum_{t=0}^{T} \frac{FV_t}{(1+r)^t} \qquad (4)$$

Then the optimal investment rule is to proceed with a project if its NPV is greater than zero and, in case of a portfolio with two or more different projects, the priority will be given to that one with the larger NPV [73].

Another famous decision parameter that belongs to the DCF approach is the Internal Rate of Return (IRR). This profitability indicator is the discount rate at which the NPV became equal to zero. Greater is the IRR of a project, more attractive is the investment.

In the energy field, one more DCF based method is the Levelised Cost of Electricity (LCOE) approach, that allows to compare the costs, per energy produced, of different energy supply technologies. It is for example useful for an investor that has to decide between a nuclear power plant and a CCGT (Combined Cycle Gas Turbine), or between different projects for two or more SMRs. LCOE is calculated in this way:

$$LCOE = \frac{\sum_{t=0}^{T} \frac{R_t}{(1+r)^t}}{\sum_{t=0}^{T} \frac{E_t}{(1+r)^t}} \qquad (5)$$

where R_t is the expenditure at time t (including investments, O&M, waste disposal, fuel costs, and so on) and E_t is the amount of electricity produced at time t.

As it is clear from the equations, the discount rate is a fundamental parameter to determine for the analysis. It can modify the final decision about an investment, because it directly changes the NPV. Actually, the meaning of the discount rate is the rate that is used to convert the future value of the project cash flow to the present value. It is adjusted for the risk perceived to be associated with the project: the higher the risk, the higher the discount rate. Indeed, business is basically about taking risks. The higher the risk and the higher the returns that investors will expect. In the finance world, risks are broadly classified as market risks and private risks. To give a rough definition, it is generally accepted that risks that can be captured in the value of a traded security are *market risks* and all the other are *private risks* [149, p. 35].

Thus, one of the biggest dilemmas an investor faces is what discount rate to use in the NPV calculations. If there is not uncertainty at all on a cash flow stream, the appropriate discount rate is a *risk-free rate*. If there is uncertainty, the next consideration is whether the stream is influenced by market or private risk. If it influenced by private risk, the investor will not pay a risk premium for the ineptness of the organization in completing the project or the ineffectiveness of the technology involved. On the other hand, if the cash flow is subject to market risk, one would account for it in some fashion, mostly common by adjusting the discount rate. Actually, in the real world it is difficult to completely separate the private risk from market risk. Furthermore any project investments requires capital, and organizations have to pay a cost to obtain capital. Therefore for discounting the cash flows that are subject to private risk, practitioners use a rate that is ether slightly higher than the risk free rate or a rate that is commensurate with the organization's *weighted average cost of capital (WACC)*. It is calculated in the following way:

$$WACC = W_d \cdot C_d \cdot (1 - tax) + W_e \cdot C_e \qquad (6)$$

where W_d and W_e represent how the stakeholders share the financing of an investment. The subscript *d* stays for debt, while the subscript *e* stays for equity. Cost of capital represents the cost of financing an organization's activities, which is normally done through some combination of debt and equity. The debt includes banks, financing agencies, public investments and so on. Their gains are predefined by contract and they have the priority for the investment recovery. On the other hand, the equity represents shareholders in general. Their investment is more flexible and risky because they don't have the priority for the investment recovery and their remuneration is not guaranteed, especially in case of bankruptcy. For these reasons, the cost of equity (C_e) is higher in comparison with the cost of debt (C_d). WACC can be used as a proxy to represent the private risks related to project investments cost. Costs of capital are the rate expected by the banker (interests) or by the investor (returns).

Therefore, the DCF methodology, although is simple and easy to implement, presents three substantial criticalities:

4. The discussed choice of the discount rate
5. The weak consideration of stochastic nature of the cash flows: DCF can't capture any market uncertainties, like electricity and fuel prices, or technical uncertainties, like construction costs that vary considerably along construction time [36]
6. The assumed passivity of the management, unable to improve the results after the resolution of some uncertainties [37]

These issues suggested in the year the creation of a new tool for economic analysis, to overcome to the problems of uncertainties and flexibility. This new tool is represented by the *Real Options*.

5.2 WHAT IS A REAL OPTION?

The conventional tools of business analysis (DCF, for example) don't consider the uncertainty of a cash flow. Uncertainty is converted only into a risk. That risk enters in the discount rate decision, like a well fixed value. Their approaches still remain deterministic, even if they consider risks. Real option approach looks at the uncertainty in a different way. Real Option Analysis is a dynamic approach to the uncertainty that supply a worth to the opportunity to have an option. ROA is a tool that give us the possibility to quantify the value added to the project due to the option.

Before explaining the approach and what real options are, it is better to clarify where real options come from.

5.2.1 Real Options and Financial Options

Financial options are contracts between two parties for a future transactions on an asset (defined as the *underlying*) at a reference price (defined as the *strike price*, or *exercise price*). The option price, or *premium*, is the price paid to acquire the option. The owner of the option has ***the right, but not the obligation***, to engage in that transaction, while the seller incurs the corresponding obligation to fulfil the transaction if the option is exercised, and receives in return the exercise price.

There are two basic types of options: calls and puts. A *call* option gives its holder the right to buy an underlying asset while a *put* option gives the holder the right to sell it, both at the strike price.

Also, financial options can be categorized by the time when they can be exercised: *American* options can be exercised at any time up to the expiration date, while *European* options can be exercised only on the fixed expiration date.

When a financial option is exercised its relative *payoff* is obtained. In call options this is the difference between the value at the exercise time t of the underlying S_t and the strike price X,

$$P_t(call) = S_t - X \tag{7}$$

while in put options it is the contrary,

$$P_t(put) = X - S_t \tag{8}$$

This is then a zero-sum game, since the gain to one party is a loss to the other side. Using a financial vocabulary, it is interesting to give some definition:

- if there is an advantage to exercise the option, the option is said *in the money.* A call option is in the money when $S_t - X > 0$, while a put option is in the money when $S_t - X < 0$.
- if there is not an advantage to exercise the option, the option is said *out of the money.* A call option is out of the money when $S_t - X < 0$, while a put option is out of the money when $S_t - X > 0$.
- if the underlying asset value is equal to the exercise price at the moment of the maturity, the option is said *at the money*

In addition, financial options are characterized from an asymmetry of return when options are exercised, as nobody will exercise an option with a negative payoff. For example, the owner of a call option, has a certain probability to gain the payoff, if S_t increases significantly during the period of the option and exceeds X. In the opposite scenario, if at the expiration date the worth of S_t is lower than X, the possible loss will not be "symmetric" to the possible gain, as the holder will not exercise the option. Hence, at expiration date, the value of an option (say C) is the maximum between zero and the payoff conditional to the exercise:

$$C = \max(0, P_t) \qquad (9)$$

This asymmetry implies that the higher the volatility of the asset the more valuable the corresponding option, as the possible gain will increase while the possible loss will remain constrained to zero.

Thus, financial assets are primarily stocks and bonds that are traded in financial markets. The options for most of these assets are listed on exchanges such as the Chicago Board Options Exchange and the American Stock Exchange. Real assets may include real estate, projects, and intellectual property, most of which are not usually traded. In this chapter will be explained that recognizing uncertainties and flexibility in a project is very important to find some "options", that in this case are "real" because they are related to real projects of the real world. For example, an option could be to invest in a new idea, or to abandon an existing design. In Table 20 there is a confront between financial options and real options. In the end, in Appendix C. there are two simple examples of a financial option and a real one.

5.2.2 DCF vs ROA

In this section will be clarified the difference between a Discounted Cash Flow approach and a Real Option Analysis (ROA). The calculation of the NPV is based on a set of fixed assumptions related to the project payoff (a deterministic approach), whereas the payoff is uncertain and probabilistic. DCF does not take in consideration the contingent decision available and the managerial flexibility to act on those decisions. For example, the value of the future flexibility to expand, contract, or abandon is not captured by DCF. Furthermore, DCF analysis accounts for only the downside of the risk without considering the rewards. This inherent bias leads to rejection of highly promising projects because of their uncertainty. Many of today's technology projects exhibit such characteristics; therefore, the limitations of DCF are of enormous significance in their valuation.

On the other hand, ROA offers new ways to address these limitations of DCF. To be sure, ROA is not a substitute for DCF. It is an auxiliary tool that fills the gaps that DCF cannot address [38]. ROA uses DCF as a building block and captures the value of the options. It integrates traditional valuation tools into a more sophisticated framework that provides practicing financial analysts and decision makers with more complete and meaningful information. Whereas real options embedded in projects are implicitly recognized by organizations, the formal valuation using ROA makes them explicit and quantifies their value, thereby helping management make rational decisions [149].

The key advantage of the real options is that, if properly managed, options can create an extra value and reduce risk for the investors that own them [39]. DCF accounts for the downside of a project by using a risk-adjusted discount rate. ROA, on the other hand, captures the value of the project for its upside potential by accounting for proper managerial decisions that would be presumably be taken to limit the downside risk.

TABLE 20. COMPARISON BETWEEN FINANCIAL AND REAL OPTIONS [149, PP. 6-7]

ITEM	FINANCIAL OPTIONS	REAL OPTIONS
Option price	Price paid to acquire the option, which is fixed by the financial markets	Price paid to acquire or create the option, keep it alive, and clear the uncertainty (for instance, price paid to acquire a patent, maintain it, and conduct market research to identify its potential). The option price is not fixed (for example, the price to buy a patent is negotiable).
Exercise price	Price paid to buy/sell the underlying stock; a fixed value defined in the option contract	Cost of buying/selling the underlying real asset (e.g., the cost of commercializing a new technology is a call option exercise price, the underlying asset being the profits from the commercialization; the selling price of abandoned manufacturing assets is a put option exercise price, the underlying asset being the manufacturing assets).
Expiration time	Defined in the options contract and is clearly known	Clearly known in some cases (e.g., leases may be signed on oil fields involving options on drilling) and not so in others (e.g., for technology projects, it depends on the market conditions and competition).
Timing of payoff	Immediately after the options are exercised; basically instantaneous	Only after some time since the option has been exercised. May be spread over a long period of time: after a decision is taken to commercialize a new technology, the commercialization itself takes months, and the profits from the sales are spread over a period of many years.
Option holder's control on its value over the option's life	None	Proper management action can increase the option value while limiting the downside potential. For example, the holder of a new, novel technology option can invest in developing other complementary technologies, increasing the value of the original option.
Option value as a function of option life	Larger for longer life of a given option	Larger for longer life of a given option, especially related to patents and property with exclusive rights. But with many options, the asset value may be diminished because of entry of competition, thereby bringing the option value down.
Option value as a function of the underlying asset's volatility	Increases	Increases
Resolution of uncertainty	Uncertainty clears automatically with time; the option holder has to do nothing to clear it	Uncertainty clears through time in some cases. In most cases, the option holder needs to actively invest in clearing the uncertainty, for instance, through market research or pilot testing.
Liquidity and tradability of the option	Liquid and tradable in financial markets	Most often neither liquid nor tradable.
Rationality behind the exercise decision	Mostly rational; dictated by the numerical difference between the underlying asset (e.g., stock) value and the exercise price	Exercise decision may have political and emotional implications (e.g., abandonment of a long-term project with a large team).

Discounted Cash Flow	Real Options Analysis
Does not capture the value of managerial flexibility during the project life cycle	Recognizes the value of managerial flexibility to alter the course of a project
Uncertainty with future project outcomes not considered	Uncertainty is the key factor that drives the option's value
Undervalues the asset that currently (or in the near term) produces little or no cash flow	The long-term strategic value of the project is considered because of the flexibility with decision making
Expected payoff is discounted at a rate adjusted for risk. Risk is expressed as a discount premium	Payoff itself is adjusted for risk and then discounted at a risk-free rate. Risk is expressed in the probability distribution of the payoff
Investment cost is typically discounted at the same rate as the payoff (risk-adjusted rate)	Investment cost is discounted at the same rate as the payoff (risk-free rate)

TABLE 21. MAIN DIFFERENCES BETWEEN DCF AND ROA [149]

ROA is most valuable when there is high uncertainty with the underlying asset value and management has significant flexibility to change the course of the project in a favourable direction and is willing to exercise the options.

In addition, ROA does not provide much value in investment decisions on projects with very high NPVs, because the projects are already attractive for investment and the additional value that may be provided would not change the decision. Similarly on projects with very low NPVs. If a project has a very high negative NPV, it probably should be rejected. Trying to justify such a project using real options would be a meaningless exercise. Real options offer the greatest value on projects with an NPV close to zero, high uncertainty and high managerial flexibility [149]. In Table 21, are reported the main differences between the two methods.

Moreover, as shown in the example in the Appendix C., the real option approach can add an extra value to the calculation of the NPV. The extra value simply comes from some kind of uncertainty or flexibility, directly proper of every specific project under valuation.

Finally, it is possible to reassume that, "*...whereas DCF provides a fixed path for investment decisions, ROA offers a strategic map that outlines the contingent decisions, especially these related to private and market uncertainties. ROA is not a substitute for but rather an extension of the DCF method. Every real option valuation starts with the underlying asset value, which is the expected payoff calculated using the DCF method, where risk premium is added to the discount rate to account for the uncertainty. Adjustments are then made to this value, taking into consideration the contingent decisions*" [149, p. 60].

Therefore, having seen the great flexibility that characterize a cogeneration plant in general, and the cogeneration plant of this study in particular, for this work has been thought that the real option approach could capture the best worth of the plant.

5.2.3 Why ROA in energy field

The real option approach is not necessary in every investment evaluation: it is more accurate as considers uncertainties but also requires a deeper analysis, requiring to model uncertainties and to implement in the model a real options valuation method. In literature, one of the first real options applications were done in the energy industry. [150] analysed the several reasons of such an huge literature concerning energy related problems:

- This industry is characterized by highly intensive capitals (e.g. the construction costs of a NPP) [151]
- Often these assets present various operational flexibilities
- Mostly the outputs of the industry are traded commodities
- Typically this sector presents an engineering culture suitable with the mathematical complexity of real options models

Finally, as discussed in the next section, there are some features of the plant evaluated in this study that perfectly match with the possibility to use a real option analysis.

5.2.4 Examples of Real Options

Options can be grouped into two basic categories: simple options and compound options. The most common simple options include [38]:

- Option to defer: the possibility to wait to take some irreversible decisions
- Option to build: the possibility to build and invest in a new project
- Option to abandon: the possibility to abandon current operations permanently if market conditions became extremely unfavourable
- Option to expand, contract, or extend the life of facility: the possibility to increase capacity if it is profitable
- Option to temporarily shut down the production process: the possibility to stop and then start again the operations if they are become profitable
- Option to switch: the possibility to change products, processes or input

Option to expand and option to defer are simple American call options, while option to contract and option to abandon are simple American put options. Option to choose and option to switch include both American calls and puts. There is another group of options called rainbow options, which may be either simple or compound. Options for which multiple sources of uncertainty exist are called rainbow options. The uncertainty may be related to one or more of the input parameters used in options valuation or to the individual components that make up an input parameter, or there may be changes in the uncertainty itself over the option lifetime.

On the other hand, compound options are "options on options" [152]. When real options are independent the value of the portfolio of options is the sum of the values of the simple options since they don't interact. This is the best case, because in other types of interactions the value of the whole portfolio will be always lower than the sum of the simple options, considered one per time.

Regarding this work, the possibility to build a new plant to couple with the nuclear central represent a clear option to invest. In the scenario considered for this study, the only certain thing assumed is the construction of the power plant. Using ROA helps us to answer to the following question: does coupling a plant in cogeneration layout with NPP add an extra worth for the investors? Therefore, to analyse the investments in a biorefinery and in a desalination plant, the option to invest can capture the uncertainties that, for example, are related to the prices of products (electricity, water, ethanol, biodiesel, glycerine and DDGS).

In addition, the possibility to work in load following mode with the desalination plant confers a great flexibility, because in this case not only the prices of the products but even the quantities are uncertain. Thus, to better evaluate the coupling with the desalination plant, also the option to switch has been considered.

In paragraph 5.5 the option to build and the option to switch are better explained, also with their advantages and the algorithms used to quantify themselves.

5.2.5 Criticalities in Real Options

Even if this modern approach pretend to solve many issues related to the risk of an investment, some critics disagree on the applicability of the options-based approaches to the real world.

"Critics of options-based approaches to valuing and managing growth opportunities often point out that there is a world of difference between relatively simple financial options and highly complex real options. [...] They are right about the differences but wrong to assume that they are insurmountable." [153].

In particular, the main criticalities arise from the fact that the real world contains many unpredictable variables that cause different scenarios. For critics these labile variables can't be summarized in short mathematical formulas, and their flexibility can't be captured by financial-options-based method.

In the next paragraph, the different methods currently adopted to evaluate an investment with ROA will be presented, together with their limits and advantages. In this study, not a simple financial-option-based method will be used, but a more complex model, that can capture many uncertainties thanks to Monte Carlo simulations.

5.3 THE ROA METHODS

As for financial options, even real options approach require a valuation tool able to calculate the worth of the options. Classical DCF methods (e.g. simulations or decision trees) are easy to understand and require simple mathematics. ROA is far more complex and requires a higher degree of mathematical understanding [149]. Real options valuation models are usually based on models thought for the valuation of financial options. This fact presents the evident advantage to have at disposal a verified and confirmed literature as support. On the other hand real options present more intrinsic uncertainties than financial options [154] and more complicated interactions between them [155]. For this reason it is necessary to create more specific methods for real options.

The choice of the valuation method depend on the complexity of the problem: higher the complexity of the problem, higher the complexity of the method.

Roughly, the valuation techniques for real options can be divided in three classes: Partial Differential Equations (PDE), lattice and simulations. PDE can be solved with Closed-Form models, using for example Black-Scholes or other similar formulas, analytical approximations or numerical methods, like finite difference method. Lattice involve the creation of matrix that can be binomial, trinomial, quadrinomial or, more in general, multinomial. Instead simulations are based on Monte Carlo methods. The techniques and main methods adopted to valuate real options are summarized in REF TO TABLE, together with their limits and advantages.

TABLE 22. ADVANTAGES AND LIMITS OF THE MAIN MODELS USED IN REAL OPTIONS APPROACH.

Technique	Method	Advantages	Limits
PDF	Closed-form models (Black-Scholes equation)	Incorporate volatility changes Easy to use Widely used in financial option	Difficult to explain, because of its mathematical complexity It refers only to European financial option, whereas the real option are closer to American options. For the reason the equation has to be adapted It uses a lognormal distribution of the underlying asset, which may not be true for cash flows related to real assets. The value of underlying asset is continuous, not accounting any drastic ups or downs (jumps). Not consider leakages It considers only one strike price, which can change for a real option during its life. Even if some limitations can be overcome by making adjustment, the already complex model would become even more complex
PDF	Finite Difference Method	Accurate Efficient	Very complicated Require unquantifiable time to resolve the equations
Lattice	Binomial	Volatility and strike price are easy to change over the option life More flexibility Transparency in its underlying framework Easy to demonstrate in management meetings Jumps and leakages can be easily add to the model	More approximation involved (less accurate) Requires higher time increments to reach a good approximation (at least 6 time steps)
Simulation	Monte Carlo	Very accurate method	Closer to European option High computational power required for American option

To solve PDF, Closed-Form models are the best approach when they are available, as they are extremely fast to compute and generate exact solutions. The most famous example of these methods is the Black-Scholes formula that derives the price of an European option written on a single underlying asset [156]. These methods cannot be used with complex real option problems because, even if analytical approaches are exact, quick and easy to implement, they depend on very strict assumptions that can lead to an unrealistic model [157].

When there is no analytical solution, numerical methods must be used. The most famous method is the finite difference method, that transforms the PDE in difference equations over a small interval, discretizing all state variables. Broadly the technique of using PDE is not suitable for complex real options because it would be extremely difficult to find the relevant PDE for the problem, as options interact between each other, and for the phenomenon called "curse of dimensionality". Therefore it is very demanding, and it becomes computationally hard (and sometimes simply impossible) to generate each combination of values.

Other numerical methods are based on very different techniques, like lattice and simulations. They can work through two opposite logic:

- Forward induction. It is an approach that values a function by unfolding uncertainty as it evolves from the past. It's recommended when present cash flows do not depend on future events.
- Backward induction. It works backwards in time, from the last time the decisions might be taken to the first time period of the evaluation. It determines the optimal time to exercise options considering at every step what to do.

Binomial (or multinomial) trees and lattices are methods based on the assumption that the stochastic variables that define the state of the model at every time step can assume only a finite number of values (two in binomial case, three in trinomial case, etc.). For example, in the binomial tree at every time step the value of the state variable will move up or down by a specific factor with a certain forecasted probability. Once the tree (or lattice) is built the option's value are calculated firstly at each final node and then at earlier nodes through dynamic programming (see Appendix D.). This method, firstly proposed by [158], is simple to understand and easily implementable with one risk driver. On the other hand it involves some approximations in the creation of the model (as the discretization of the possible values that could be assumed by the price of electricity, for example) that lead at too simplified results. To analyse a complete problem (with many state variables and uncertainties) the number of nodes required, and then the time to generate the tree, grows exponentially with the number of state variables [159].

The Monte Carlo (MC) simulation is based on the idea that simulating state variables' trajectories can approximate probability distribution of terminal asset values. For every simulation, a defined number of paths is generated, sampling the values of the stochastic processes. Since it's computationally heavy, it's indicated for complex cases with many sources of uncertainty. MC simulation is only the base for several methods: for example, the Least Squares Monte Carlo (LSMC) and the Simulation with Optimized Exercise Thresholds (SOET) are two different algorithm based on MC method. LSMC is a method developed by [160] that creates scenarios with MC simulations, but decides the exercise of the options backwards, with dynamic programming (allowing the valuation of American options) (see Appendix D.). The inputs of this method are both deterministic (e.g. the capacity of a power plant) and stochastic (e.g. the price of electricity) and the result of the model is the expanded NPV. Instead SOET method is still based on the MC simulations, but then the core of this method is the evaluation model that generates the distribution of the output variables (e.g. the NPV) from the inputs, both deterministic (e.g. the discount rate), and stochastic. Indeed in the evaluation model, a threshold is created to permit the exercise of the options: every time that, during the MC simulation, the threshold is reached (say for example the price of electricity overtakes a fixed value, e.g 9 $cent/kWh), the option is exercised and the investment results are recorded [74]. Therefore, the most important aspect to clarify is that there are several method based on MC simulations. In regard to the kind of options with are dealing with, an appropriate algorithm can be used or built.

In this study there are many variables that can bring to positive or negative NPV. They are for example the prices of all the products involved: electricity, water (for desalination case), biodiesel, bioethanol, glycerol, DDGS (for biorefinery case). Moreover, even the costs of the plants can represent a source of uncertainty, and the O&M costs that affects the annual cash flows, as well. Because of the great complexity that surrounds the decision about the investment in such a plants, in this study Monte Carlo simulations have been used to simulate the following uncertainties:

- Price of electricity
- Price of water
- Price of biodiesel
- Price of bioethanol
- Price of glycerol (80% pure grade, see Appendix A.)
- Price of DDGS (see Appendix A.)
- Capital Cost of desalination plant
- Capital Cost of microalgae cultivation plant
- Capital Cost of microalgae processing plant
- Capital Cost of biodiesel plant
- Capital Cost of bioethanol plant
- O&Ms for every plant just listed

Monte Carlo method gave only the possibility to simulate the future. Then, a whole model has been created in this thesis, both for desalination case and for biorefinery case. The models used for every proper case permit to the investors to decide if proceed or not in the construction of the plants, and the relative forecasted NPVs. In paragraph 5.4 the data used for the economic analysis will be introduced, while in paragraph 6.1 and in paragraph 5.5, the models created to evaluate the investments will be explained, for the biorefinery and for the desalination plant, respectively.

5.4 DATA FOR THE ANALYSIS

In this chapter are reported all data inserted in the model in matter of costs. As it happened for the power requirements, a comparison between costs reported in literature is done. Costs needed to be introduced regard all the production plants, except the SMR reactor. Indeed we assume that the construction of the nuclear site is already planned and confirmed. With this work we want to investigate only if it is possible to enlarge the investment's worth, taking advantage of the wasted energy. Therefore the suitability of the investment in the nuclear reactors isn't discussed and consequently overnight costs and operation and maintenance costs (O&M) for the SMRs are not considered. In the following, analysed costs are gathered into 2 main plants: cultivation and biodiesel plant (that involve all processes from the cultivation of algae up to the production of biodiesel) and bioethanol plant. Actually costs for the desalination plant are already introduced in paragraph 3.3 Costs in desalination. At the end of the chapter a particular section is also dedicated to the modifications that occur to couple the SMRs with the auxiliary plants (i.e. heat exchanger, security condenser, etc.).

Before reporting data it is important to specify that all data belonging to past papers are escalated to the present value, through an escalation index. For escalation index a weighted average between Power Capital Costs Index (PCCI), Chemical Engineering Plant Cost Index (CEPCI) and Marshall and Swift index (M&S) is done. In particular, considering the chemical nature of the plants, more importance is given to CEPCI (50% of weight), whereas the rest is shared between PCCI (30%) and M&S (20%).

5.4.1 Cultivation and biodiesel costs

In literature for the cultivation costs more data are available for the open ponds layout. However, for the assumptions debated previously, here they are not analysed. Only the fermentors system is considered. In Table 23 the list of data collected in matter of costs for cultivation, processing (dewatering and oil extraction), and biodiesel production phases is reported. Whereas in Table 24 the same specific costs are escalated to the present value with the also indicated escalation index. Data just reported in tables don't consider items like boiler, or purchase of gas and electricity because in our layout they are not needed. In this study data reported by [161] are used because they are more detailed and complete, they usually match also data of other references (especially the capital costs) and they are more recent than others. In particular for cultivation phase data are adapted to volume of fermenters, while for processing and biodiesel they refer to litres (of biodiesel) per year produced.

TABLE 23. CAPITAL COSTS AND O&M FOR CULTIVATION, PROCESSING AND BIODIESEL PHASES. IN PROCESSING ARE INCLUDED THE STEPS OF HARVESTING, DEWATERING AND OIL EXTRACTION.

REFERENCE		[26]	[161]	[162]	[163]
Year		2009	2012	2006	2008
Capacity fermenters [ML_ferm]		1,2	69,75	-	-
Capacity biofuels [MLPY]		1,34	11,3	37,8	9
Yield [L/L]		112%	16%	-	-
Cultivation	Capital Cost [M$]	2,8	150	-	-
	O&M [M$]	-	7,6	-	-
	O&M [$/L]	1,28	0,11	-	-
	Capital Cost [$/L]	2,33	2,15	-	-
Processing	Capital Cost [M$]	-	56,7	-	-
	O&M [M$]	-	2,9	-	-
	O&M [$/LY]	-	0,26	-	-
	Capital Cost [$/LY]	-	5,02	-	-
Biodiesel	Capital Cost [M$]	-	4	10,5	1,35
	O&M [M$]	-	0,202	2,33	0,85
	O&M [$/LY]	-	0,018	0,062	0,094
	Capital Cost [$/LY]	-	0,354	0,278	0,150

TABLE 24. ESCALATED COSTS FOR THE 4 REFERENCES REPORTED IN TABLE 23.

PRESENT VALUE COSTS		[26]	[161]	[162]	[163]
escalation index		1,1272	1,1116	1,2394	1,0611
Year		2013	2013	2013	2013
Cultivation	O&M [$/L]	1,443	0,121		
	Capital Cost [$/L]	2,630	2,391		
Processing	O&M [$/LY]		0,285		
	Capital Cost [$/LY]		5,578		
Biodiesel	O&M [$/LY]		0,020	0,076	0,100
	Capital Cost [$/LY]		0,393	0,344	0,159

In addition, to investigate the optimal capacity of the biorefinery, economies of scale are used (exponential factor equal to 8).

5.4.2 Bioethanol costs

Data reported in literature for capital cost and O&M regarding the production of ethanol are listed in Table 25. Data of column (A) represent the whole capital costs, including costs of boiler and milling process. In [125] the cost of boiler is assumed to be the 2,8% of the entire cost, while in [164], the cost of the beginning milling process accounts for the 7,3% of the plant. Both of them are saving items since they are not needed. In particular the milling process is not required in this stage because the biomass arrives from the oil extraction process, after been already passed through a grinding phase.

In addition it is possible to see in the table the existing strong economies of scale. In particular in [165] is reported an equation that links the specific capital cost to the capacity of the ethanol biorefinery:

$$\frac{K}{Q} = 1{,}971677 - 0{,}034146 * Q + 0{,}000264 * Q^2 \qquad (1)$$

where K represent capital costs in million of 1998 US dollar, whereas Q is the size of the plant in million of US gallons per year. The equation has been converted to current dollar value and into a capacity of million of litres per year (MLPY). The converted equation is plotted in Figure 38.

In Figure 38 is plotted even the curve for 2009 US dollar, used to help us to make a comparison between data of column (A). Data reported in Table 25 seem to match with the graph. Therefore in this work the equation of [165] is used to take account of economies of scale. Then the 10,1% is subtracted to this value to consider the saving provided by the no-requirements of boiler and milling. Hence, for example for an ethanol plant of 45 MLPY capacity a specific cost of 0,729 $/LY is obtained. Subtracting the 10,1%, the final value of 0,656 $/LY would be used in this case.

Regarding O&Ms, they are already listed without the item of purchasing of gas and electric energy. Therefore, since data are quite similar between references, a simple average is done, and for O&M a final value of 0,0572 $/L is chosen in this work.

TABLE 25. CAPITAL COST AND O&M FOR ETHANOL PLANT REPORTED IN LITERATURE.

ITEM	CAPACITY (MLPY)	$/L (A)	7,3% saving	2,8% saving	SUBTOTAL ($/L)	REFERENCE	escalation index	2013 WORTH
Capital Cost	378,5	0,476	0,035	0,013	0,428	[125]	1,1272	0,482
	200	0,346	0,025	0,010	0,311	[166]	1,2394	0,386
	189	0,594	0,043	0,017	0,534	[125]	1,1272	0,602
	151	0,309	0,023	0,009	0,278	[164]	1,4177	0,394
	50	0,625	0,046	0,018	0,562	[166]	1,2394	0,696
	1	0,750	0,055	0,021	0,674	[166]	1,2394	0,836
O&M					0,0455	[125]	1,1272	0,0513
					0,0436	[166]	1,2394	0,0540
					0,0469	[164]	1,4177	0,0665

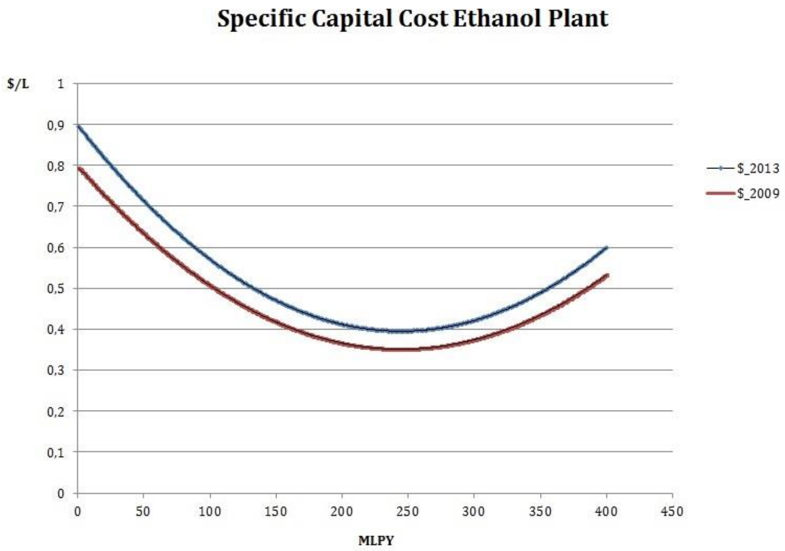


FIGURE 38. SPECIFIC CAPITAL COST IN DOLLAR PER LITRE PRODUCED FOR ETHANOL PLANT, OBTAINED WITH EQUATION OF [227].

5.4.3 Coupling with the NPP

Some modifications that coupling between plants involves are already discussed in the two previous sections: i.e. not considering costs of boilers and purchase of energy.

On the same plane, also for the desalination case the cost of boiler is not considered. Therefore, remembering from paragraph 3.3 that capital costs for desalination plant are assumed to be 1300 \$ per m^3/d installed, we subtracted to this value the 7% of the total investment cost, as is reported by [125].

Then the costs for other modifications like heat exchangers, additional security condenser, etc. are calculated as described in paragraph 4.3.

5.5 THE ALGORITHMS

In this paragraph, the developed algorithms to analyse the profitability of investment with Real Option Approach are inserted.

5.5.1 Option to build

The advantage of the option to build is to proceed with the investment only when the uncertainties are solved in a positive way. In the real life, the investor can wait for a period to see how some uncertainties (i.e. market prices) are evolving.

In this way, the risk associated with the project became lower and the possibility to have a success is greater. Therefore the investor has the possibility to choose whether to build the plant or not. Instead, with the DCF method it is possible to calculate the NPV of the investment in a static way. The investor owns some inputs, like overnight costs, prices, taxes, discount rate, etc. and simply calculates the NPV, without taking into account the possibility that, in case of a not favourable scenario, the investor can decide to do not proceed with the project. This affects the average profitability of the investment.

Then, the option to build gives an extra worth to the investment, considering that for the investor is not mandatory to invest, if the solved uncertainties are negative. To develop an algorithm for the option to build means to capture the eventual profitability for positive scenario, and to reject negative scenarios.

The analysis starts with the development of the classic DCF. Then the uncertainties are introduced and simulated with the Monte Carlo method. Finally NPVs are calculated with the algorithm and the mean is saved. The difference between the previous value of the NPV and the new value represents the worth of the option to build.

The DCF method can be schematised as in Table 26. Assuming that all the investments costs are paid at year 0 (overnight costs), in the balance sheet they are depreciated until year 13, with a depreciation index of 8%. Therefore as yearly cost item, the 8% of the overnight costs enter in the sheet, until the total investment costs are fully considered. Actually, at the last year, not the 8% of the investment is inserted in the balance sheet, but the difference between the total amount already considered and the real overnight cost. Therefore, for example, at year 13 the annual investment cost introduced in the balance sheet is half of the sum considered in the previous years.

TABLE 26. SCHEMA FOR DCF METHOD. ANNUAL CALCULATION.

Investment	Annual Depreciated Investment cost	+
Annual Net Cash Flow	Annual revenues	-
	Annual costs (i.e. O&M)	-
Taxes	Taxes	=
	Annual Free Cash Flow (Future Value)	*
Actualization	Discount Factor	=
	Present Value of Annual Free Cash Flow	

Then the future free cash flows are actualized as follows:

$$PV_t = \frac{FV_t}{(1+r)^t}$$

where r is the discount rate, as discussed in 5.1. In this work, the discount rate used is the *weighted average cost of capital (WACC)*, calculated as described in paragraph 5.1.

Important economic fixed parameters used in this work are:

- Share of debt in financing (W_d)
- Share of equity in financing (W_e)
- Cost of debt (C_d)
- Cost of equity (C_e)
- Tax
- WACC (calculated with previous parameters)
- Depreciation index
- Average Drift Price (D)
- Plant Escalation Cost (E)
- O&M Escalation Cost (M)

- Life time for biorefinery
- Life time for desalination plant

Then, in addition to depreciation index and WACC, also other parameters are considered. Average drift price and O&M escalation costs take into account the fact that the yearly net cash flow is not constant during the life period of the plant. Prices of electricity and products historically are subjected to fluctuation and they usually slightly increase every year. Similarly the annual costs can change, due to inflation or general increasing of prices and needed services (i.e. shipping of raw materials) that in turn affects the direct production costs. Plant escalation cost instead is used only in the calculation of NPVs with real options approach. When all the present values (PVs) are calculated, the NPV of the investment is simply their sum:

$$NPV = \sum_{t=0}^{T} PV_t = \sum_{t=0}^{T} \frac{FV_t}{(1+WACC)^t}$$

Apart from NPV, also the IRR is calculated. The whole general schema for DCF is reported in Table 27. Although the parameters of DCF listed previously are steady, the inputs of inserted variables (e.g. costs and prices) can be subjected to different approaches. For example it is possible to have a DCF that calculates the NPV picking up the "punctual" expected values of a variable, or the entire distribution, or a combination of the two approaches. The differences are underlined in Table 28. Distributions are inserted within the sheets and are simulated with Monte Carlo method, thanks to @Risk tool (of Palisade Decision Tool) for Microsoft Excel. For the option to build, items simulated with Monte Carlo method are the following:

- overnight costs of the plants
- O&M costs
- yearly prices of electricity
- yearly prices of sold products.

The meaning of DCFs forecast and real are explained in the following, whereas the difference between DCF static and MC is that, even if the mean value is the same, DCF static owns just that single value, while the DCF MC is characterized by a symmetric distribution of values.

TABLE 27. SCHEMA DCF METHOD.

Year	0	1	2	…	13	14
Investment cost	X	X*8%	X*8%	X*8%	4%*X	-
Annual Costs	Y	Y*(1+M)^1	Y*(1+M)^2	Y*(1+M)^…	Y*(1+M)^13	Y*(1+M)^14
Annual Revenues	R	R*(1+D)^1	R*(1+D)^2	R*(1+D)^…	R*(1+D)^13	R*(1+D)^14
Net Cash Flow (NCF)		Revenues – Costs	Revenues - Costs	Revenues – Costs	Revenues – Costs	Revenues – Costs
Taxes		For every year, if NCF is greater than X*8%, then paid taxes are calculated as 40% * (NCF – X*8%)				40%*NCF
Free Cash Flow		NCF – taxes	NCF - taxes	NCF - taxes	NCF – taxes	NCF – taxes
Discount factor	$\frac{1}{(1+WACC)^t}$	0,929	0,862	…	0,382	0,355
Present Value		(NCF-taxes)*0,929	(NCF-taxes)*0,862	…	(NCF-taxes)*0,382	(NCF-taxes)*0,355

TABLE 28. DIFFERENCES BETWEEN DCFS CALCULATED FOR OPTION TO BUILD.

NAME	COSTS	PRICES
DCF (static)	Expected value	Expected value
DCF (MC)	Distribution	Distribution
DCF (forecast)	Expected value	Distribution
DCF (real)	Distribution	Simulated "real" values

Distributions for the values are always defined as Pert function, because it is the most common distribution used in literature. The only exception is the distribution for the random component for the simulation of prices. This aspect is clarified in Table 29.

TABLE 29. CHARACTERISTICS FOR DISTRIBUTIONS FOR VARIABLES CONSIDERED.

ITEM	DISTRIBUTION	EXTREME VALUES
Overnight costs biorefinery	Pert function	-10%, +10% of nominal value
O&M costs biorefinery	Pert function	-10%, +10% of nominal value
Overnight costs desalination	Pert function	-30%, +30% of nominal value
O&M costs desalination	Pert function	-30%, +30% of nominal value
First value. Prices electricity, water, biodiesel, ethanol	Pert function	-5%, +5% of nominal value
Drift. Prices electricity, water, biodiesel, ethanol	Pert function	Nominal value -2%, nominal value + 2%
Random. Prices electricity, water, biodiesel, ethanol	Normal function, truncate	Mean = 0 σ = 10% of nominal value Truncate = -10%, +10% of nominal value
Prices glycerol, DDGS	Pert function	-5%, +5% of nominal value

Therefore, regarding costs, the values inserted in the balance sheet are simply extracted from the distributions. For example, if the nominal value of overnight costs for MED-TVC is assumed to be 1300 $\$/(m^3/d)$, the range that characterized this distribution will be between 1300*0,7 = 910 $\$/(m^3/d)$ and 1300*1,3 = 1690 $\$/(m^3/d)$. The range for costs of biorefinery is more narrow because in literature values are rather similar between different references. For a similar reason, instead, the range for costs of desalination is wider.

The situation is slightly more elaborated for the simulation of yearly prices:

- firstly the first value for the price at year 0 is extracted from its distribution. Small uncertainty is given (only -5%, +5%) in this case, because it is assumed that, if a scenario is defined with a particular price, the real value inserted in the model can't differ so much from that value. For example if we consider the scenario of a country where the price of water is 1,2 $\$/m^3$ it is not reasonable to vary the price between 1 and 1,4 $\$/m^3$ (that represent a variation of 17%)
- secondly the average drift between two years is extracted from its distribution. For example if the nominal value of average drift is assumed of 1%, the range will be (-1%, 3%)
- finally a random component is added to confer a Brownian path on the trend of the price.

For NPV calculation with DCF MC method, only the first two points of the above list are taken into account: firstly the 1st value is extracted, subsequently its drift is chosen, as in Figure 40. In this way, i.e. considering a constant increment rate, the evolution of prices is exponential.

Then, since now it is a normal DCF analysis. With this method two values of NPV (one called *static*, one called *MC*) are calculated for year 0. The NPV calculated with DCF MC is the benchmark to compare with the highest NPV found with ROA in order to calculate the option's value.

For ROA, the evolution of prices includes also a random component. In Table 30 an example is reported. This is only one example for the simulation of prices of biodiesel for year 0 and year 1, in one of the scenarios supposed. The extraction of the 1st year value and for the drift are done only once per trend of the yearly price, while the extraction of the random component is done for every year.

FIGURE 39. THE PERT DISTRIBUTION USED TO EXTRACT THE INVESTMENT COSTS FOR THE MED-TVC PLANT.

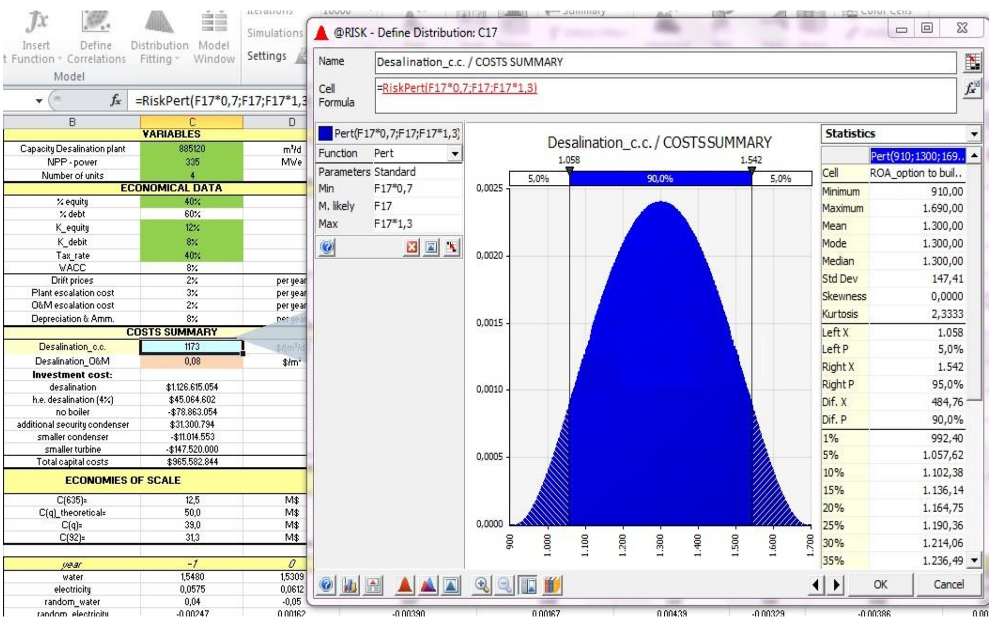


FIGURE 40. HOW THE MC METHOD WORKS FOR SIMULATING PRICES.

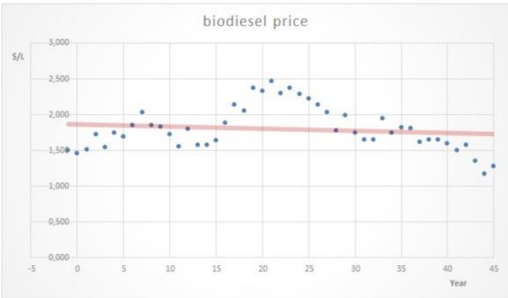
The general results of the trends from year -1 to year 45 can have very different shapes between each other. In Figure 41, Figure 42 and Figure 43 there are some examples for the yearly prices of biodiesel, ethanol and electricity, respectively. Prices are simulated from year -1 to year 45. Indeed, the price for year -1 is used to forecast the profitability of the investment at year 0. Moreover, since the life time of the plant is 35 years and the value of the option is studied for 10 years, the prices are simulated up to the year 45.

To introduce the RAO a trigger is selected. In this work the trigger is the NPV of the DCF forecast. It is also useful to remember that ROA is more worthwhile when big uncertainties surround the project. For example, ROA could be useful if the project involves construction of a totally modern biorefinery, selling new products. In Europe, biofuels own a just 15 years old market and whole European capacity grown exponentially since 2007.

Therefore prices for biofuels are evolving rapidly and are difficult to forecast. In addition biofuels prices strongly depend on the biomass used.

TABLE 30. EXAMPLE OF EVOLUTION OF BIODIESEL PRICE BETWEEN YEAR 0 AND YEAR 1.

Price of biodiesel	Nominal value	1,5 $/L
	Range (-5%; + 5%)	1,425 $/L ; 1,575 $/L
	Extraction of year 0 price	1,483 $/L
Drift of biodiesel	Nominal value	+1%
	Range	-1% ; +3%
	Extraction of drift	+2,03%
Random value	Nominal value	0
	Range	-0,15 ; +0,15
	Generation of random	-0,002
RESULT	Biodiesel price year 1	(1,483 * 1,0203) – 0,002 = 1,5111 $/L



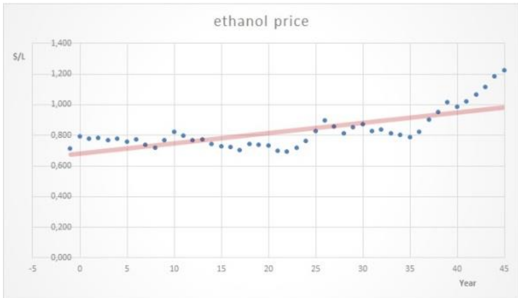


FIGURE 41. YEARLY BIODIESEL PRICE. EXAMPLE ON THE TOP LEFT.

FIGURE 42. YEARLY ETHANOL PRICE. EXAMPLE ON THE TOP RIGHT.

FIGURE 43. YEARLY ELECTRICITY PRICE. EXAMPLE ON THE LEFT.

Algae are not yet used on the real market, and great uncertainty involve this feedstock. Moreover the historical data for biofuels prices could be no longer valid in the future, especially if the European Community, that posed demanding goals in matter of consumption of biofuels in transport' sector, introduces incentive to increase the usage. Similarly electricity prices can be unknown in Arabic countries where a desalination plant could be proposed, or they can vary considerably even in European countries, characterized by a new liberal energetic market (i.e. Italy).

Hence, it is possible to reckon that the investor knows the expected values for overnight costs and O&M. On the other hand, historical trends of prices of biofuels, electricity and desalinated water are not well known. In this situation the investor builds a DCF sheet (that has been called *forecast*), with all the punctual values that he owns: expected values for costs and current wholesale prices for the products. In addition one important simplification is introduced: at 1st of January of every year, the investor can take his decision. At that date, he knows the prices of the previous years and builds the DCF sheet. Actually, if he is at year n, he knows the prices of all the previous (n+1) years (see Table 31).

TABLE 31. EVOLVING KNOWLEDGE OF THE INVESTOR.

YEAR	WHAT INVESTOR KNOWS
0	Prices year -1 (the drift in this case is that one used for DCF static)
1	Prices for years -1, 0. Drift between -1 and 0
2	Prices for years -1, 0, 1. Drift between -1 and 1
3	Prices for years -1, 0, 1, 2. Drift between -1 and 2
4	Prices for years -1, 0, 1, 2, 3. Drift between -1 and 3
...	

If the scenario seems to be profitable and he decides to build, the works start immediately the day after and at the end of the year the plant is ready. With this strong assumption, the following year the project begins to give revenues. This rough assumption eases the analysis, but doesn't affect the study of this work, that has also the aim to evaluate the potential role of Real Options. Therefore, the effect of eventual delays, the interests on debts and the real cash flow profile are not considered.

The result of the DCF forecast (read Table 28 to see how it inserts data) is the NPV: if it is positive the investor decides to build, otherwise he can wait for the 1st of January of the next year and then to build a new DCF, with an updated knowledge.

Simultaneously, another DCF is built (called *real*). As said in Table 28, it works in the same way, but with a different approach: regarding costs, they are extracted from distributions like for DCF MC, whereas regarding prices, they are calculated as explained in Table 30, and for this reason they are called "real" prices. Therefore for DCF MC the drift is constant and the prices evolve as an exponential; for DCF real a random component is continuously extracted and the prices evolve as a Brownian. Also in this case the NPV is calculated. If the correspondent DCF forecast has suggested a positive NPV, then the investment is done and the NPV for the DCF real is saved, otherwise a value of zero is given. In this way, most of negative situations are cut and the mean value of NPV is consequently and reasonably greater.

The DCF forecast and the DCF real are built for 10 years: as seen in the previous table, the investor gains more data with passing of years. Therefore it is frequent that the investor needs more information to take a decision. For example the price of electricity could suggest the not profitability of a project at year 3, but a sudden diversion could happen at year 4. Subsequently the project could become profitable only at that year. Similarly the same project could return not profitable at year 5, because of the drop of the price of water. In the real life, if the investment is done, the production is usually started and the project is carried on (even if sometimes, recognizing mistakes could save a lot of money: option to abandon). However, in the model developed here, a decision taken at year n, doesn't affect the decision at year (n+1). In this way, it is possible to catch how a project is or is not objectively profitable for the eyes of an investor.

To succeed with this aim, we completed both of DCFs for 10 years and the mean value for NPV is calculated with several Monte Carlo simulations. For example, at year 6 if the DCF forecast suggests to invest, the NPV of the DCF real at year 6 is recorded. The last considered year is the 10^{th}. Indeed it is reasonable to think that the model can lose validity in 10 years because the technology could change, the costs of the plant could vary significantly, the cost of financing could be modified by new agreements, etc. In addition, to affirm that an investment can be profitable in 11 or more years has low attraction for stakeholders.

TABLE 32. IN THIS EXAMPLE, THE SHEET IS REPORTED FOR YEAR 4, DESALINATION CASE. THE FORECASTED NPV CALCULATED WITH THE DCF (UP) IS NEGATIVE. CONSEQUENTLY, THE VALUE OF NPV FOR THE REAL CASE (DOWN) IS NOT CONSIDERED. DATA IN M$.

DCF forecast – year 4														
Year		*0*	*1*	*2*	*4*	*5*	*6*	*7*	*8*	*...*	*27*	*28*	*29*	*30*
Investment cost	Capital cost				-1113									
	Annual capital cost					-89	-89	-89	-89		0	0	0	0
Annual net cash flow						70,7	70,6	70,5	70,4		64,5	63,9	93,4	0
Taxes						0	0	0	0		25,8	25,6	25,4	0
Annual free cash flow						70,7	70,6	70,5	70,4		38,7	38,4	38,0	0
PV of annual cash flow					827,9	48,9	45,3	42	38,9		5,25	4,83	4,44	0
NPV		-324 M$												
IRR		4%												
PI		0,45												
DCF real – year 4														
Year		*0*	*1*	*2*	*4*	*5*	*6*	*7*	*8*	*...*	*27*	*28*	*29*	*30*
Investment cost	Capital cost				-1210									
	Annual capital cost					-96,8	-96,8	-96,8	-96,8		0	0	0	0
Annual net cash flow						75,6	76,3	67,8	52,1		28,8	23,5	32,9	0
Taxes						0	0	0	0		11,5	9,4	13,2	0
Annual free cash flow						75,6	76,3	67,8	52,1		17,3	14,1	19,8	0
PV of annual cash flow						52,2	48,9	40,4	28,8		2,34	1,78	2,31	0
NPV		-476 M$												
IRR		0%												
PI		0,35												

5.5.2 *Option to switch*

The option to switch is applicable in all the situations in which there is a margin of operability. If the operations in a plant are fixed, maybe even variable, but previously scheduled, there is no possibility to take advantage from the switching.

A common example in literature of the option to switch is the possibility to burn a different fuel within a boiler (i.e. oil or natural gas), accordingly to the price market in that moment.

In this study, the option to switch is developed only for the cogeneration with the MED-TVC. As said in paragraph 4.2, the nuclear site coupled with the desalination plant studied here is characterized by two production ways: one is mainly daily, producing almost entirely electricity; the other one is mainly nocturne, producing more water. Therefore, in our study the flexibility is given by the possibility to produce either water or electricity in case of cogeneration with a desalination plant. Actually, the only fact to work in load following mode (therefore variable mode) doesn't give a switching option. This is because a load following could be static: for example at 2 a.m. the plant switches to a different production level, producing more water, and at 6 a.m. the plant comes back to the daily production way. In this operation mode there is not choice to take advantage of evolving prices in the market, because the production ways are pre-scheduled.

The scenario is totally different if the flexibility to choose is given to the manager of the plant: accordingly to the live prices of water and E.E., the plant can switch to the more profitable production way, increasing the revenues. With this approach, every day can be characterized by a different level of production. However, it is reasonable to think that, because of the usual daily trend of electricity price, the production oscillations will be not very different from the load following mode. The main difference is that the manager can choose the perfect time to switch into the other production way. In addition during the weekend, when E.E. price is lower, it is possible to increase the production of water.

With this option, we studied the advantages of a flexible production mode, in comparison to static ones. Therefore, in this case costs (both overnight and O&M) are not taken into account. To compare operation configurations the EBIT is calculated for 1 year. For the option to switch only the trend of electricity's price within the week is simulated. Indeed the price of water is assumed to don't vary so much within the year and is kept constant. Then the possible revenues are calculated for both the operation modes, with the prices of water and electric energy in a particular period of the day. It is assumed that the manager would take advantage of the situation, switching to the more profitable operation mode.

To simulate the daily price of E.E., the daily time is divided in 48 intervals, once every half an hour. For every time interval a drift and a random component are extracted by their distributions. The random component is kept constant for all the intervals. Instead the range for the extraction is continuously changed between different intervals. The reason is that, statistically, the price evolves with a simile path during the day: at night the electric energy is cheaper, the minimum is usually reached at 4 a.m., at 6 a.m. the price begins to rise until noon, then there is a smooth decrease, that precedes a new positive ramp. At 8 p.m., the maximum price of the day is usually reached. After that, a continuous rapid decrease starts again.

The random component is extracted from a Normal distribution, with mean, sigma and extreme values equal to 0, 0,015 and (-0,001; 0,001) respectively. The graph of this distribution is reported in Figure 44. Instead to simulate the daily prices, trend were observed for the electricity prices in UK, using the website www.bmreports.com [167]. Consequently, the drifts are chosen to reproduce the same trend. They are not fixed but extracted from Pert distributions, characterized by values reported in Table 33.

The real and simulated daily trends are drawn in Figure 45 and Figure 46, respectively. In particular in Figure 45 the trend is reported for two consecutive days, separated by the dashed line. During the week, the following day begins with a slightly increased value in comparison to the price at the same hour of the day before. The most important consideration is that during the week days, the shape is rather constant, with the picks that are observed in the same time intervals. The situation is different during the weekend. Following this approach, dissimilar drifts values are given for Saturday and Sunday.

Hence, in this study firstly the daily price trend is simulated, secondly they are reproduced for all the days of the week. The weekly trend of the E.E. price instead is reported in Figure 47. Extracting different drifts for every time interval and adding a random component, produces different prices history for every week of the year. The main anomaly that could be observed between the reality and our simulations is that real prices own more drastic picks. These drastic picks are not interesting for the aim of this study. Indeed, it's less reasonable to assume that a plant could switch the production mode several times per day, following every single pick. Instead is definitely more interesting observing the general trend.

FIGURE 44. NORMAL TRUNCATED DISTRIBUTION FOR THE EXTRACTION OF THE RANDOM COMPONENT IN CASE OF SIMULATION OF DAILY ELECTRICITY PRICE.

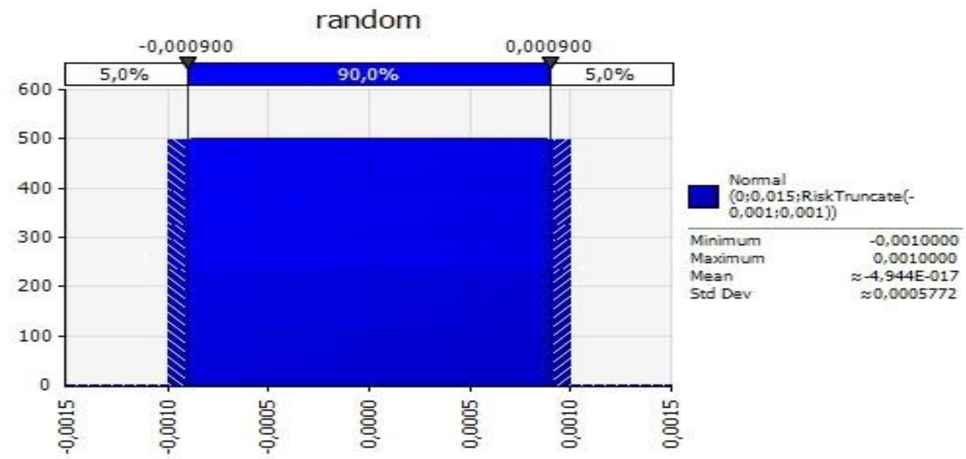


TABLE 33. DRIFT FOR E.E. PRICES PER HALF AN HOUR, FOR THE WEE DAYS ALL DRIFTS ARE EXTRACTED FROM PERT FUNCTION.

TIME	min	mode	max
0.00	- 5%	- 4%	- 2%
0.30	- 6%	- 5%	- 2%
1.00	- 7%	- 6%	- 3%
1.30	- 7%	- 6%	- 3%
2.00	- 7%	- 4%	- 3%
2.30	- 7%	- 4%	- 3%
3.00	- 5%	- 3%	- 1%
3.30	- 1%	0%	+ 3%
4.00	- 1%	+ 2%	+ 5%
4.30	- 1%	+ 2%	+ 5%
5.00	+ 5%	+ 8%	+ 9%
5.30	+ 5%	+ 8%	+ 9%
6.00	+ 5%	+ 8%	+ 9%
6.30	+ 5%	+ 8%	+ 9%
7.00	- 1%	+ 6%	+ 8%
7.30	- 1%	+ 6%	+ 8%
8.00	- 1%	+ 6%	+ 8%
8.30	- 1%	+ 6%	+ 8%
9.00	- 1%	+ 6%	+ 7%
9.30	- 1%	+ 6%	+ 7%
10.00	- 2%	+ 1%	+ 2%
10.30	- 2%	+ 1%	+ 2%
11.00	- 2%	+ 1%	+ 2%
11.30	- 2%	+ 1%	+ 2%
12.00	- 2%	+ 1%	+ 2%
12.30	- 5%	- 3%	+ 1%
13.00	- 5%	- 3%	+ 1%
13.30	- 1%	- 0,5%	+ 0,5%
14.00	- 1%	- 0,5%	+ 0,5%
14.30	- 1%	- 0,5%	+ 0,5%
15.00	- 1%	- 0,5%	+ 0,5%
15.30	- 1%	- 0,5%	+ 0,5%
16.00	- 1%	- 0,5%	+ 0,5%
16.30	- 1%	+ 7%	+ 8%
17.00	- 1%	+ 1%	+ 5%
17.30	- 1%	+ 3%	+ 5%
18.00	- 1%	+ 3%	+ 5%
18.30	- 1%	+ 3%	+ 5%
19.00	- 2%	+ 1%	+ 2%
19.30	- 2%	+ 1%	+ 2%
20.00	- 2%	+ 1%	+ 2%
20.30	- 4%	- 3%	+ 1%
21.00	- 9%	- 8%	+ 1%
21.30	- 9%	- 8%	- 5%
22.00	- 9%	- 8%	- 5%
22.30	- 9%	- 7%	- 1%
23.00	- 9%	- 7%	- 1%
23.30	- 9%	- 7%	- 1%

For the week end, the drifts are built with the same logic, but the values are dissimilar, to better follow the trends of the real prices for the weekend.

FIGURE 45. SYSTEM E.E. SELL PRICE IN UK. DATA REFERRING TO DATES 21ST AND 22ND OF AUGUST 2013. DATA ARE REPORTED EVERY HALF AN HOUR AND FOR THIS REASON THE X-AXIS OF THE DAY IS DIVIDED IN 48 INTERVALS.

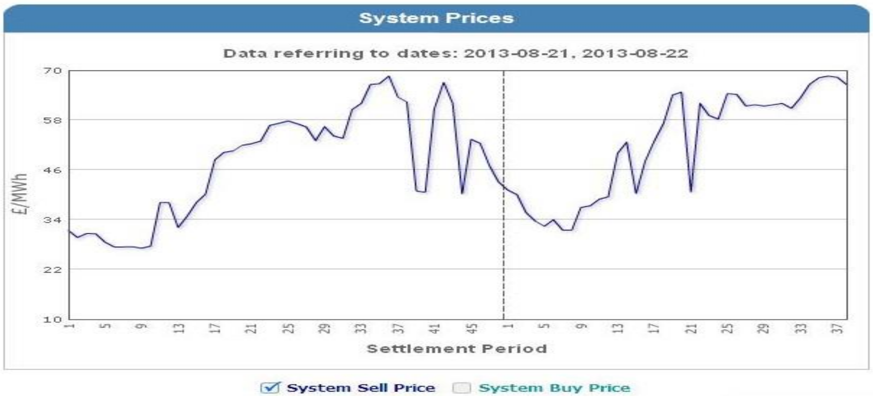
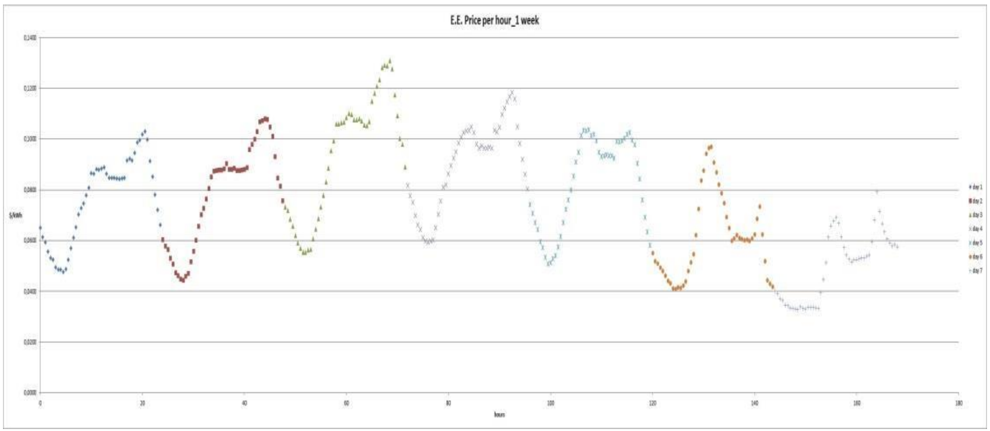


FIGURE 46. SIMULATED ELECTRICITY PRICE FOR ONE DAY.

FIGURE 47. EXAMPLE OF SIMULATED WEEKLY PRICES FOR ELECTRIC ENERGY.



Therefore this is the followed path to simulate the daily and weekly prices. Then, when prices are know, they are multiplied to electricity and water produced to calculate the revenues. To confront the profitability, only variable revenues are compared. Indeed during the day there is always a minimum constant level of production both for the water and the electric energy.

We analyse in the following the option to switch for a nuclear site formed by 4 IRIS reactor. With the plant layout of Figure 7, 2 IRIS reactors (say number 1 and 3) always produce only electricity at nominal power level. Other 2 IRIS reactors instead are connected to the MED-TVC units and they switch the production.

Hence, the products of nuclear reactors are as in Table 34:

TABLE 34. ELECTRICITY AND WATER PRODUCED PER HOUR BY 4 IRIS REACTOR IN BY-DESIGN AND OFF-DESIGN WORKING MODE.

NUCLEAR SITE. PRODUCTION BY-DESIGN		
	1st and 3rd IRIS	2nd and 4th IRIS
Power per unit (electric)	335 MWe	335 MWe
Power per unit (thermal)	1000 MWt	1000 MWt
Number of units	2	2
Total thermal power	1000 MWt	1000 MWt
Constant thermal power MED-TVC	0	231 MWt
Thermal power to turbine	1000 MWt	770 MWe
Electric power	335 MWe	258 MWe
Constant electric power MED-TVC	0	10 MWe
Electric power to the grid	335MWe	248 MWe
E.E. produced per hour	*335 MWh*	*248 MWh*
Water produced per hour	*0 m^3*	*4610 m^3*
NUCLEAR SITE. PRODUCTION OFF-DESIGN		
	1st and 3rd IRIS	2nd and 4th IRIS
Power per unit (electric)	335 MWe	335 MWe
Power per unit (thermal)	1000 MWt	1000 MWt
Number of units	2	2
Total thermal power	1000 MWt	1000 MWt
Constant thermal power MED-TVC	0	922 MWt
Thermal power to turbine	1000 MWt	0
Electric power	335 MWe	0
Constant electric power MED-TVC	41 MWe	41 MWe
Electric power to the grid	294MWe	- 41 MWe
E.E. produced per hour	*294 MWh*	*0 MWh*
Water produced per hour	*0 m^3*	*18440 m^3*

In this table is possible to see how, even if reactors number 1 and 2 produce constantly electric energy, when the site operates in off-design, they have to supply the electricity for the desalination plant. Indeed, in off-design mode reactors 2 and 4 direct all their steam to the MED-TVC plant. Only the minimum level (7,8%) of the steam is directed to the turbine, but only to prevent failures, without producing electricity.

Therefore all reactors have a minimum output constantly produced: it is 294 MWh for reactors 1 and 3, while it is 4610 m^3 of water for reactors 2 and 4[18].

[18] Data refer to production per hour.

Then the stationary and variable products are listed in Table 35.

TABLE 35. STATIONARY AND VARIABLE PRODUCTS PER HOUR, FOR EVERY NUCLEAR REACTOR.

	STATIONARY		**VARIABLE. BY-DESIGN**		**VARIABLE OFF-DESIGN**	
	E.E.	water	E.E.	water	E.E.	water
IRIS 1	294 MWh	0	41 MWh	0	0	0
IRIS 2	0	4610 m^3	248 MWh	0	0	13830 m^3
IRIS 3	294 MWh	0	41 MWh	0	0	0
IRIS 4	0	4610 m^3	248 MWh	0	0	13830 m^3
TOTAL	**588 MWh**	**9220 m^3**	**577 MWh**	**0**	**0**	**27660 m^3**

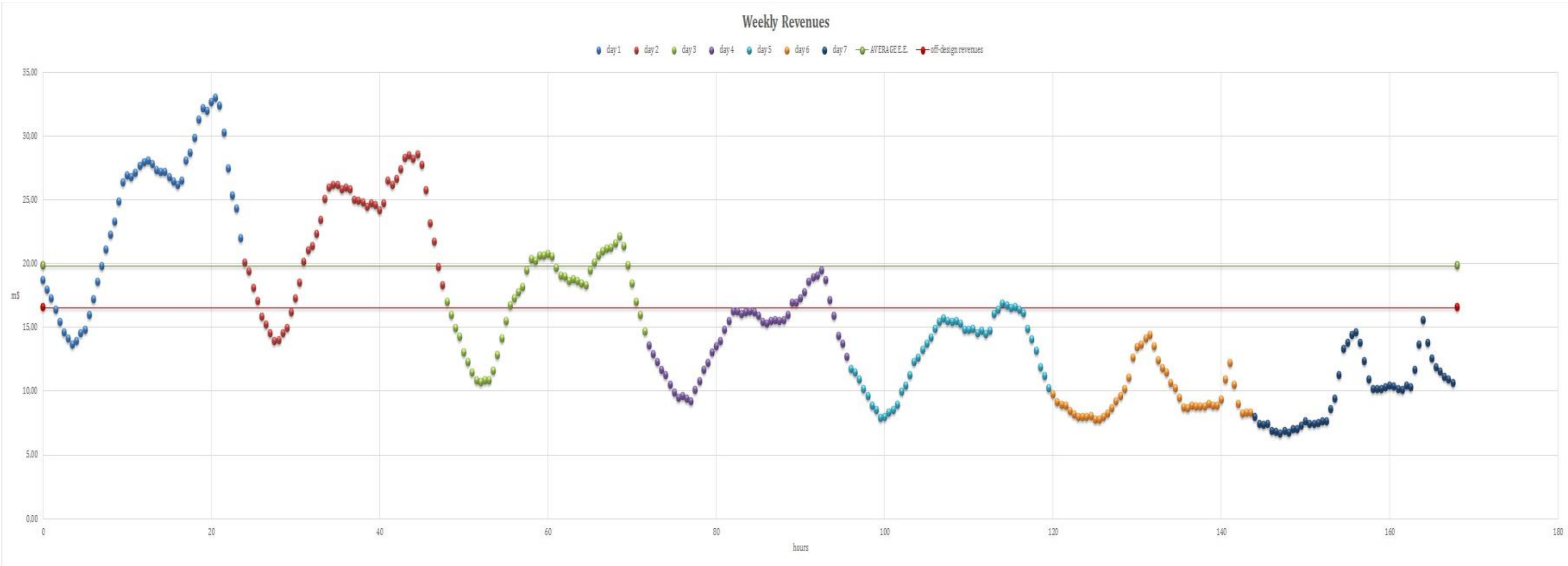
Knowing prices and yields, for every half-an-hour-interval revenues are calculated for both the by-design and off-design arrangements and then they are compared. It is assumed that if the revenues from the production of electricity are higher than the revenues from production of water, the plant switches to the by-design operation mode. Also the vice-versa is true.

Simultaneously, also a counting system is inserted to count how many times the production is in by-design and in off-design: if for example the revenues for selling water are higher, the counting system adds a 1 to the respective cell. With the counting system it is possible to calculate how many hours the plant worked in by-design and off-design mode at the end of the day and, also, at the end of the week. Knowing this data or knowing the total water and E.E. produced it is possible to forecast what is the capacity effectively used for both the plants. Percentages obtained are subsequently double checked with graph in Figure 36.

In the next 2 figures there is a comparison between revenues for by-design and off-design modes. In Figure 48 the comparison is for the day 1 of the week; in Figure 49 all the week is shown. Graphs refer to the same example. The green line (average revenues by selling electricity) is used to comment the results in the next chapter. With the trend as in Figure 48, it is reasonable to switch to off-design only between 2 a.m. and 6 a.m.; but as underlined in Figure 49, this trend can change significantly during the week.

FIGURE 48. DAILY REVENUES FOR THE FIRST DAY OF THE WEEK. IN THIS EXAMPLE THE REVENUES FOR THE OFF-DESIGN CONFIGURATION ARE CALCULATED WITH A WATER PRICE OF 1,2 $/M^3.

FIGURE 49. WEEKLY VARIABLE REVENUES FOR BY-DESIGN MODE (DOTS) AND OFF DESIGN MODE (RED LINE). IN THIS EXAMPLE THE REVENUES FOR THE OFF-DESIGN PRODUCTION ARE CALCULATED WITH A WATER PRICE OF 1,2 $/M^3.



In this figure the red line represents the variable revenues by selling water, whereas the dots are the revenues per half an hour by selling electricity. The green line is the level of the average revenue by selling E.E.. It is calculated as max value + min value, divided per 2. In this case the price of electricity in the week evolves downward. In this example the potential of the option to switch is shown. In the first 3 days of the week is better to produce electricity, while at the end of the week is better to produce water, because the respective revenues are higher. Having a fixed load following scheduled mode (e.g. switching to off-design mode only between 2a.m. and 6 a.m.) can't capture the profitability according to the evolution of prices.

Then the yearly revenues for the entire time life of the plant are calculated. They are simply found as the product of the weekly revenues, multiplied for the number of the weeks of the year (52,14) and for the availability of the system (90%). This is done for different operation mode:

- Load following *static*: 2 a.m. – 6 a.m. off-design, 6 a.m. – 2 a.m. by design
- Load following *flexible*: variable
- By-design: the plant always work in by-design arrangement
- Off-design: the plant always work in off-design arrangement

To compare the profitability of the project, the investment costs are not considered, as well as the O&Ms and the eventual drifts of prices and costs between years. Indeed, we want to evaluate the advantage to switch with maximum flexibility, in comparison with different static operation modes. Drifts and costs are quantities that are in common between modes. Moreover introducing distributions of costs and drifts could affect the real simple difference between revenues that comes from only the operation modes.

The revenues are instead simply actualized with the WACC as discount rate. The final confront is done between the sum of all the actualized revenues during the whole life of the plant (25 years for MED-TVC).

In this work therefore the prices are modelled only daily and weekly. On the contrary it is implicitly assumed that they don't vary significantly between months. In the real life the prices are usually slightly lower during the summer. However the fluctuation is less severe in comparison with other fluctuations simulated (especially the daily one).

In this case, plant and O&M costs are not considered. For the option to switch, we assume that the MED-TVC is already built, together with the NPP. Even the difference between O&M costs is considered negligible in comparison with the differences of revenues. Hence in this case, the object used to do the confront and to evaluate the worth of the option is the EBIT (Earnings Before Interests and Taxes).

CHAPTER 6.
RESULTS AND DISCUSSIONS

In the sixth chapter, the final results of the economic analysis are reported. At the beginning the chapter lists the common parameters involved for all the options have been studied here. The results are presented gathered per option. At first the two options to build, then the option to switch. For all the options, before the scenarios evaluated are introduced and then the graphs that summarize the results. Subsequently, a discussion is done to critically analyse what we obtained. At the end of the chapter a sensitivity analysis concludes this work.

6.1 COMMON PARAMETERS

For the economic analysis, the values for the common used parameters are listed in Table 36.

The depreciation index and the life time for the plants are found in literature [125], [161]. The parameters involved for the calculation of WACC are rather high. Indeed, with this type of investments, the associated risk is usually high. For this motivation it is reasonable to assume that banks and stakeholders would require a higher cost of capital. Therefore the 8% value of WACC is conservative.

The other values (drifts and escalation costs) are reasonable hypothesis.

TABLE 36. COMMON ECONOMIC PARAMETERS, USED IN THIS WORK.

Depreciation index	8%
W_d	60%
W_e	1- W_d = 40%
C_d	8%
Tax	40%
C_e	12%
WACC	8%
Average Drift Price (D)	2% per year
Plant Escalation Cost (E)	3% per year
O&M Escalation Cost (M)	2% per year
Life time for biorefinery	35 years
Life time for desalination plant	25 years

6.2 OPTIONS' VALUES

6.2.1 Option to build: Biorefinery

SCENARIOS' DEFINITION

The developed model is very flexible and has the aim to be as much as general possible. In particular, the aim is not to study the applicability of a cogenerative plant in a selected country. The goal of this work is to study the eventual profitability of such a plant, in the real uncertain market. Therefore, defined scenarios are not limited to precise and very specific parameters. They want to provide some outputs, in relation to the variations of some peculiar inputs. Therefore, thanks to a static DCF analysis, it is possible to affirm that definitely the most important products (in matter of revenues) and costs are the biodiesel and the not-sold electricity.

This fact also comes out looking at Figure 50 and Figure 51. Graphs are drawn for the case of scenario 2 (see next table). This scenario is characterized by standard prices: 1,5 $/L for biodiesel, 0,7 $/L for ethanol and 0,08 $/kWh for electricity. In addition prices for glycerol (80% pure grade) and DDGS are set to 350 $/ton and 220 $/ton respectively [168], [169].

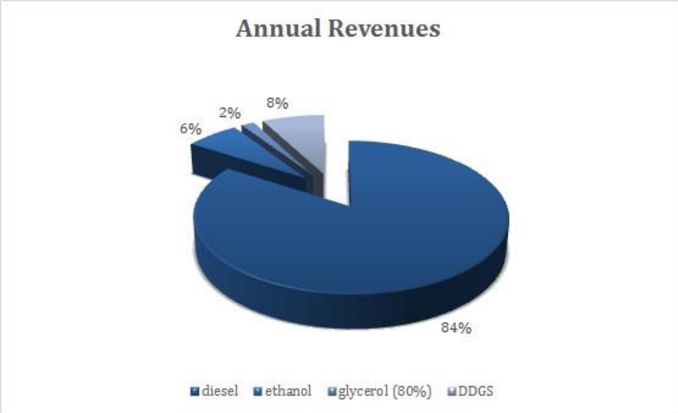


FIGURE 50. SHARE OF THE ANNUAL REVENUES FOR THE BIOREFINERY.

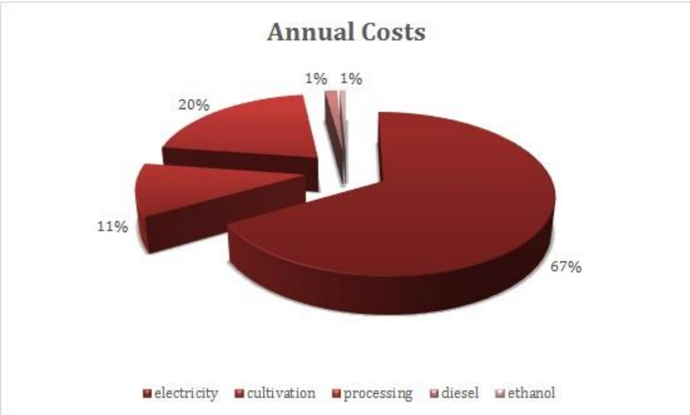


FIGURE 51. SHARE OF THE ANNUAL COSTS FOR THE BIOREFINERY.

Therefore, even if there is the possibility to study hundreds of scenarios, modifying prices, costs or general parameters (e.g. taxes in a particular country or different financial share between debt and equity), in this work scenarios are elaborated changing the biodiesel and the electricity prices. The sensitivity upon other parameters are studied with a sensitivity analysis. The 7 scenarios developed are reported in Table 37. The 3 values in the column are, from left to right, the minimum, the average and the maximum value inserted for defining the Pert distribution. They refer to the value of the product at year 0.

TABLE 37. LIST OF SCENARIOS STUDIED FOR THE BIOREFINERY.

SCENARIO	BIODIESEL [$/L]	ELECTRICITY [$/kWh]
1 – cheap biodiesel	1,235; 1,3; 1,365	0,076; 0,08; 0,084
2 – standard biodiesel	1,425; 1,5; 1,575	0,076; 0,08; 0,084
3 – breakeven price	1,52; 1,6; 1,68	0,076; 0,08; 0,084
4 – expensive biodiesel	1,615; 1,7; 1,785	0,076; 0,08; 0,084
5 – incentive biodiesel	1,71; 1,8; 1,89	0,076; 0,08; 0,084
6 – cheap electricity	1,425; 1,5; 1,575	0,067; 0,07; 0,074
7 – expensive electricity	1,425; 1,5; 1,575	0,086; 0,09; 0,095

As it is possible to note, scenarios change only one value per time, to investigate only one effect per time. For all scenarios, the ethanol price distribution is a Pert function with average value of 0,7 $/L (average wholesale price in USA) [169]. According to Table 29, the minimum and the maximum value are the (-5%; +5%) of the average price. Also glycerol and DDGS average prices are that said at the beginning of this section.

To propose an example, the wholesale prices for biodiesel are 1,3 $/L in USA and 1,3-2 $/L in Europe.

Scenarios 1, 2, 4, 6 and 7 don't require any interesting explanation. They simply consider possible different prices for biodiesel and electricity. Scenario 3 has been defined "break-even" because the price of 1,6 $/L is very close to the break-even point of the price of biodiesel with a price of electricity of 0,08 $/kWh. Finally scenario 5 has been defined "incentive" in order to take into account the possibility to have a high price of biodiesel even in countries where it is usually cheaper, thanks to the integration of state incentive to promote this biofuel. This is assumed to be a probable scenario, because of the demanding target fixed by [61].

EXEMPLIFICATIVE RESULT

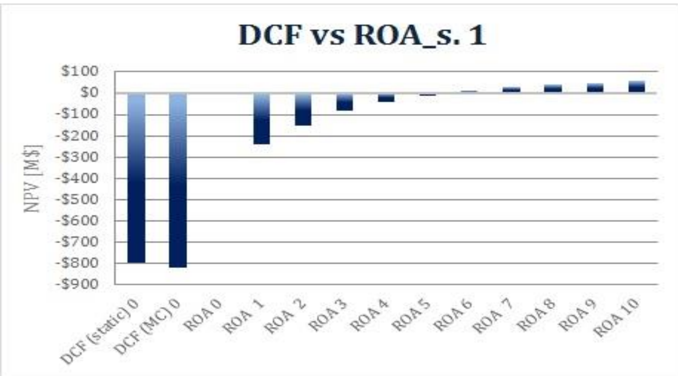
For all scenarios the following relations are found:

- Trend of NPV in function of the DCF and the ROA methods. Trend expressed even in function of the year of the ROA
- Trend of IRR, characterized as above
- Probabilities to don't invest, have a lucky investment or have an unlucky investment, with passing of years.

All the graphs are collected in Appendix E. Here only the most exemplificative results are reported and discussed.

As will be better explicated in paragraph 6.3, the ROA with the option to build shows its potential in broadly negative scenarios. Therefore for the option to build in case of Biorefinery the scenario 1 is discussed (cheap biodiesel. Very negative scenario). Instead for the option to build in case of desalination the scenario 7 (slightly negative, big uncertainty) is carefully analysed.

FIGURE 52. NPVS CALCULATED WITH DCF METHODS AND REAL OPTIONS APPROACH AT DIFFERENT YEARS. RESULTS FOR SCENARIO 1 OF BIOREFINERY CASE.



In scenario 1 (low biodiesel price, fixed at 1,3 $/L), the traditional DCF analysis shows that the investment is definitely not profitable (see Figure 52). Indeed, both the static NPV and the Monte Carlo NPV are strongly negative (about -800 M$). Consequently, at the same year 0, even with the possibility to observe some change in the price at the previous year, it is not possible to have a positive profitability of the investment, because the prices still remain unattractive. For this reason the NPV calculated with ROA at year 0 is equal to zero: it means that the investor never invests. Even if the investor can't capture any profit, the value of the option in this case is already approximately 820 M$, that is the difference between the NPV calculated with DCF MC and that one with ROA at year 0. The manager has the possibility to don't invest and to save 820 M$.

After year 0, it is possible to see a well-defined curve: from year 1 up year 4, the NPV is still negative, at year 5 and 6 it is almost equal to 0, then it starts to rise very smoothly. This is explained as following:

- Since year 1, thanks to the DCF *forecast* the investor begins to see some profitable evolution. Consequently the 43,3% of times he decides to proceed with the investment (percentage found as one's complement of blue line in Figure 54)
- Because of the prevision is based only on two values and one drift, at year 1 the DCF forecast can't capture very well the profitable scenarios. For this reason the average NPV calculated with ROA at year 1 is still rather negative (-240 M$). However it is possible to note a drastic improvement in respect to the case of DCF MC
- The following years, the investor gathers even more data, and the prediction becomes more precise. For this reason the NPV improves
- However the NPV seems to evolve to a just slightly positive value: indeed whereas between year 1 and year 5 the improvement for the NPV is of about 230 M$, in the subsequent 5 years (from 6 to 10), it improves of only about 45 M$
- These considerations are also confirmed looking at Figure 53. The IRR obtained with DCF methods is very low. At year 0, with ROA it is equal to 7,7% (that is the WACC): indeed, all the times that the investor doesn't risk with this project, the model gives a value of 0 to the NPV and a value equal to the discount rate (WACC in this case) to the IRR. Therefore even with the graph of IRR it is confirmed that the investor never invests at year 0
- After year 0, thanks to the possibility to forecast better the evolution of prices, the IRR improves, remaining almost constant
- The most interesting consideration concerning Figure 54 is that the trend of the statistics of the investment change direction, with passing of years: with DCF is confirmed that the investment is strongly negative (red dot on the top) and consequently the ROA suggests to never invests (blue dot on 100%). Later the possibility to have a lucky investment (green line) continues to rise and at year 5 it overtakes the red line of unlucky investments
- The probability to have negative investment is halved in ten years, from 27% to 14%. In addition also the probability to do not invest rise from 56% to 66%

The last considerations explain how it is possible to have a final improvements of the mean value of the simulated NPVs.

Finally, what it is interesting to affirm for this scenario is that the scenario itself still remains not very attractive, because the probability to have a positive investment is only 18-20%. This is because the beginning price of biodiesel is too low to suggest a good profit. But, thanks to ROA, after about 5 years of collecting data, it is possible to complete a precise prediction of the future. Then, although the DCF suggests to completely abandon the project, the ROA suggest to wait the next 4-5 years because the project could turn in a very profitable investment. Indeed, whereas at the beginning (says year 1) the probability to have a positive NPV is lower than having a negative NPV, after 5 years the probability is the same and at the end (year 10) the situation is overturned.

Then, even if it is more probable to don't invest, and consequently the average NPV is approximately 0, the ROA permits to capture large earnings: when the investor understand a positive evolution of prices, he can take advantage of the situation..

FIGURE 53. IRR CALCULATED WITH DCF METHODS AND REAL OPTIONS APPROACH AT DIFFERENT YEARS. RESULTS FOR SCENARIO 1 OF BIOREFINERY CASE.

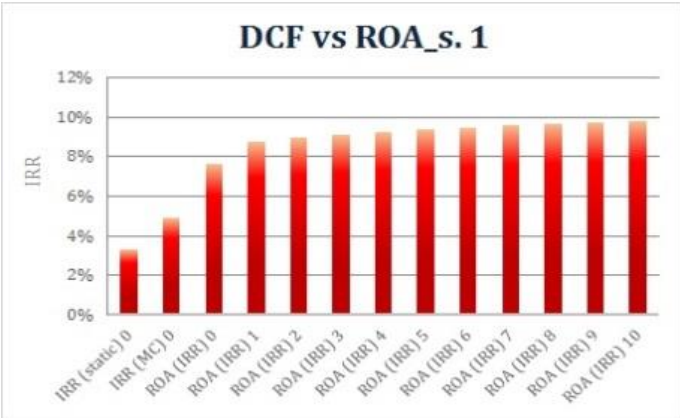
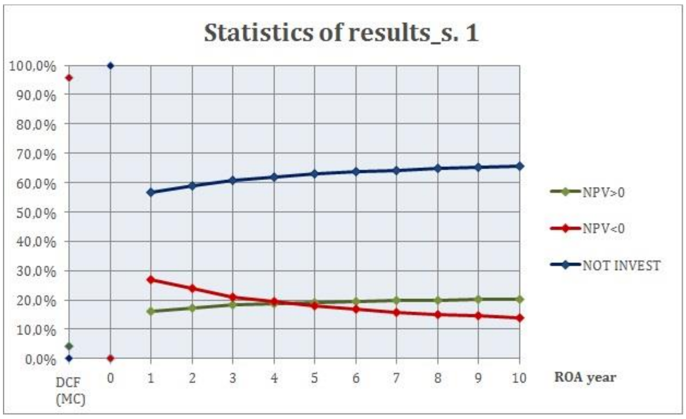


FIGURE 54. PROBABILITY TO DON'T INVEST, HAVING A FINAL POSITIVE NPV AND HAVING A FINAL NEGATIVE NPV. RESULTS OBTAINED FOR SCENARIO 1 OF BIOREFINERY CASE.



6.2.2 *Option to build: Desalination*

The option to build for the desalination studies the suitability to build a MED-TVC plant to couple with the NPP. The site in this case would work in a static load following mode: at 2 a.m., 2 SMRs are disconnected totally to the electric grid and they direct their steam (except for the part that is needed to warm the turbines) to the MED-TVC. At 6 a.m., the site come back to the daily production arrangement.

SCENARIOS' DEFINITION

Even in case of cogeneration with a desalination plant, the scenarios are studied regarding possible prices of involved products. The 7 defined scenarios to study the option to build for MED-TVC plant are listed in Table 38. The drifts are extracted from the same Pert distribution as in the case of the Biorefinery.

TABLE 38. LIST OF DEFINED SCENARIOS TO EVALUATE THE OPTION TO BUILD FOR DESALINATION.

SCENARIO	WATER [$/m^3]	ELECTRICITY. Night Window [$/kWh]
1 - cheap water	1,14; 1,2; 1,26	0,038; 0,040; 0,042
2 - expensive water	2,38; 2,5; 2,63	0,038; 0,040; 0,042
3 - standard case	1,52; 1,6; 1,68	0,038; 0,040; 0,042
4 - pure load following	1,52; 1,6; 1,68	0
5 - cheap electricity	1,52; 1,6; 1,68	0,019; 0,02; 0,021
6 - night price	1,52; 1,6; 1,68	0,057; 0,06; 0,063
7 - breakeven case	1,43; 1,5; 1,58	0,029; 0,030; 0,032

Scenario 3 is defined by standard prices: they are 1,6 $/m^3 for water and 0,04 $/kWh for electricity. The reader could think that the price of electricity is very low and therefore optimistic. Indeed in the option to build the electricity represent an item of cost, because this is the price of the not sold electricity, in order to produce water. Actually this price is not the average of the daily or yearly price in one country (that in the previous scenario has been fixed at 0,08 $/kWh). This price is the worth of E.E. in the nocturne time window. Moreover this is not the proper normal market price: usually the price of electric energy at night is about 0,06 $/kWh. More precisely this is the price at which the E.E. produced could be sold, during the night. Since the possibility to work in load following mode has been debated and it seems to be applicable from the technical point of view, this E.E. would be the electricity "in surplus". Hence, although the normal nocturne market price could be 0,06 $/kWh, we assume that the saturation of market (that entails the requirement of load following) lowers the demand. So, the lower the demand, the lower the prices of saleable E.E. produced in excess.

Then scenarios 1 and 2 change the prices of water, keeping fixed the standard price of electricity.

Instead scenarios 4, 5 and 6 in turn change the price of E.E. keeping fixed the price of water. In particular scenario 4 hypothesizes that the value of E.E. in the night window is zero. This is the case when the control rods are inserted, in order to absorb neutrons to follow the electric demand during the night. For example this is what happens in France. Scenario 5 considers a case in which the market suggests a very cheap sell of E.E. On the contrary scenario 6 considers a case in which the market is not so saturated and therefore the sale of E.E follows the normal price (0,06 $/kWh).

Finally scenario 7 is defined to study what happens in a uncertain scenario: indeed priced of water and E.E. are close to the breakeven point, found with the DCF method (see Figure 67).

EXEMPLIFICATIVE RESULT

Scenario 7 confirms very well the theory of real options approach (NPV drawn in Figure 55). The prices of scenario permit to find a slightly negative NPV with the classic DCF methods. Using the developed algorithm for the option to build, instead it is possible to cut the negative evolutions and to take advantage of the positive ones. At year 0, the prevision is not done accurately and the value is still negative. Instead after about 5 years the asymptotic value of the option is joined. Moreover, the reason of the asymptotic trend comes from the fact that the prevision is already done accurately (the green line of Figure 56 remains constant at about 27%), but the passing of time creates a depreciation of revenues. In other words, the depreciation of revenues is not balanced by the gain of new information to make a more precise prevision.

FIGURE 55. NPVS CALCULATED WITH DCF METHODS AND REAL OPTIONS APPROACH AT DIFFERENT YEARS. RESULTS FOR SCENARIO 7 OF DESALINATION CASE.

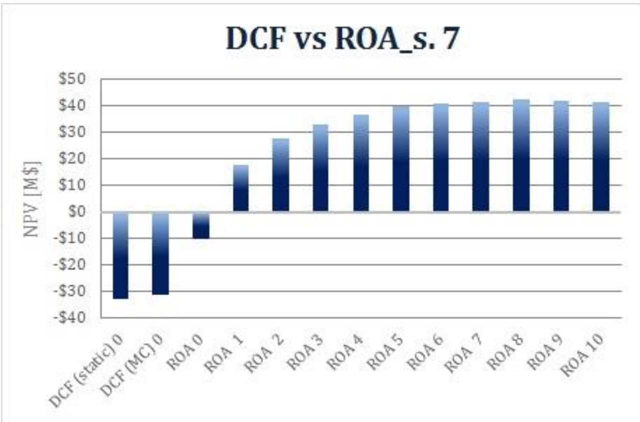
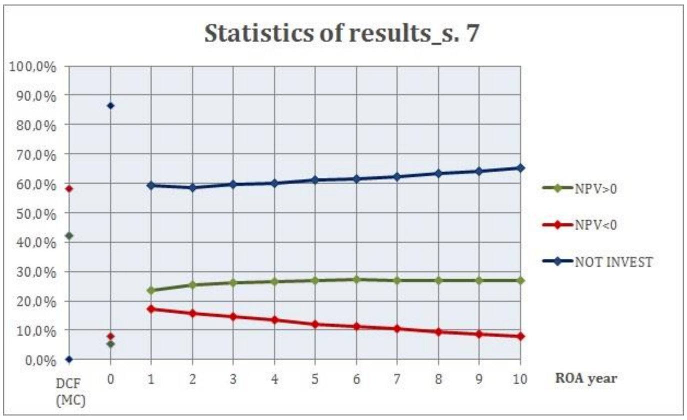
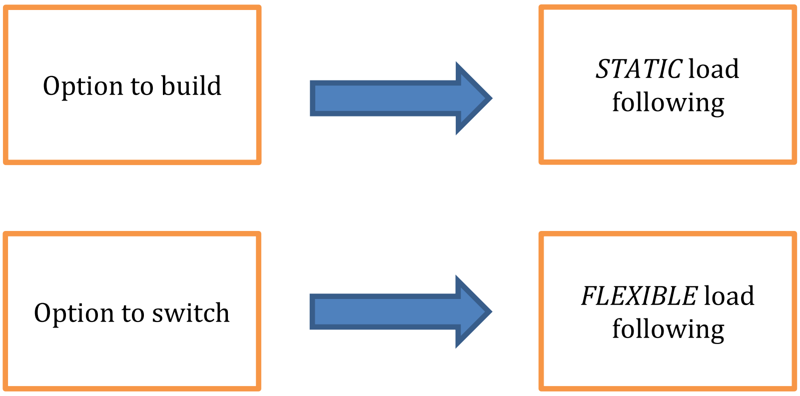


FIGURE 56. PROBABILITY TO DON'T INVEST, HAVING A FINAL POSITIVE NPV AND HAVING A FINAL NEGATIVE NPV. RESULTS OBTAINED FOR SCENARIO 1 OF DESALINATION CASE.



6.2.3 Option to switch: Desalination



With the option to switch the suitability for the NPP to work in a <u>flexible</u> load following mode is studied. This operation mode gives the required flexibility to the operator to take advantage of the evolution of the market prices.

SCENARIOS' DEFINITION

Even in this case, studied scenarios consider different prices of products. Indeed, the switch can add worth only in a particular combination of prices. For example, if the water is much more expensive than E.E., the plant would work always in off-design mode, maximising the usage of the desalination plant. Also the vice versa is truth.

Therefore, the prices for E.E. are chosen as in Table 39: they are the values of the electricity in excess only in the nocturne window 2-6 a.m.. So the E.E. price fluctuate as usual during its daily trend, but to take account of an eventual excess of electricity, it is studied the possibility that in that narrow window the price could be fixed at a conventional value.

The price of water instead is varied in the range 0,4 – 2,8 $\$/m^3$. The Figure 57 is inserted to make a comparison between the chosen range and the real price in some countries. The prices reported in that figure are the buy price for the customer. The selling price for a company is usually one third, because in the middle there are expensive distribution costs.

The range of water price is divided in 20 intervals. Consequently the correct number of simulated scenarios is 5x20=100. The experiment plan is reported in Table 40.

TABLE 39. ELECTRICITY PRICES FOR 5 SCENARIOS STUDIED FOR THE OPTION TO SWITCH.

SCENARIO	E.E. price [$/kWh]
1 – France	0,00
2 – cheap electricity	0,02
3 – standard electricity	0,04
4 – night price	0,06
5 – market price	variable

FIGURE 57. PRICE OF WATER FOR THE CUSTOMER IN DIFFERENT COUNTRIES [170].

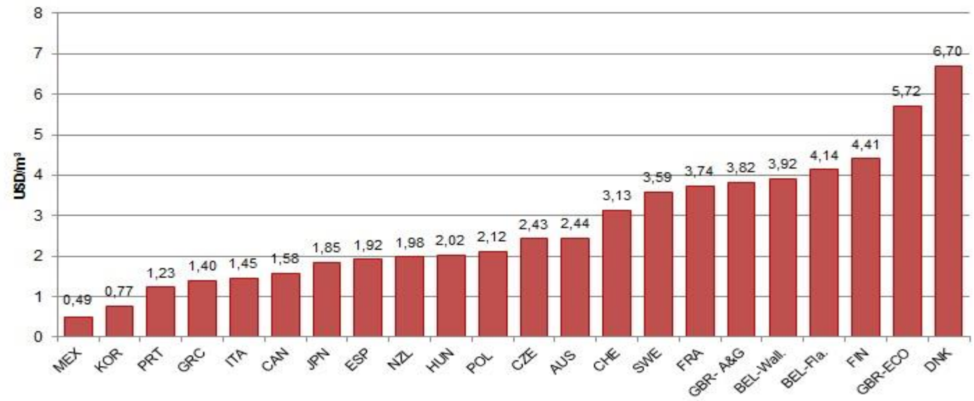


TABLE 40. THE COMPLETE EXPERIMENT PLAN FOR THE OPTION TO SWITCH.

		Water [$/m³]																			
Scenario	E.E. [$/kWh]	0,4	0,6	0,8	1,0	1,1	1,2	1,3	1,35	1,4	1,45	1,5	1,55	1,6	1,8	1,9	2,0	2,2	2,4	2,6	2,8
1 France	0																				
2 cheap	0,02																				
3 Standard	0,04																				
4 Night	0,06																				
5 Market	variable																				

EXEMPLIFICATIVE RESULT

For every scenario the following values are investigated:

- option value: difference between static load following and flexible load following
- option value: difference between NO load following and flexible load following
- percentage of the NPP used capacity
- percentage of the MED-TVC used capacity

FIGURE 58. HOW TO CALCULATE THE VALUE OF THE OPTION. THE OPTION TO SWITCH HAS WORTH ONLY IN A SPECIFIC COMBINATION OF PRICES OF WATER AND E.E.

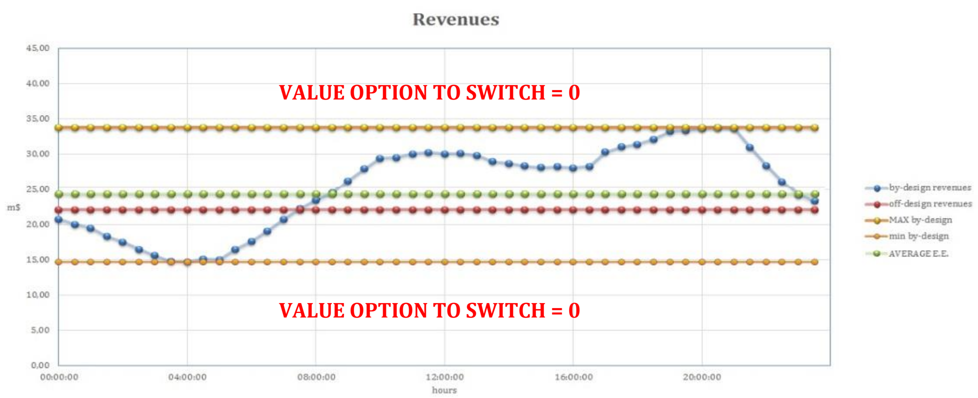


FIGURE 59. VALUE OF THE OPTION TO SWITCH IN RELATION WITH WATER PRICE. COMPARISON BETWEEN A FLEXIBLE AND A STATIC LOAD FOLLOWING.

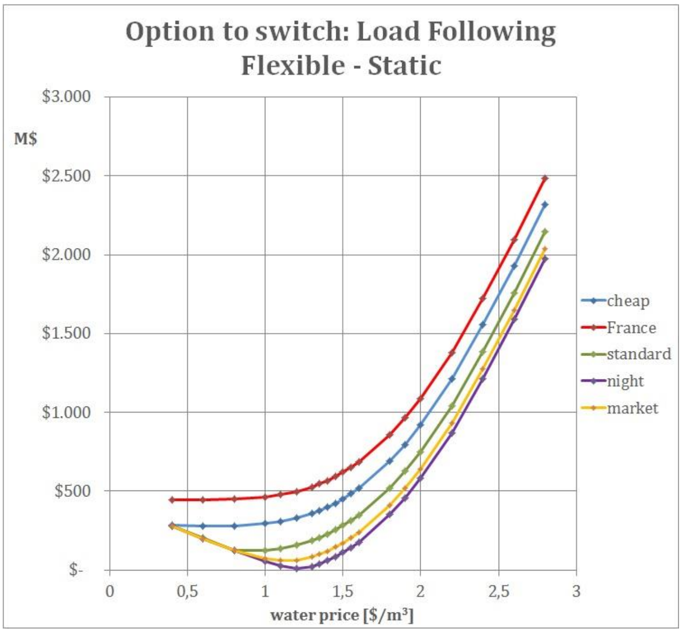
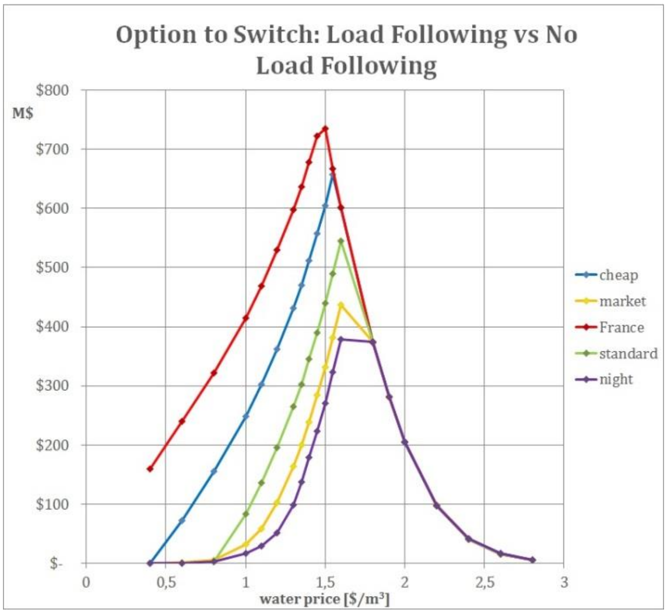


FIGURE 60. VALUE OF THE OPTION TO SWITCH IN RELATION WITH WATER PRICE. COMPARISON WITH NO LOAD FOLLOWING MODE.



As it is possible to see in Figure 58, the option to switch gives a worth only when prices of water and E.E. give revenues that are comparable. This is the meaning of the lines:

- blue = revenues from by-design operation mode (E.E.)

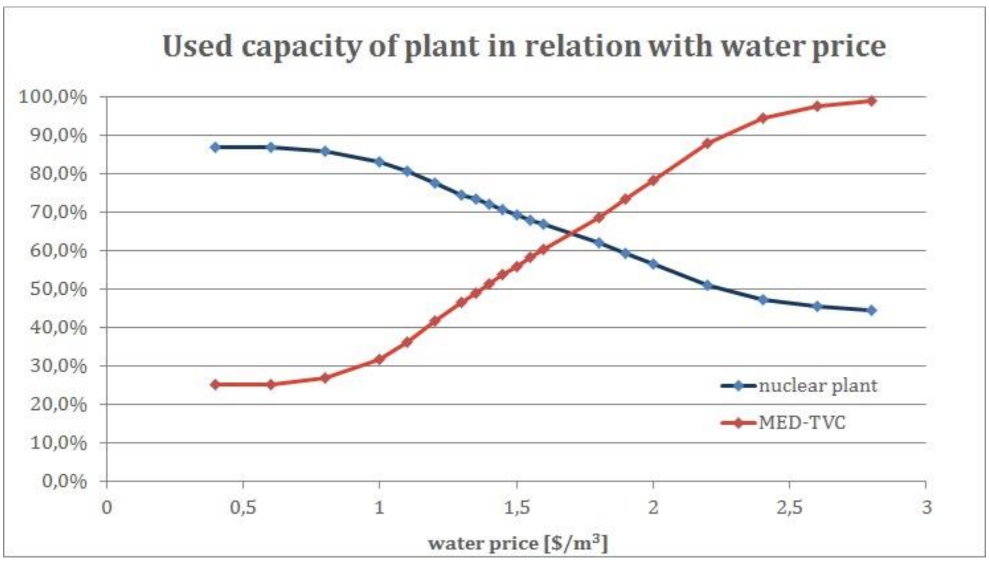
- red = revenues from off-design operation mode (water)
- green = average revenues from by-design operation mode
- yellow = MAX and min revenue from by-design operation mode

Figure 59 displays the difference between the revenues gained with the flexible and the static load following. Figure 60 instead displays the same difference, but the flexible load following and the no-load following: if the off-design revenues are lower than the average by-design revenues (red line is lower than the green one), the no-load following considered is the case of “always by-design operation mode”. Otherwise the no-load following considered is the case of “always off-design operation mode”. Both of the graphs are important to understand the value of the option: for example, it is possible to consider a water price of 2,8 $\$/m^3$. Looking at only Figure 59, one could think that the value of the option is maximum. Moreover one could think that the value could even rise, looking at the trend of Figure 59. Instead the Figure 60 remembers that upper a fixed value, the option to switch has no worth, because it could be better to work always in the off-design operation mode. Consequently the value of the option displayed in Figure 60 is zero.

Therefore, if the red line is outside the area limited by yellow lines, the option to switch has no worth. Whether it is under the yellow line on the bottom, it is not reasonable switching on the desalination plant, because the by-design mode can always guarantees higher revenues. On the contrary, whether the red line is above the upper yellow line, it is advisable to produce as much as water possible. Consequently, the more interesting case is within the yellow lines.

The Figure 59 underlines the advantage to have a flexible load following. Remembering that the static load following mode works in off-design only between 2 a.m. and 6 a.m., if the price of water is too cheap, the advantage given by the flexibility regards the fact that in this case the plant never switches in the off design production arrangement. This is confirmed looking at Figure 61. The plant starts to work in off-design mode only few hours per day, only if the price of water is at least 0,8 $\$/m^3$. Otherwise the used capacity of the MED-TVC is fixed to 25%, that is the assumed the minimum constant working level.

FIGURE 61. PERCENTAGE OF USED CAPACITY FOR BOTH THE PLANTS, AT VARYING OF WATER PRICE. RESULTS DISPLAYED FOR SCENARIO 5 (MARKET PRICE).



Then in the range of 1-1,5 $/m^3 the advantage of the flexible load following is minimum in comparison with the static one, because the suggested share between the two operation modes is very close to the static one. However there is always an advantage with the flexible mode, because it is possible to catch the perfect time for the switching. Instead the advantage respect to a plant working always in by-design becomes very important (see Figure 60). The advantage is even more pronounced if the E.E. is sold with a cheaper price.

The optimal situation is when the water price is in the range 1,5-2 $/m^3.within this range, the value of the option is important both in comparison with the static load following, and in comparison with the no-load following (always off-design production). Within this range, there is also a good compromise regarding the usage of the plants: the used capacity is approximately 60-70% (see Figure 61). When the price of water rises over the value of 2-2,5 $/m^3, the option loses worth because it becomes advisable to produce as much as water possible: indeed in Figure 61 the used capacity of the desalination plant rises over the 95%.

6.3 GENERAL DISCUSSION OF RESULTS

6.3.1 Biorefinery

In order to make a comparison in the same order of magnitude with the desalination case, approximately the same percentage of used electric capacity of the nuclear site has been chosen as confront parameter.

This required the study of a large size biorefinery, of which the total capital costs are about 1,5 b$. The investment cost is subdivided as in Figure 62.

With a fixed biodiesel price at 1,5 $/L, the dependence of the IRR of the investment to the size of the biorefinery is displayed in Figure 63. The trend of the curve underlines the dependence to the economies of scale (exponential value h=0,8).

FIGURE 62. SHARE OF INVESTMENT COSTS FOR THE BIOREFINERY, BETWEEN DIFFERENT PROCESSES..

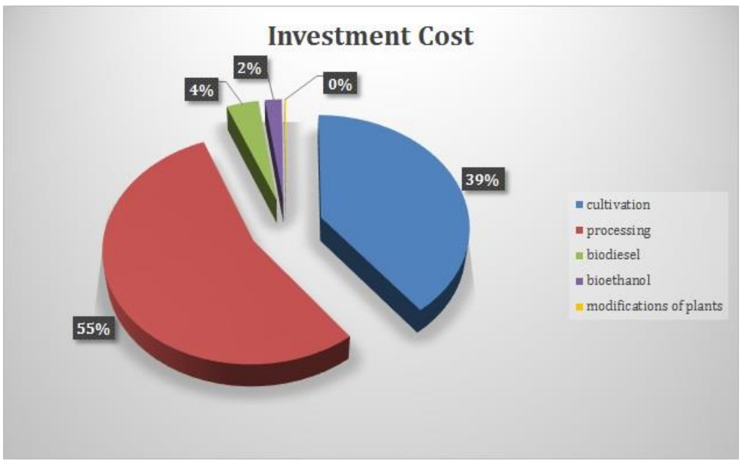
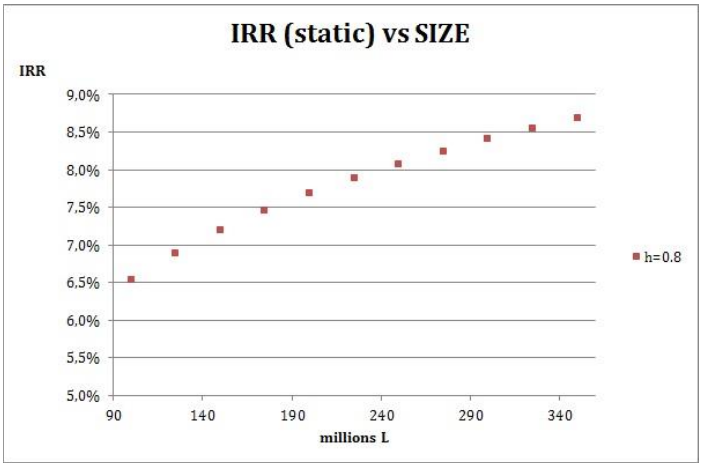
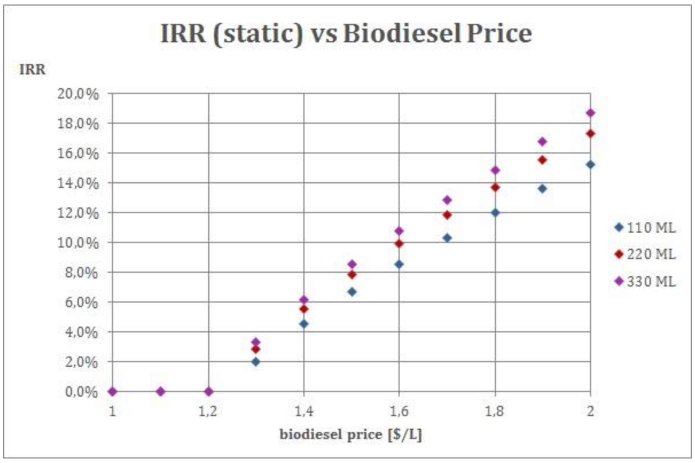


FIGURE 63. DEPENDENCE OF THE IRR TO THE SIZE OF THE BIOREFINERY. RESULTS CALCULATED WITH THE DCF STATIC AND BIODIESEL PRICE OF 1,5 $/L.



The price of biodiesel chosen to investigate this relation is a standard price, and it is also a price very close to the break-even point. Indeed in Figure 64 the relation between the IRR and the price of biodiesel is displayed.

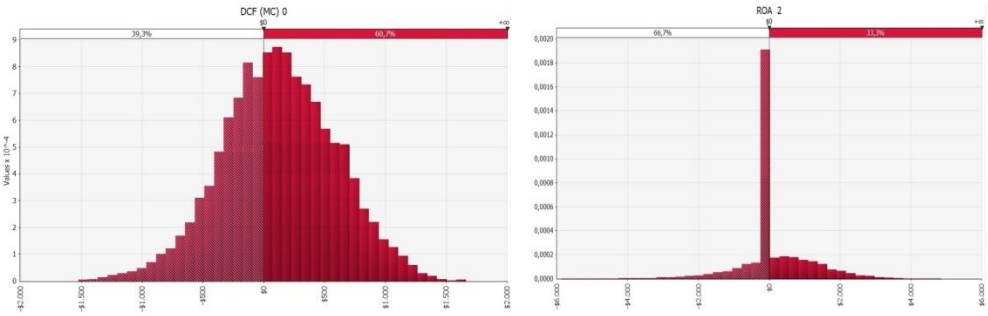
FIGURE 64. PROFITABILITY OF THE INVESTMENT, IN RELATION TO THE PRICE OF BIODIESEL.



Therefore, the classic DCF suggests that the installation of a large algal biorefinery could be profitable, because the break-even point for the price of biodiesel is in the order of magnitude of the real prices. Anyway a strong dependence to theses prices of products is underlined.

The ROA analysis confirms what found with the DCF method. In particular it is possible to assert that the ROA analysis is particularly useful to analyse negative scenarios. Indeed, if the scenario is already attractive, it isn't advisable to wait for the evolution of prices. It is not possible to take advantage from the option to build. The option to build permits to enhance the mean of the calculated NPV because it doesn't consider (cut) the negative tail that, on the contrary, is considered in the DCF MC analysis (see Figure 65).

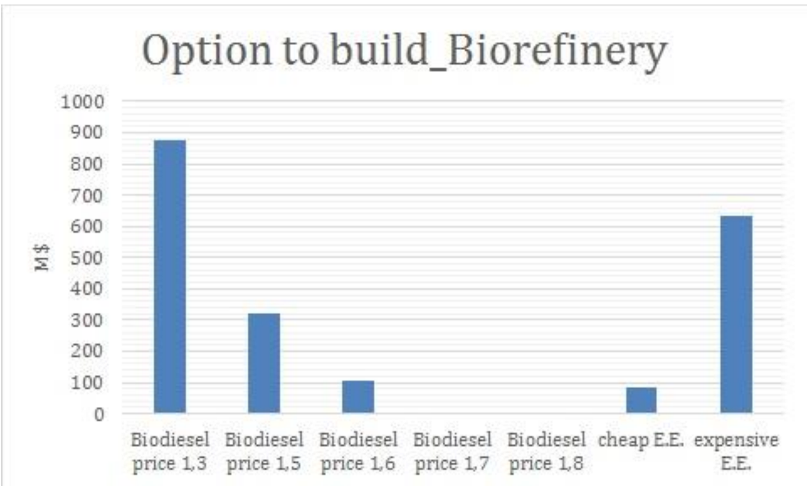
FIGURE 65. RESULTS OF A SIMULATION. ON THE LEFT THE GRAPH OF THE NPV CALCULATED WITH A CLASSIC DCF MONTE CARLO. ON THE RIGHT, THE SAME GRAPH HAS BEEN CALCULATED WITH ROA AT YEAR 2.



This is reasonable in comparison to the real life, because it's sure that a manager wouldn't consider the opportunity to build a plant if the situation is not attractive. Therefore it is right do not take into account some very negative evolutions of uncertainties. All this negative scenarios are collected on the point NPV = 0.

Therefore the option to build confirms the suitability of such an investment. In addition, the worth of this option increases significantly for more negative scenarios, when for example the biodiesel is cheap or the electricity is more expensive (see Figure 66). This is possible because this option acts like a life preserver: when the scenario is negative, it advises to don't invest; otherwise when the evolution of uncertainties is attractive, the option is able to take full advantage of the profitability of the project.

FIGURE 66. WORTH OF THE OPTION TO BUILD FOR THE BIOREFINERY.



6.3.2 Desalination

Also in case of desalination, the investment required is very important (about 1 b$). For this reason, as it happens for the nuclear plant, the literature suggests to use as much as possible the installed capacity of the MED-TVC. However, the possibility to save the costs of energy, seems to permit to have a break-even point for the price of water that is in the order of magnitude of the real water prices. The break-even point of water price is displayed in Figure 67, in relation to the nocturne price of E.E.. The figure shows that if the market is not saturated during the night and therefore the E.E. is not in excess and consequently sold at normal nocturne price (about 0,05-0,06 $/kWh), the price of water to reach the break-even is quiet high: about 2-2,2 $/m^3.

On the other hand, if the market is saturated and the requirement to work in load following mode becomes fundamental, it is reasonable to assume that the E.E in excess could be sold with a lower price (even not sold: case of France where control rods are inserted). With these scenarios, the requirements in matter of the price of water are less demanding: about 1-1,8 $/m^3.

FIGURE 67. COMPARISON BETWEEN THE PRICE OF WATER AND THE PRICE OF E.E. TO REACH THE BREAK-EVEN POINT.

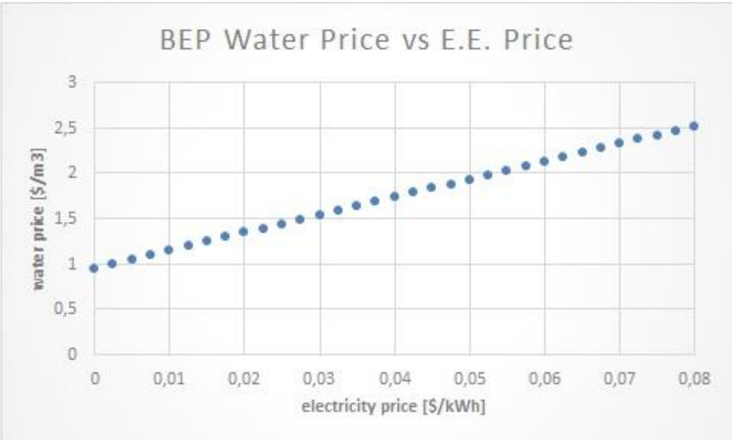
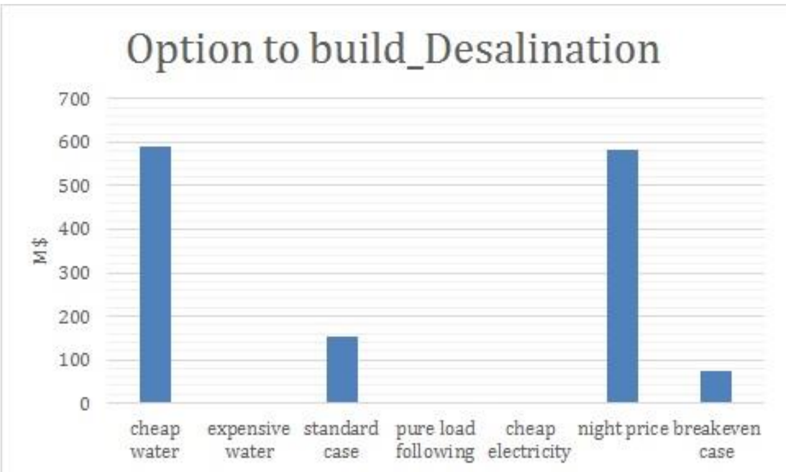


FIGURE 68. WORTH OF THE OPTION TO BUILD FOR THE DESALINATION PLANT.



Even in this case the profitability of the investment could improve thanks to the option to build, especially if the scenarios is highly negative at the beginning.

Therefore the value of the option highly depends on the scenario. However the order of magnitude of this worth is comparable between the biorefinery case and the desalination case.

Finally, observing all the graphs inserted in Appendix E., the maximum value of the option is usually reached at year 4-6. It means that, if the investor doesn't own a historical data collection in matter of market prices, after approximately 5 years it is possible to develop a precise forecast, and to limit the risk associated to the investment, caused by the uncertainties of the market.

Instead, regarding the option to switch, it is interesting to observe how the ranges of water prices obtained to maximise the option, match perfectly with the range required to reach the break-even point. In addition, if the option to build seems applicable mostly with negative scenarios, the option to switch depends only by the combination of prices of water and E.E..

Subsequently the option to switch seems to have a greater potential and applicability. Finally, it is not a coincidence that the most attractive water price is about 1,7 $\$/m^3$: actually with this price there is a perfect compromise in matter of the share between the weekly usage of the NPP and the MED-TVC (both at about 65% of their nominal capacity).

6.4 METHODOLOGICAL DISCUSSION: SENSITIVITY ANALYSIS

In this paragraph the sensitivity analysis of the economic models is reported. In the Appendix E. all the main graphs are inserted for the sensitivity analysis. The latter is studied for both the biorefinery and the desalination. The analysis is done for the DCF MC and for the ROA at year 5. In addition the analysis for the Monte Carlo is done with and without the prices of products (that are doubtless the main variables of the model) to zoom the dependence of the results from other parameters (such as costs, discounts rate, escalation index, etc.).

BIOREFINERY

As said, the most important variables are the prices of biodiesel and electricity (Figure 69 confirms these dependences). For this reason they are well-investigated with particular scenarios within this work.

A part from that, the other important parameters are the discount rate of the cash flow (WACC), the O&M costs for the processing phase, the average drift of the prices and the capital cost of the plant.

It is interesting to note how the final output doesn't depend much from the production of ethanol: this confirms that in this case it is only a secondary source of revenues. Furthermore, looking at Figure 70 that shows the sensitivity analysis at year 5, it is possible to note how the yearly drift of prices acquires more importance. Instead the escalation costs (especially that one of the plant) don't affect considerably the final results.

FIGURE 69. TORNADO GRAPH FOR THE DCF MC OF THE BIOREFINERY.

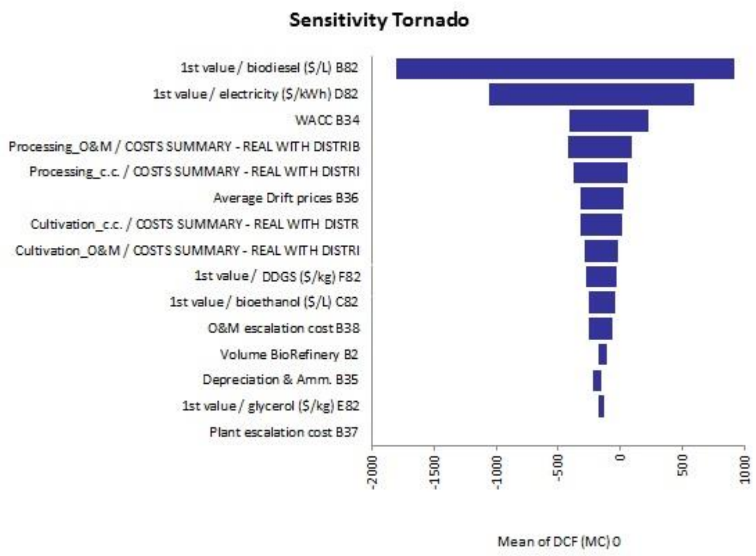
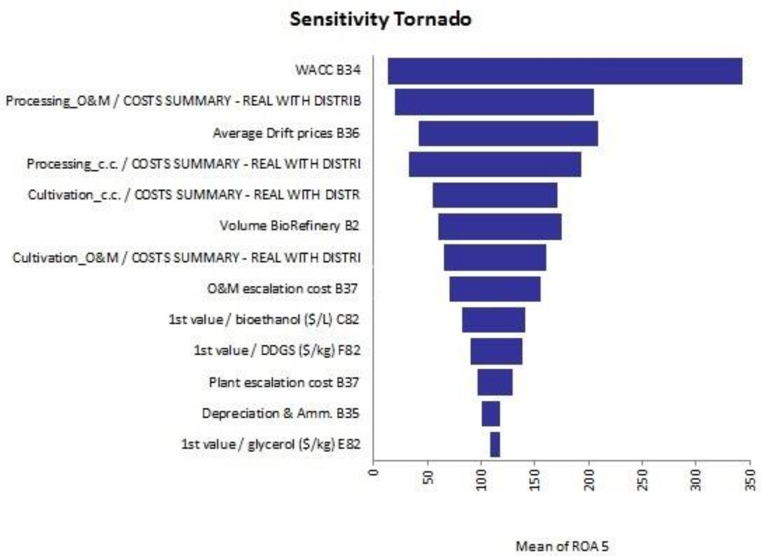


FIGURE 70. TORNADO GRAPH FOR THE ROA AT YEAR 5 FOR THE BIOREFINERY.

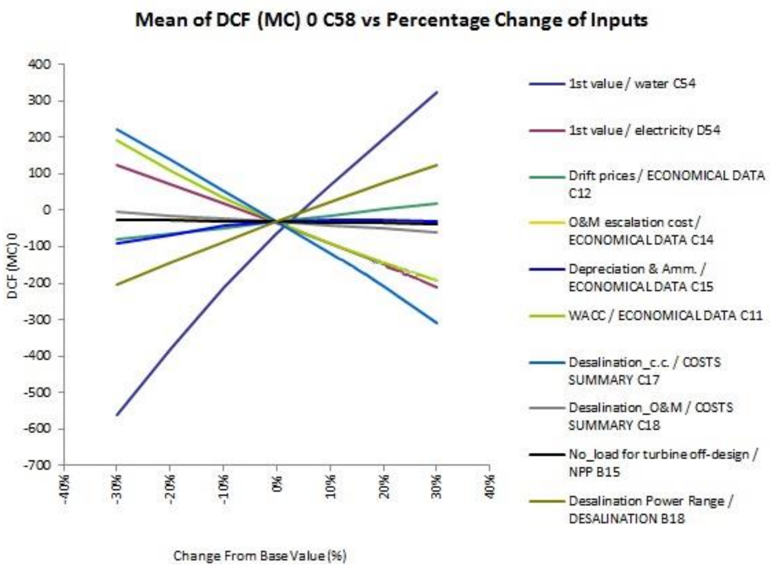


DESALINATION

Even in case of desalination, the most significant variable is the price of the main product (water in this case). In contrast with the biorefinery, the E.E. price has a less central role. Probably because in case of static load following, the quantity of not-sold electricity plays a minor role, especially if this is sold with a cheap price. Instead the capital cost of the desalination plant is peculiar for the profitability of the investment. Finally also in this case, the WACC has a key role.

About technical parameters imposed, the minimum no-load factor don't affect considerably the results. On the contrary, the minimum level chosen for the desalination plant in by-design arrangement owns a discrete importance.

FIGURE 71. PERCENTAGE CHANGE GRAPH FOR THE DCF MC FOR THE DESALINATION.



CHAPTER 7. CONCLUSIONS

Nuclear energy represents a proved opportunity to tackle most of the criticalities in the energy field, such as the growing energy demand, the greenhouse gas emissions, the scarcity of fossil fuels sources at cost-effective price, etc. On the other hand, to be competitive in the long-term energy needs, various problems must be addressed, regarding waste, safety, security, and non-proliferation issues. Therefore an important goal to promote a wide adoption of nuclear energy is to improve its sustainability.

Considering this energetic-economic-social background, the idea of this work is to investigate the future role of nuclear energy, in particular the role of SMRs, since they are currently surrounded by great interesting from scientific community and nuclear industry.

Between the aspects that nuclear technology has to face, there is also the load following of the grid. Although nuclear energy has been mainly seen as a base load source, the evolving energetic portfolios and the recent limitations posed by institutions require to work in load following mode even for nuclear power plants. Nuclear technology is very capital intensive and most of the costs are fixed. Consequently there is not advantage to reduce the power production, because the lower O&M costs are less important than the lower load factor. To take full advantage of a nuclear reactor, the load factor has to be as high as possible. Moreover, currently control rods and neutrons absorbers are inserted into the vessel to follow the grid. To insert a reactivity control introduces within the primary loop also various thermo-mechanical stresses. Therefore, the idea of this work is to address the thermal power produced in excess during low-load hours (i.e. during the night) to other plants, coupled with the NPP in a cogenerative layout.

For these motivations, the results of this thesis try to give an answer concerning the flexibility of nuclear power (SMRs in this specific case), as an alternative source even for not-electrical applications, such as the production of biofuels or desalinated water.

The proposal studied here is the suitability to produce both a worthwhile product like the Electric Energy, and some alternative not-electric products, using a cogeneration layout. This fact allows to take full advantage of the thermal power generated within the core and to enhance the employment of the low enthalpy heat, required for example by a biorefinery and a by desalination plant.

The strong points of the developed models are a deep analysis of literature information in matter of technical considerations and values. Instead from the economic point of view, the models carefully consider the uncertainty that characterizes the evolution of prices, in an undefined and rapidly evolving financial markets. In addition, the algorithms permit to study how the profitability of an investment changes in relation to an increasing acquisition of market data.

The obtained results can be classed in technical and economical. The main **technical results** are the following:

- between all the possible technologies to cultivate the biomass for the biorefinery, only the fermenter scenario is usable for our purposes. All other technologies require much space (thousands of hectares) for a reasonable coupling with a plant, that has an installed power in the order of magnitude of GW. The fermenter scenario however reduces part of the environmental benefits and, consequently, of the sustainability
- the studied biorefinery can't work in a flexible way, because of the perishability of the biomass and the importance of the power requirements in the first (constant) steps of the production chain. Consequently the biorefinery is not suitable for the load following
- on the contrary the desalination plant gives this flexibility to the nuclear site: it is possible to work in load following mode
- the required size of the plant is similar to the largest plants worldwide
- a minimum quantity of steam must be supplied to the steam turbines, even when they are not producing E.E. (no load value equals to 7,8%), in order to prevent superheating issues
- it is also advisable to supply a minimum quantity of power even to the MED-TVC plant during the daily hours. This fact is not mandatory to prevent technical problems on the system or on the material, but it is reasonable from a managerial point of view. Actually the plant begins to produce a sealable quality of water, at least at 20-30% of the power load. Consequently, in order to prevent losses and to take better advantage of a capital intensive plant, it is assumed to work with a minimum power level of 25% also during the daily hours.

Therefore in this case, the suitability of working in load following mode, would be limited only by economical (and not technical) considerations. Indeed, because of the desalination plant is capital intensive and therefore it is important to maximize its availability, in this thesis not only one scheduled working mode has been studied. A range of utilization factor is taken into consideration to study the profitability of the investment in relation to the used capacity of the plants.

The **economical results** confirm what is suggested by the Real Options theory. If there is uncertainty about the outcome of an investment, the ROA can evaluate more positively the profitability of this project in comparison to what it is obtained with a classic Discounted Cash Flow method.

In particular, with the *Option to Build*, the approach is more promising with scenarios with a wide NPV standard deviation around zero. Actually, if the NPV calculated with the Monte Carlo DCF is already very positive because the investment is very attractive, the ROA theory can't add an extra value to the found result. Another outcome of this work regards the necessary years to better forecast a scenario. As it is possible to see from all graphs inserted in the Appendix E., after 4-6 years most of uncertainties are solved. This is affirmed because after 4-6 years it is usually reached the maximum value for the NPV or an asymptotic value is approached.

Final considerations regard positive/attractive scenarios:

- At year 0, the ROA usually can't forecast the profitability of an investment. The reason comes from the way the algorithm is built and prices are simulated. Actually the Brownian component of the price fluctuation is quite important and consequently, even if the drift could be positive, the model doesn't recognize the correct trend. If the scenario is positive the best way to act should be to always invest. Instead, the developed model sometimes suggest to don't invest and consequently the profitability calculated with the NPV is lower.
- Even if at year 0 the ROA method can't provide a reasonable result and broadly the ROA is not more useful than the DCF, the trend of the NPV obtained with the ROA at different years confirm the theory. It means that the value grows in the first years, it reach a maximum (after 4-6 years) and then it starts to decrease. This is reasonable because the investor gathers more information during the first years. Then this advantage become less important than the fact that the revenues are discounted with a higher discount factor.

Also the *Option to Switch* confirms to add an extra worth to the project given by its flexibility. However, this option can't always guarantee a fixed advantage. The magnitude of order of the advantage given by the possibility to switch, strongly depends by the combination of prices of water and E.E.. Extremely related to the prices of the products, also the utilization's percentage of plants' capacity varies accordingly. Consequently, the option to switch allows to study also the profitability in relations to prices and utilization's factor of plants. The best compromise seems to be when both the plants work at 60-70% of their designed capacity[19].

The final relevant result is that the calculated order of magnitude for the prices in order to reach the break-even point is of the same order of magnitude of the real market prices. This statement suggests that the modelled plants can be profitable even from an economic point of view. Moreover, the cost-effectiveness can enhance if some state incentives are proposed in order to sustain the production of these "green" products. Actually, what it usually happens to help the green economy is to tax pollutant products (i.e. conventional fossil fuels) for supporting green energy (e.g. photovoltaic panels).

Therefore, the biorefinery could be attractive, but results obtained with the option to build strongly depend from the scenarios. On the contrary, the simple installation of a desalination plant (option to build) has similar results of the biorefinery case; but the possibility to have a flexible production (option to switch) could add an ulterior value to the project. Consequently, the desalination case seems more interesting.

However, some approximations are introduced to obtain the results. Firstly, the investment item is considered as instantaneous. It has a symbolic duration of one year, without taking into account the real negative cash flow profile. This approximation limits this study, because it is not consider any possible delays in construction, or the importance in matter of costs of different construction phases,

[19] For the NPP, the capacity refers to the electric one. Actually, the reactor (primary circuit) always works at full power rate.

such as shipping, installation or licensing. Consequently it is not considered how these items affect the investment financing (i.e. interests of debts, delays of revenues, etc.). Secondly, the Pert distribution that gathers the possible specific costs for the plants is symmetric. This assumption was necessary to assure a correct working way of the economic models. Instead, in the real world, it is more probable to have an increase of forecasted overnights costs, in comparison to their decrease.

Consequently, further studies about this project can regard the following aspects:

- To develop a more precise analysis of the outlet cash flow profile, for the investment phase. For example the modularity of the desalination plant could be useful to better predict the whole investment cost (through a learning process). From the financial point of view, this could affect the total costs
- To study if and eventually how, this connection with the green economy affects the acceptance of nuclear energy. From the social and economic point of view, it could be interesting to study how the acceptance could influence the project
- in case of developments and improvements of algae technology, perhaps the case of the biorefinery could become more interesting. For example, if algae were produced at lower costs, the biorefinery coupled with the NPP could simply transform the dried algae in biofuels, without considering the cultivation and dewatering phases. In this way, load following could be possible even with the biorefinery case
- To enhance the availability of the desalination plant, during the daily hours. A way to pursue this goal could be to add a secondary energy plant to drive the MED-TVC plant. For example, during the day it could be possible to burn gas, while during the night the energy in excess from the nuclear power could be addressed to the MED-TVC. However this solution highly contrast with the pursuit of the green economy
- To study how different sizes of the SMR could affect the load following. Thanks to the high modularity degree of the nuclear plant, having smaller basic module could guarantee to better match the demand of E.E. In this way it would be possible to better catch the live volatility of prices.

APPENDIX A.

ROUTE TO BIOFUELS

In this Appendix, all the production phases for the biofuels are deeply analysed.

A. Cultivation

Three main different layouts currently exist for the cultivation of microalgae: open ponds, photobioreactors (PBRs) and fermenter tanks. In open ponds and PBRs the biomass grows autotrophically. Autotrophic cultivation occurs when the microalgae use light, such as sunlight or artificial light, as the energy source, and inorganic carbon (e.g., carbon dioxide) as the carbon source to form chemical energy through photosynthesis [171]. Some microalgae species can not only grow under phototrophic conditions, but also use organic carbon under dark conditions, just like bacteria. The situation when microalgae use organic carbon as both the energy and carbon source is called heterotrophic cultivation [172]. In this process microalgae are grown on organic carbon substrates such as glucose in stirred tank bioreactors or fermenters. In this case, algae growth is independent of light energy [80]. In this paragraph the main characteristics for every system will be delved. Finally a comparison will be worked out to prove the reasons of the system chosen for this study.

OPEN PONDS

The large-scale outdoor culture of microalgae and cyanobacteria in open ponds is well established [173]. Nowadays, algae are cultivated for pharmaceutical, cosmetic and food industries. Open pond systems are usually shallow systems (between 10-50 cm, to allow appropriate illumination) in which the microalgae are grown. The ponds can be excavated and used unlined, or lined with impermeable materials, or built up with walls of concrete or other materials. The culture medium is directly exposed to the atmosphere, allowing liquid evaporation and thus helping to regulate the temperature of the process [174]. The successful culture of microalgae in outdoor ponds is limited to a small number of algae species, which can tolerate extreme environmental conditions to the exclusion of most other species. These include fast growers such as *Chlorella* and species that require highly selective environments such as *Spirulina* and *Dunaliella* which thrive in highly alkaline or saline selective environments. There are four main types of open ponds: unmixed open ponds, raceway ponds, circular ponds and inclined ponds [26]. Raceway ponds are the most common type of open pond currently in practice and they are widely used for the commercial cultivation.

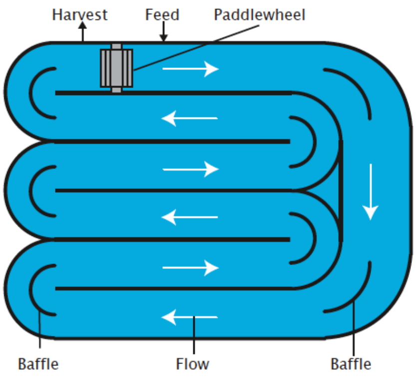


FIGURE 72. A SCHEMA OF A RACEWAY POND SYSTEM.

These ponds incorporate low-energy-consuming paddle wheels for agitation and mixing of the gas and liquids of the cultures, and for circulation too (see Figure 72). A continuous agitating paddle wheel provide the mixing and circulation of the broth. Ahead the wheel, the medium is fed with algae, they grow in the pond until they don't reach the end of the raceway behind the wheel. Then they are harvested. The pond is directly exposed to the atmosphere. Reported flow rates range from 10-30 cm/s [175], with depths ranging from 10-30 cm and individual ponds being up to 1 ha in area. Much higher productivities have been reported from raceway ponds compared to other open ponds system. For these reason they have been selected as benchmark for open ponds cultivation system. The microalgae's CO_2 requirement can be satisfied from the surface air, but a submerged system, composed by pipes and aerators, may be installed to enhance the participation of CO_2 in the medium [176]. Flue gases from close factories that contain CO_2 at concentration ranging from 5 to 15% (v/v) have indeed been introduced directly into the ponds of various configurations that contain several microalgal species [177]. In Figure 73 an example of a commercial scale raceway ponds layout for cultivation of microalgae in USA.

In a raceway pond there is a continuous circulation of the medium, that requires energy for the puddle wheel and for pumping both feed water (inlet) and harvested medium (outlet). The inlet water stream is needed to replace the portion of the culture that is harvested daily [178] and the water lost for evaporation. Usually the 25% of the entire batch is harvested every day, therefore the medium needs 4 days

FIGURE 73. ALGAE OPEN POND FACILITY IN HAWAII (USA).

to complete the cycle in the raceway pond. A cell concentration of about 0,5 g/L (grams of alga per litre of medium) is usually obtained in open systems [179]. Hence, according to the literature, in this study the following energy requirements have been considered for the cultivation phase in an open pond system: pumping of water (inlet/outlet), circulation of gases (CO_2) and mixing and circulation of medium (thanks to the puddle wheel). The critical analysis of data about energy consumptions will be performed at the end of this paragraph (A).

PHOBIOREACTORS

Closed PBRs were developed to overcome the problems associated with open pond systems. A wide variety of PBRs have been designed and built ranging from tubular and cylindrical systems, conical systems, helical systems, flat plate systems, to flexible tubing coiled around a cylindrical framework or vertical bubble columns and airlift reactors [26]. They can be located indoors and provided with artificial light or natural light via light collection and distribution systems [14] or outdoors to use sunlight directly. The former option currently involves complex and costly light collection and distribution systems or the use of artificial light and are not feasible for the production of algae for commercial bioenergy applications [180]. Consequently, only the latter options will be briefly analysed here. Indeed, traditionally, photobioreactors have suffered from problems of scalability, especially in terms of mixing and gas exchange (both CO_2 and O_2)[20]. In Figure 74 it is reported as example of a commercial cultivation PBR layout system. PBRs lose much less water than open ponds due evaporative cooling and so temperature must be carefully maintained. Open ponds, however, are subject to daily and seasonal changes in temperature and humidity. Photobioreactors are unlikely to be sterilizable and may require periodic cleaning due to biofilm formation. PBRs can also provide a higher surface to volume ratio and so can support higher volumetric cell densities, reducing the amount of water that must be processed and thus the cost of harvest [11]. Raceway ponds are made of less expensive materials, their construction involves lower costs and they require less energy for mixing [181]. All the advantages and disadvantages, also in relation with fermenter tanks, are gathered in the Table 41.

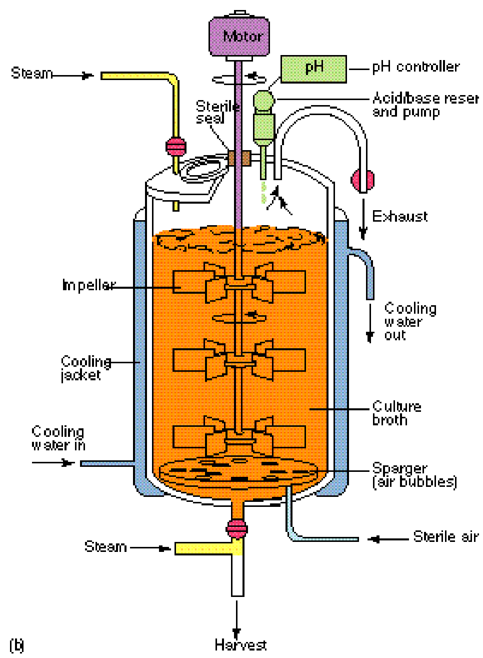
FIGURE 74. THE COMPLEXITY OF THE PBRS. IN PARTICULAR HERE IN THE PICTURE A HORIZONTAL TUBULAR, NATURALLY ILLUMINATED, PHOTOBIOREACTOR SYSTEM.

[20] PBRs are a phototrophic cultivation system, as open ponds. So PBRs require circulation of auxiliary CO_2 as carbon source. However, on the contrary of open ponds, PBRs are closed systems and then they require a supplementary circulation of O_2, produced during photosynthesis reaction and that has to be drained. This operation can't be done easily, like in open system that are in direct contact to the atmosphere and can release gases in the air.

FERMENTER TANKS (or FERMENTORS)

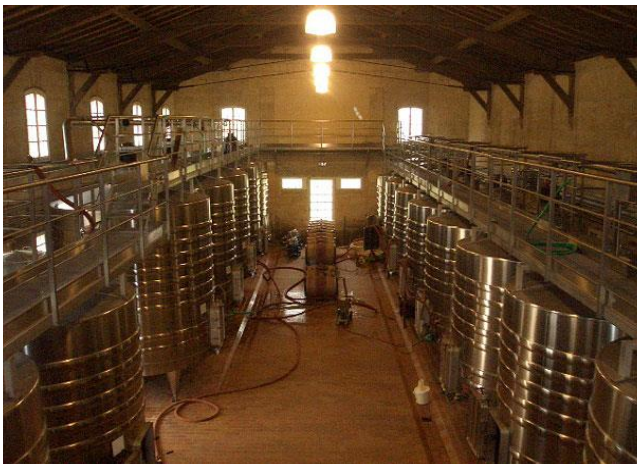
While most algae grow phototrophically, some are capable of heterotrophic growth using organic substrates as the sole carbon and energy sources, as discussed in paragraph 2.2. This way of algal cultivation is well established [182] and has several advantages over phototrophic modes of growth (see Table 41). These include the large, existing fermentation technology knowledge base, the high degree of process control for consistent, reproducible production, the elimination of light requirements, the independence from weather and climatic conditions, and lower harvesting costs [183]. Generally, heterotrophic cultivation has been found to increase the total lipid content in algae compared to phototrophic cultivation grown cells [184]. Furthermore, heterotrophic cultivation of algae usually results in higher yields [185]. One more advantage is the possibility to utilize inexpensive sugars (like starch) for carbon source [11]. Fermenters are simply a kind of stirred tanks in which water, inoculum, carbon source (starch) and all the required nutrients are inserted. Fermentation process is very delicate, from the chemical point of view. It's fundamental to reach and maintain an optimum level of PH and temperature inside the fermenters. Also, before to fill the tanks, an accurate sterilization process has to be performed to avoid the contamination of any kind of bacteria. For the same reasons, the inlet water of the medium has to be fresh, pure and distilled, such as that one produced via a thermal process in a desalination plant. As it is possible to see in Figure 75, only 70-80% of the volume of the reactor is filled with liquid so that sufficient amount of head space is available to hold foam if formed [26] and the cultivation time takes about 7 days (see paragraph 2.4.2). The medium inside is stirred, thanks to the usage of some impellers. Air is introduced from the bottom to supply a correct amount of oxygen. The bottom is usually made round to enhance the mixing capability inside the tank. Thermal and chemical condition are also very important and therefore a ph controller and a cooling system that surround the walls are usually installed.

FIGURE 75. THE INSIDE OF A FERMENTER.



Common fermentors used in today's industry include the following: tray fermentors, static bed/tunnel fermentors, rotary disk fermentors, rotary drum fermentors, fluidized beds, agitated tank fermentors, and continuous screw fermentors [186]. One of the most common fermentors used on a large scale, and the type this thesis is concerned with, is the agitated/stirred tank fermenter. The stirred tank fermentor can be divided into two subsets: the bioreactor, which is used for mammalian cells, and the fermentor which is used for bacteria, yeasts, and algae. The fermentor is typically used for cells that are more robust, tolerant of high shear rates and have higher oxygen demands. The oxygen demand of the cells in stirred tank fermentors plays a vital role in cell culture and growth. For this purpose, fermentors typically employ a sprinkler located near the bottom of the tank to introduce air (sterile air) into the broth. Fermentors also employ one, or several, impellers to provide bubble break up and bulk mixing of the media (see Figure 75). Then, some baffles are inserted in the tank, attached to the interior walls to promote mixing. During the design phase of a fermentor, the number and the spacing between impellers are key factors to optimize efficiency of mixing and power consumption. Industrial fermentors are characterized by large volume: up to 250 m^3 [187]. However, one other important design parameter is high-to-diameter ratio [188]. Typical values range between 2 and 3; however, taller tanks (up to H/d=4) have been used to reduce the power requirement of the impellers [189] and the area occupied. Typical tanks also employ a dish-shaped bottom to enhance mixing and prevent dead zones. The most difficult task in tank design is getting the fermentor capability to match the oxygen demand of the fermentation culture [190]. When designing a stirred tank fermentor, the main concern is providing sufficient oxygen to the cells without exceeding any limits of shear or power consumption. Actually, to obtain higher oxygen absorption, increasing the impeller speed could be a solution. However this causes a great increase in impeller tip speed which can damage the organisms because of the increased shear. Also it creates an exponential increase in power consumption. In order to avoid these pitfalls, correlative models can be used by imposing limit on power consumption and impeller tip speed. With this, the rest of the tank can be designed to match the oxygen demand of the organism being fermented [188]. Here, not all these specific design parameters have been considered: only the area occupied by fermenters and the power requirements are further discussed.

FIGURE 76. THE INSIDE OF AN INDUSTRIAL WAREHOUSE WHERE SOME FERMENTERS ARE ALLOCATED. IN THIS LAYOUT THEY ARE ALIGNED IN TWO ROWS. ALSO, IT IS POSSIBLE TO APPRECIATE THE BIG SIZE OF SUCH MACHINERIES FOR INDUSTRIAL APPLICATIONS AND THE EXTRA SPACE THAT THEY REQUIRE FOR OPERATION AND MAINTENANCE, I.E. THE CORRIDORS INSTALLED OVER THEIR TOPS.



In particular, power requirements are compared in Table 42, while the following dimensions are used for the fermenters (listed data refer to a single fermenter):

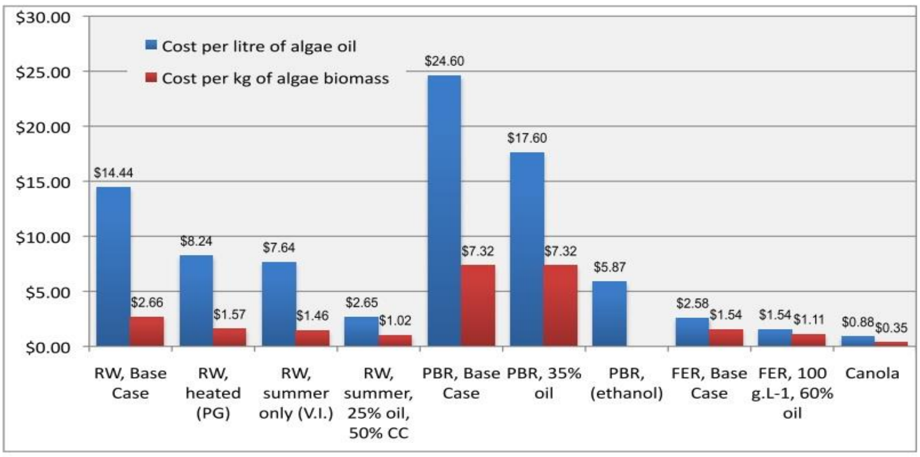
- Capacity: 100 m^3
- Height/Diameter ratio: 3 (height 10,5 m; diameter 3,5 m)
- Area occupied: 9,6 m^2

Actually when fermenters are installed in the facility, they require some extra space: if the internal diameter is 3,5 m, they have a layer 10 cm thick for insulation. Moreover, when installed, they lay nearby each other to limit the space required, but still a little passage (corridor) has to remain between two neighbouring fermenters to permit operations (see Figure 76). This space is just wide enough for the passage of one man and it is assumed to be 1 m. Therefore, the total area needed for one single fermenter is about 17 m^2.

COMPARISON BETWEEN SYSTEMS

To recapitulate, for autotrophic growth, open pond is a cheaper arrangement than closed PBRs, for large-scale algal biomass production. Open ponds also have lower energy requirement, and regular maintenance and cleaning operations are easier. This has been confirmed from the study of the British Columbia Innovation Council [26] which results are drawn in the graph of the Figure 77. From their work, it's clear that currently PBRs are less cost-effective for a commercial-scale development. In the Figure 77 the first four cases (RW) refer to a raceway system, whereas PBR indicates photobioreactor layout and FER is for fermenter. Three main considerations have to be done. The first is that in every cases evaluated, opens systems are cheaper than PBRs. Secondly, fermenters seem to be cheaper than opens systems in turn and they are also the only system that approaches the prices of conventional biofuels. Finally, currently for the hypothesis done in that report, biofuels from microalgae are more expensive in comparison with first generation biofuels. In this study, energy is supplied to the biorefinery without any additional cost. For this reason, it's rational to reckon that prices to obtain biofuels will be lower. Moreover, many cases of bankruptcy for PBR systems are reported in [191]. In addition, algae grown in photobioreactors consume nearly as much energy as is produced [141]. For these reasons, PBRs are not further evaluated in this study. Because of open ponds are already widely adopted and are the most well-established system for cultivation of microalgae in commercial-scale plants, here open ponds system (raceway) has been considered as the benchmark layout,

FIGURE 77. ECONOMICAL RESULTS REGARDING DIFFERENT CULTIVATION SYSTEMS [27].



whereas fermenters have been evaluated like an interesting alternative, thanks to their advantages. Actually Figure 77 and our results (see paragraph 4.1) confirm that fermenters gave a proof of a better performance. Indeed, they have a very low land requirements for the cultivation of algae and very high cell concentration as well, that results in a much higher yield per square meter per day. Land requirement is found to be an important limit for commercial scale applications.

Indeed to have a sufficient amount of energy to address to the biorefinery in order to work in load following mode, and to have a reasonable yield in matter of biofuels produced (and sold) per year, the size of the cultivation system in case of open ponds would require hundreds of hectares. On the contrary the same yields of algae harvested, and subsequently biofuels produced, require no more than 1 ha in case of fermenters. Therefore, the power requirements calculated for the scenario that involves fermenters better match the coupling with a NPP (see paragraph 2.4.2). All these considerations are deeply analysed in paragraph 4.1.

Hence for raceway ponds and fermenter tanks, the energy consumptions have been analysed (Table 42).

TABLE 41. A COMPARISON BETWEEN THE 3 DIFFERENT LAYOUTS FOR THE CULTIVATION PHASE OF THE BIOMASS.

ITEM	RACEWAY	PBR	FERMENTER
Cell density in culture	Low	Medium	High
Limiting factor for growth	Light	Light	Oxygen
Culture volume necessary to harvest a unit weight of cells	High	Medium	Low
Surface area-to-volume ratio	High	Very High	N/A
Cost of the system	Low	High	Low
Energy consumptions	Medium	High	Low
Control over parameters	Low	Medium	Very High
Commercial availability	Readily Available	Under custom built	Readily Available
Construction costs per unit volume produced	Medium	High	Low
Operating costs	Medium	High	Low
Technology base	Readily Available	Under development	Readily Available
Risk of contamination	High	Medium	Low
Evaporative water losses	High	High	Low
Weather dependence	High	Medium	Low
Maintenance	Easy	Difficult	Specialized
Susceptibility to overheating	Low	High	N/A
Susceptibility to excessive O_2 levels	Low	High	N/A
Ease of cleaning	Very Easy	Difficult	Difficult
Ease of Scale-up	High	Variable	High
Land requirement	High	Variable	Low
Applicability to different species	Low	High	Low
Possibility of scaling[21]	Low	Medium	Very High

[21] Possibility to expand the size of the plant, installing multiple single unit of production (say one pond or one fermenter). In case of open ponds, for large plants the number of installed units could be restricted by the whole land requirement (hundreds of hectares). On the contrary, it is easier expand the size in case of fermenters, due to their lower area occupied and easy reproducibility.

TABLE 42. DATE REPORTED IN LITERATURE FOR THE CULTIVATION PHASE IN MATTER OF POWER REQUIRED BY THE OPEN PONDS AND BY THE FERMENTERS. PBRS ARE NOT ANALYSED (SEE TEXT).

CULTIVATION LAYOUT	ITEM	SPECIF DATA IN REFERENCE (A)[22]		DATA ADAPTED TO THIS STUDY (B)[23]		POWER REQUIRED		REFERENCE
Raceway pond	water mixing	1,23	kWh/m^3	600000	m^3/d	30,75	MW	[22]
	CO_2 circulation	2,01	kWh/m^3	600000	m^3/d	50,25	MW	
	TOTAL CULTIVATION	*3,24*	*kWh/m^3*	*600000*	*m^3/d*	***81,00***	***MW***	
	CO2 circulation	363,3	kWh/ton	180	ton/d	2,72	MW	[33] Base case scenario, standard Nitrogen
	water mixing	64,7	kWh/ton	180	ton/d	0,49	MW	
	water pumping	71,2	kWh/ton	180	ton/d	0,53	MW	
	TOTAL CULTIVATION	*499,2*	*kWh/ton*	*180*	*ton/d*	***3,74***	***MW***	
	CO2 circulation	403,3	kWh/ton	180	ton/d	3,02	MW	[33] Base case scenario, starvation Nitrogen
	water mixing	84,5	kWh/ton	180	ton/d	0,63	MW	
	water pumping	92,8	kWh/ton	180	ton/d	0,70	MW	
	TOTAL CULTIVATION	*580,6*	*kWh/ton*	*180*	*ton/d*	***4,35***	***MW***	
	CO2 circulation	1171	MJ/ton	180	ton/d	2,44	MW	[33] Best case scenario
	water mixing	165	MJ/ton	180	ton/d	0,34	MW	
	water pumping	18,7	MJ/ton	180	ton/d	0,04	MW	
	TOTAL CULTIVATION	*1354,7*	*MJ/ton*	*180*	*ton/d*	***2,82***	***MW***	
	TOTAL CULTIVATION	*21890*	*kWh/ha*y*	*800*	*ha*	***2,22***	***MW***	[24]
	water mixing & CO2 circulation	*3,72*	*W/m^3*	*2400000*	*m^3*	***8,93***	***MW***	[174]
	TOTAL CULTIVATION	*8,9*	*kWh/L_oil*	*28500*	*L/d*	***10,57***	***MW***	[26]
Fermenter	Electricity	2,4	kWh/L_oil	28500	L/d	***2,85***	***MW***	[26]
	Heat	8,5	MJ/L_oil	28500	L/d	***1,40***	***MW***	
	Spark plug	0,375	kW/10000 L	30000	m^3	1,125	MW	[192]
	Stirring	1	kW/10000 L	30000	m^3	3	MW	
	TOTAL CULTIVATION (electric)	*1,375*	*kW/10000L*	30000	m^3	***4,13***	***MW***	
	TOTAL CULTIVATION (electric)	*3*	*kW/m^3*	30000	m^3	***90,00***	***MW***	[187]
	TOTAL CULTIVATION (electric)	*1 - 10*	*kW/m^3*	30000	m^3	***30-300***	***MW***	[193]
	TOTAL CULTIVATION (electric)	*28,8*	*kW/m^3*	30000	m^3	***864***	***MW***	[194]
	TOTAL CULTIVATION (electric)	*0,15*	*kW/m^3*	30000	m^3	***4,50***	***MW***	[195]
	TOTAL CULTIVATION (electric)	*0,5 - 16*	*kW/m^3*	30000	m^3	***15-480***	***MW***	[196]

[22] In this table “m^3“ for raceway refers to cubic meters of treated water, while for fermenters refer to the capacity of the plant, “ton” to metric tonnes of algae produced, “L_oil” to litres of oil produced

[23] To make a fair comparison between references, specific data found (column A) were multiplied for the values contained in this column (B). Column B (raceway ponds rows) contain values calculated in this study for a 800 ha raceway pond facility (calculation are detailed in paragraph 4.1). In literature authors usually refer to facilities of 100-400 ha, but it is also reported that companies are studying the feasibility of very large size projects by 2018: i.e. up to 880 for Sapphire Energy (San Diego, USA) and up to more than 1000 ha for Independence Bio Products – IBP (Texas, USA). To make a rough initial comparison, 800 ha facility has therefore been chosen for open ponds. Because of land requirements are much less demanding for fermenters, to have a reasonable confront of power requirements, we have calculated the yield of algae harvested per day in case of 800 ha facility. As it will be explained in the following (paragraph 0), the yield depends on many factors, but our calculations say that it is about 130-200 ton/day for every “open ponds” scenarios evaluated. Hence we have fixed the value of about 180 tons per day for fermenters case. It correspond to a facility of total volume of 30 million of m^3. This facility, in contrast with open ponds, could occupy only 0,5 ha. This table aims only to analyze power data for cultivation phase. For this aim, it has been necessary to use data introduced later in the thesis (par. 0 and par. 4.1).

Let's divide the analysis for open ponds and for fermenters. Comparing data found in literature, for a 800 ha raceway system, a good compromise could be a power consumption (electric) of about 10 MW. In particular, I preferred to exclude the usage of the value of 10,57 MW reported in [26], because it is referred to litre of oil produced. Therefore this number also involve some subsequent phase of the process, like the dewatering phase and the oil extraction. Both these processes are characterized by yields, efficiency and assumptions that strictly depend on the kind of process employed. Therefore this value is influenced by hypothesis not reported in that paper. For open ponds system are considered three main sources of energy requirements: water mixing (by puddle wheel) and CO_2 circulation (by auxiliary pumps and piped system), and pumping of inlet/outlet fluids (water and medium with algae). The values chosen are that marked in red in the Table 42: water pumping 71,2 kWh/ton of algae harvested [33], while water mixing and CO_2 circulation 3,72 W/m^3 of filling water in ponds [174].

On the other side, for fermenters data collected in literature are rather spread. Without considering some extreme cases like 4 or 800 MW, a fair optimistic conciliation could be represented by 30 MW. Therefore the value reported in [193] of 1 kW/m^3 installed will be used in our model.

B. Harvesting and Dewatering

Algae are grown in diluted liquid cultures, and achieving the low water contents required for subsequent extraction phase represents one of the greatest challenges for the production of algae-derived biofuels. A proper downstream processing is necessary to minimize energy requirements and optimize costs [141]. Indeed, in literature it has been pointed out that the dewatering of microalgae is one of the main bottlenecks in algal culturing [33]. For instance, in the model of [89] the energy required for the dewatering process accounted for 84,9% of the total energy consumption. In addition, recovery of the biomass from the broth has been claimed to contribute 20–30% to the total cost [197]. Recovery of biomass can be a significant problem because of the small size of the algal cells (paragraph 2.2). Culture broths are generally relatively diluted (< 0.5 kg/m^3 of dry biomass in open ponds systems) and hence large volumes need to be handled to recover the biomass [197].

Many are the techniques to harvest the microalgae and they are usually used simultaneously: harvesting of biomass requires generally more separation steps. The most interesting debate in literature is about which level of drying has to be reached to permit to biomass to pass at the oil extraction process. In other words, which is the goal of the dewatering process expressed in percentage of dry biomass. In particular, the main difference is between studies that consider a dewatering up to about 10-30% (of dry biomass in the sludge) and others that consider a dewatering up to 80-95%, achievable only thanks an energy intensive thermal drying. The former follows a "wet route" to biofuel, while the latter follows a "dry route".

In the wet route, the dewatering processes consume obviously a less quantity of energy, but the extraction process became more difficult and more energetically expensive (see Table 52). In the dry route, the thermal drying requires a considerable amount of heat, but this permits to have a more conventional and energetically cheap oil extraction process. The overall results indicate that based on current available technologies, the dry route has higher

FER[24]. [33]. This means that the dry route is more energetically convenient. Moreover, a dried biomass can be stored easily, requiring less space (the volume needed is low in comparison with a wet biomass) and for longer times; in contrast a wet biomass has to be processed immediately (in few hours maximum), otherwise it putrefies. Choosing a dry route gives more flexibility in operations and also supplies an insurance in case of unexpected unavailability of the plant. Finally

"According to [198], *algae oil extraction is very similar to soybean extraction. However soybean has a solid content around 90%. Hence to preserve consistency of the study, algal paste has to be dried up to a solid content of 90% before being processed in the oil mill. […]"* [89].

For these reasons, here a dry route has been followed, even if the wet route could be suitable in the future, especially if additional improvements will affect new techniques for wet extractions [141]. In the following of the paragraph, the main methods for dewatering will be briefly described, together to a summary of the techniques adopted by other authors and a critical analysis of methods, shown in Table 43.

Biomass can be harvested by centrifugation, filtration, floatation, gravity sedimentation or ultrasounds separation. These processes may be preceded by a flocculation step.

FLOCCULATION

Microalgae cell surfaces are negatively charged [199], the intensity of which, depends on the species, ionic strength of the medium, pH and other environmental conditions. In a stable culture, the electrical repulsion between the cells and the cell interactions with the surrounding water prevents the aggregation of the cells and contributes to the stability of the algal suspension. The neutralization of these surface charges destabilizes the algae culture, leading to the agglomeration of the algae into large clumps or "flocks", which can then be more readily separated from the culture medium. There are various methods of inducing flocculation in algal cultures, such chemical flocculation, electro-flocculation or bio-flocculation. Chemical flocculation is the most popular and has been used here, because it doesn't require any extra energy, apart pumping water: algae may be induced to flocculate by the addition of inorganic chemicals, such as aluminum sulfate ($Al_2(SO_4)_3$), ferric sulfate $Fe_2(SO_4)_3$, ferric chloride $FeCl_3$ or lime ($Ca(OH)_2$) [26].

SEDIMENTATION

Sedimentation has also been suggested to harvest microalgae cells based on the tendency of microalgae to settle when the energy is withdrawn and they are made quiescent. Sedimentation may or may not be coupled with flocculation. More often, sedimentation can also be used to harvest flocculated algae [200], where the larger particle sizes of the flocks increases the settling velocities of the algae.

ULTRASOUND SEPARATION

[201] described a method of harvesting microalgae by the use of ultrasound induced flocculation, using high frequency, low sound amplitude waves which is

[24] FER means Fossil Energy Ratio. It is often used as parameter to evaluate a Life-Cycle Assessment of a technology. It is calculated as a ratio between the energy extracted from products obtained and the total fossil energy used in a technology. Having a high FER means being capable to extract a great amount of energy from a particular source of energy, using a particular process.

then followed by sedimentation. The microalgae are pumped into a resonator chamber which consists of a transducer and a reflector. When the apparatus is turned on, it creates fields of maximum energy (bellies) and fields of minimum energy (nodes). The microalgae are driven into the node planes, where they aggregate and settle when the field is nullified. Ultrasound separation has the advantage of maintaining the viability of the microalgae as it causes no shear, avoids mechanical failures as it has no moving parts and enables continuous operation. These authors reported harvesting efficiencies of up to 92%. However, scale up of the system is difficult as the resonator chamber needs to be cooled. Moreover the energy costs for this method of harvesting are very high.

DISSOLVED AIR FLOATATION

This involves pressurizing some of the liquid to dissolve additional air. When the pressurized liquid is mixed with the algae culture at atmospheric pressure, the air comes out of solution as bubbles that attach to the flocks, making them float.

CENTRIFUGATION

This is a well-established industrial process that uses gravitational force to achieve separation. The morphology and sizes of the cells being harvested affect the recovery (and costs) as filamentous cells and large colonial cells will settle more readily than single smaller cells. Centrifugation is energy intensive, with estimates of the energy consumption required for various types of centrifuges estimated to range from 0,3 to 8 kWh/m^3. The high capital and running costs associated with centrifuges limit their use to second-stage filtration in the processing of microalgae for biofuels. [26]

FILTRATION

The principle of filtration is introducing the particles onto a screen of given aperture sizes. The particles either pass through or are retained on the screen according to their size. Filtration can be performed under pressure or vacuum with energy requirement estimates ranging from 0.2 to 0.88 kWh/m^3 and 0.1 to 5.9 kWh/m^3 respectively [197]. Filtration can also be carried out by micro-strainers which consist of a rotating drum covered by a straining fabric. A backwash spray collects the particles into an axial trough. Low power requirements of between 0.02 to 0.2 kWh/m^3 have been reported [202]. Although the costs associated with filtration are low, screen clogging and membrane fouling limit its suitability to larger species of algae. Filter presses operating under pressure and belt filters have been reported to operate satisfactorily for the recovery of microalgae and have been reported in use in commercial facilities [26].

COMPARISON BETWEEN PROCESSES

To make a comparison, the use of only filtration recovery may be unsatisfactory because filtration can be relatively slow. For extremely low value products, gravity sedimentation, possibly enhanced by flocculation, may be the method of choice. Sedimentation tanks or settling ponds are generally used in biomass recovery from sewage-based processes [203]. Centrifugal recovery of the biomass is feasible for high-value products. Centrifuges can process large volumes relatively rapidly and the biomass can remain fully contained during recovery [197].

To conserve energy and reduce costs, algae are often harvested in more steps process, as it is possible to understand from Table 43. In the first step, the algae are concentrated, often by flocculation, which concentrates the diluted cultures to about 1-5% w/w. In the second step, the cells are further concentrated by

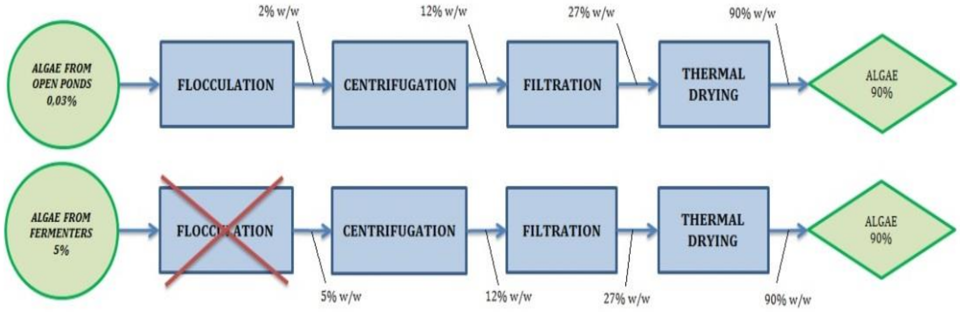


FIGURE 78. STEPS SELECTED IN OUR MODEL FOR HARVESTING AND DEWATERING PHASE. BETWEEN THE ROUTE FOR OPEN PONDS SYSTEM AND THAT ONE FOR FERMENTERS THERE IS ONLY ONE DIFFERENCE: THE CONCENTRATION OF ALGAE WITHIN THE MEDIUM OF FERMENTERS IS HIGHER IN COMPARISON TO THAT ONE WITHIN THE BROTH IN OPEN PONDS. FOR THIS REASON, THE FLOCCULATION IS NOT NECESSARY IN CASE OF FERMENTERS.

centrifugation, to get a solids concentration of 12-25%. Then the biomass has to reach the concentration of 80-90% by a thermal drying. Actually the water to be removed is not only that is contained in the broth, but also that is present in the microorganisms. Thermal drying is inevitably required due to bound intercellular water. Typical algae cells can contain between 40 and 80 % water [204]. This water cannot be removed by mechanical operations such as centrifugation or filtration. A thermal dryer can achieve the low moisture contents needed [141]. The thermal drying is a very energy intensive process. For this reason, it is suggested to reduce the amount of water before beginning the final drying trough a mechanical dewatering process. As studied by [33], this permits to reduce drastically the global energy requirements of the overall dewatering phase (see Figure 79). In the latter work, two factors are underlined. Firstly the thermal drying is the most energy intensive process, being accountable of 70-99,9% of the whole energy consumptions in the dewatering phase. Secondly, adopting more than one technology before the thermal drying permits to reduce the total amount of energy required from the process. As it is possible to see from Table 43, the most common method is to use a chamber filter press between the centrifugation step and the thermal drying. In Figure 78 is anticipated what is chosen in this work in matter of route for harvesting and dewatering phase.

Therefore in this study the following dewatering steps have been evaluated:

- Flocculation (to 2%)
- Centrifugation (to 12%)
- Filtration (to 27%)
- Thermal drying (to 90%)

They are the methods selected for the open ponds because the algal slurry is very dilute (only 0,03% w/w of dried alga). Since the concentration of algae is higher in case of cultivation in fermenters (5% w/w), for that scenario the flocculation is not needed. However the rest of the processes is supposed being identical.

The final moisture level of 10% is optimal for grinding the biomass later in the process, and ensures that, under conditions of extended storage, the microalgae will not deteriorate appreciably. However, these are only the processes selected.

TABLE 43. A COMPARISON BETWEEN TECHNOLOGIES APPLIED IN DIFFERENT STUDIES FOR THE DEWATERING PHASE. THIS IS A VERY CRUCIAL PHASE, AND BECAUSE OF THE MODERNITY OF THIS BIOMASS CONCERNED TO THE PRODUCTION OF BIOFUELS, A PREDOMINANT ROUTE IS STILL NOT ESTABLISHED.

DEWATERING PROCESS	[33] – **base case**		[33] – **best case**		[89]		[9]		[9]		[22]		[26]		[26]		[26]		[197]		[141] – **case 3**		[141] – **case 3**		[141] – **case 5**		**AVERAGE**
	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	Y/N	RESULT	
chemical flocculation	✓	to 2%	✓	to 2%	✓	to 2%					✓	-	✓	to 2%	✓	to 2%	✓	to 2%	✓	to 2%	✓	to 5%	✓	to 5%	✓	to 5%	2,6%
disk stack centrifuge	✓	to 16%	✓	to 16%			✓	-			✓	-	✓	to 20%					✓	to 12%	✓	to 12%					15%
spiral plate centrifuge																							✓	to 31,5%			31,5%
chamber filter press	✓	to 30%	✓	to 50%	✓	to 20%			✓	-									✓	to 27%	✓	to 27%			✓	to 27%	30,8%
heat assisted rotary pressure filter																							✓	to 56%	✓	to 56%	56%
thermal drying	✓	to 85%	✓	to 85%	✓	to 90%	✓	to 91%	✓	to 91%											✓	to 95%	✓	to 95%	✓	to 95%	90,3%
floatation															✓	to 8%											8%
sedimentation																	✓	to 5%									5%

As just discussed, many different machineries exist to complete a single step. In Table 46 a confront is done in matter of power requirements, while in Table 47 a confront is done in matter of efficiency of the processes. It means that during the dewatering steps, a part of the biomass is lost.

Initial dewatering is performed by a flocculation unit. It is assumed to add Aluminum sulfate at 250 mg/L and lime at 0.73 g/g aluminum sulfate [205]. However the only energy consumption is required to pump water. In this study, the energy required for pumping water outlet from cultivation system (and therefore inlet flocculation step) is already inserted into the cultivation phase. Therefore for this study, no supplementary energy is accounted for flocculation process.

For centrifugation step, three different types of centrifuges were considered. The first was a disk-stack centrifuge, capable of removing water to approximately 12% dry algae content. This centrifuge has a processing capacity of 85 m^3/h and a power consumption of 45 kW. The second centrifuge used was a decanter bowl centrifuge. The decanter bowl centrifuge produces a 22% dry algae, is available at commercial capacities, and consumes 8 kWh/kg of water removed. The third centrifuge was a novel spiral plate centrifuge developed specifically for dewatering algae. It is capable of achieving 31.5 % dry algae weight, consumes 0.95 kWh/m^3 of algae slurry, and can process up to 40 m^3 of slurry/h [141].

Then three filtration methods were investigated for the mechanical dewatering step: a tangential flow filtration, chamber filter press, and a heat assisted rotary pressure filter. Although the tangential flow filtration has an energy consumption of only 0.00206 kWh/kg of water removed and it can achieve a final dry algae content of 8.8%. For this reason is not applicable after a centrifugation process. The chamber filter press consumes 0.88 kWh/kg of water removed and can achieve 27% dry algae content. The heat assisted rotary filter increases the solids concentration from 33 to 56% while using 60 kWh/dry ton of biomass. This filter was capable of operating at a capacity of 200 t of sludge/h. These filtration units require electricity as energy input [141].

Finally the last dying step is the most delicate, because it consumes the biggest quantity of energy. A steam rotary and a heat integrated dryer were considered as

FIGURE 79. IMPORTANCE OF MULTISTEP DEWATERING PHASE. THE ROUTE FOLLOWED IN THIS STUDY IS VERY SIMILAR TO THE CASE III SUMMARIZED IN THIS TABLE [34].

Different dewatering processes and their energy consumption for processing 1 ton of microalgae.

Dewatering step	Dry solid weight percentage	Distribution of the energy consumption for dewatering			
		Case I	Case II	Case III[d]	Case IV[d]
0 Before processing	0.5 wt.%	-	-	-	-
1 Lime flocculation[a]	2 wt.%	0.1%	0.8%	1.6%	3.4%
2 Disk stack centrifuge[b]	16 wt.%	-[f]	1.7%	3.5%	7.6%
3 Rotary pressure Filter[c]	30 wt.% or 50 wt.%[d]	-[f]	-[f]	7.9%	17.1%
4 The Delta Dryer[e]	85 wt.%	99.9%	97.6%	87.0%	71.9%
Energy consumption for dewatering ($E_{dewatering}$, GJ)		97.7	10.40	4.96	2.29
Total energy consumption of the dry route (E_{total}, GJ)		108.28	20.96	15.6	12.85
$E_{dewatering}/HHV_{microalgae}$		5.3/1	6/10	1/4	1/8
$E_{dewatering}/E_{total}$		90.2%	49.6%	31.7%	17.7%

[a] It is assumed that an adjustment of pH of culture to 11 by lime can achieve a floc concentration of 20 g L^{-1}.
[b] Based on the operation of a Alfa Laval PX-115 high-capacity (85 m^3 h^{-1}, 45 kW) disc stack centrifuge from Alfa Laval Tumba AB (Alfa Laval, 2010).
[c] Assumed to be 60 kwh per dry ton of microalgae based on a pilot scale industrial sludge impulse dewatering press (Mahmood et al., 1998).
[d] It is assumed that the dry solid weight percentage after the rotary pressure filter step are 30 and 50 wt.% for case III and case IV, respectively.
[e] 2 MJ per kilogram of evaporated water (van Gemert, 2009).
[f] This unit is not used in this case.

alternative methods of thermal drying. The values reported in literature for energy consumption of these driers range from about 4 to 2 MJ/kg of water removed, as reported in Table 48. In particular, the heat integrated dryer, developed by Delft University for drying a biomass- type sludge, consumes only 2 MJ/kg of water removed [206]. This dryer uses hot balls to contact the algae slurry under a vacuum, and condenses the water vapour over the metal balls to recover the heat. Although this is a new and modern method of drying, the value reported has been considered too optimistic in this study, and a more conservative value of 2,26 MJ/kg, equal to the latent heat of vaporization of water at ambient pressure, has been chosen here. Converting the value, it correspond to 628 kWh/m^3 of water evaporated. All the concepts expressed in this paragraph and results used for developing the model are summarized in the next two tables, Table 44 and Table 45.

TABLE 44. THE MAIN FEATURES OF THE DEWATERING PHASES CHOSEN FOR THE MODEL.

DEWATERING PHASE	INITIAL WATER CONTENT	FINAL WATER CONTENT	POWER CONSUMPTION
Flocculation	Very High	High	Very Low
Centrifugation	High	Medium	Medium
Filtration	High	Medium	Medium
Thermal drying	Medium	Very Low	Very High

TABLE 45. THE VALUES USED IN THE MODEL FOR THE DEWATERING STEPS.

DEWATERING PHASE	EFFICIENCY	POWER CONSUMPTION[25]
Flocculation	91%	0 (see text)
Centrifugation	90%	1 kWh/m3
Filtration	90%	0,88 kWh/m3
Thermal drying	95%	628kWh/m3

C. Oil Extraction

After the preparation of the dried microalgae, the next step is the extraction of the main bricks that form the biomass. This means to isolate the lipids and the carbohydrates to direct to the different chemical processes for the production of biodiesel and bioethanol, respectively. Hence in this paragraph the extraction of oil will be debated. The main product of this phase is obviously the crude oil that has to be refined before the transesterification can begin. On the other hand, the "waste" of this process is actually a worthwhile co-product, called in the following "alga cake", rich in carbohydrates, that is sent to the fermentation process to obtain bioethanol. In this section the main methods to extract oil will be briefly introduced and compared, while the selected method for this study will be deeply analysed, also reporting their energy requirements.

[25] For centrifugation and filtration, data refer to volume of processed water (then water inlet). For thermal drying, data refer to volume of water evaporated

DEWATERING PROCESS	[33] – **base case**	[33] – **base case**[26]	[89]	[22]	[26]	[197]	[141]	**THIS STUDY**
flocculation	21,7 kWh/ton	Not interesting	-	4,11 kWh/m^3	0 - 0,3 kWh/m^3	-	-	0 (power for pumping water is counted in cultivation phase)
disk stack centrifuge	26,5 kWh/ton	0,54 kWh/m^3[27]	-		0,3 - 8 kWh/m^3	1 kWh/m^3	-	1 kWh/m^3 (of water inlet)
filtration (pressure)	60 kWh/ton	11,43 kWh/m^3[28]	-	-	0,2 - 0,88 kWh/m^3	0,88 kWh/m^3	0,88 kWh/m^3	0,88 kWh/m^3 (of water inlet)

DEWATERING PROCESS	[13]	[9]	[89]	[26]	[197]	[141]	**THIS STUDY**
flocculation	> 90% (90-98%)	-	90%	70-96% (av. 83%)	95%	95%	91%
disk stack centrifuge	-	-	-	90%	60-95%	-	90%
filtration (pressure)	-	90%	-	-	-	-	90%
thermal drying	-	100%	-	-	-	-	95%[29]

ENERGY REQUIREMENT	REFERENCE
3,3-3,9 MJ/kg	[207]
3 MJ/kg	[208]
2 MJ/kg	[206] (p=vacuum)
3,556 MJ/kg	[9]
3,54 MJ/kg	[89]
2,5 MJ/kg	[141]
2 MJ/kg	[33]
2,26 MJ/kg	latent heat of evaporation (p=1 bar)

TABLE 46. AT THE TOP OF THE PAGE. ENERGY NEEDED BY THE PROCESSES SELECTED FOR THE DEWATERING PHASE.

TABLE 47. IN THE MIDDLE OF THE PAGE. EFFICIENCY OF THE PROCESSES SELECTED FOR THE DEWATERING PHASE.

TABLE 48. ON THE LEFT. ENERGY REQUIRED FOR THE THERMAL DRYING PER UNIT OF WATER EVAPORATED.

26 Data of the column on the left were converted in this column in kWh/m3 to make a fair comparison with other references. Method of conversion is found in Appendix B.

27 Data for centrifugation step in similar with other data of other references, because the same centrifuge (disk stack c.) has been used

28 Data for filtration step here is higher in comparison with other authors because a rotary pressure filter has been used in [34]. This filter consumes more energy in comparison with a chamber filter press, as discussed in the text.

29 Conservative hypothesis

There are a few well-documented procedures for extracting oil from microalgae, those being mechanical pressing, homogenization, milling, solvent extraction, subcritical or supercritical fluid extraction, enzymatic extractions, ultrasonic-assisted extraction and osmotic shock [105]. All these methods have their individual benefits and drawbacks. In Table 49 advantages and limits of the main technologies are listed. Pressing and homogenization essentially involve using pressures to rupture cell walls, in order to recover the oil stored within the cells. Milling on the other hand, uses grinding media (consisting of small beads) and agitation to disrupt cells. These methods are usually used in combination with some kind of solvent extractions. Solvent extraction entails extracting oil from microalgae by repeated washing or percolation with an organic solvent. Hexane is a popular choice due to its relatively low cost and high extraction efficiency. Alternatively, solvent extraction can be enhanced by using organic solvents at temperatures and pressures above their boiling point; this is called accelerated solvent extraction (ASE) [105]. The solvents used are those normally proposed in typical distillation based methods such as Soxhlet extraction. In application a solid sample is enclosed in a sample cartridge that is filled with an extraction fluid that statically extracts the sample under elevated temperature (50–200°C) and pressure (35–210 bar) for short time periods (5–10 min). Subcritical water extraction is based on the use of water, at temperatures just below the critical temperature, and pressure high enough to keep the liquid state[30]. The basic premise to subcritical water extraction is that water, under these conditions, becomes less polar (i.e., the dielectric constant is altered as the temperature is increased) and organic compounds are more soluble than at room temperature. There is also the benefit that as the water is cooled back down to room temperature, its dielectric constant changes such that it is no longer miscible with the extracted lipids and therefore separates easily. Supercritical fluid extraction is a relatively recent extraction technique that involves the use of substances that have properties of both liquids and gases (i.e. CO_2) when exposed to increased temperatures and pressures. This property allows them to act as an extracting solvent, leaving no residues behind when the system is brought back to atmospheric pressure and room temperature. The majority of applications have used carbon dioxide because of its preferred critical properties (i.e., moderate critical temperature of 31.1°C and pressure of 72.8 atm), low toxicity, and chemical inertness, but other fluids used have included ethane, methanol, ethane, nitrous oxide, sulfur hexafluoride as well as n-butane and pentane.

Super critical extraction is often employed in batch mode, but the process can also be operated continuously. Lipids have been selectively extracted from microalgae at temperatures between 40 to 50°C and pressures of 241 to 379 bar [209]. Enzymes can also be used to facilitate the hydrolysis of cell walls to release oil into a suitable solvent. The use of enzymes alone, or in combination with a physical disruption method such as sonnication, has the potential to make extractions faster and with higher yields. The use of sonnication alone can also enhance the extraction process immensely due to a process called cavitation. Ultrasonic waves create bubbles in the solvent, the bubbles burst near the cell walls of microalgae, which produce shock waves, causing the contents (i.e. lipids) to be released into the solvent [210]. Osmotic shock, a less-employed procedure, makes use of an abrupt lowering of osmotic pressure that causes cells to burst and release their contents [105].

[30] Critical thermo-dynamical conditions of water are: temperature of 373.9°C and pressure of 22.059 MPa.

TABLE 49. COMPARISON OF ADVANTAGES AND LIMITATIONS BETWEEN THE MAIN METHODS TO EXTRACT OIL FROM MICROALGAE.

EXTRACTION METHOD	ADVANTAGES	DISADVANTAGES	REFERENCES
Pressing	Easy to use No solvent involved	Large amount of sample required Slow process Low efficiency with single cell microorganism	[211] [28]
Solvent extraction	Solvent used are usually inexpensive Results are reproducible	Most organic solvents are highly flammable and/or toxic Solvent recovery is expensive and energy intensive Large volume of solvent is required	[212] [213]
Subcritical water extraction	Shorter extract times Higher quality of extract Lower costs associated with the extracting agent Environmental compatibility It eliminates the need for the dewatering step	Require a significant cooling system to rapidly cool the product down to room temperature in order to avoid product degradation High power consumption Expensive/difficult to scale up at this time Extreme thermo dynamical conditions	[214] [28]
Supercritical fluid extraction	Non-toxic (no organic solvent residue in extracts) 'green solvent' Non-flammable and simple operation Solvent and product are easily separated downstream	High power consumption Expensive/difficult to scale up at this time	[215] [216]
Ultrasonic assisted	Reduced extraction time Reduced solvent consumption Greater penetration of solvent into cellular materials Improved release of cell contents into bulk medium	High power consumption Difficult to scale up	[217] [218]

The application of mechanical pressing alone is limited because microalgae, as single cell microorganisms, some of which contain rigid cells walls, will not be crushed but will rather flow with the water through the thousands of water micro-channels that exist in pressing equipment [28]. In addition, methods like sub- and supercritical fluid extraction, and ultrasonic extraction involve extreme condition of temperature and pressure, and high power consumptions, and usually suffer of issues of scalability. For these reasons in this study the solvent extraction method has been selected, that is also the most used and therefore that one for which more data are available.

Solvent extraction of oil from biomass is a process whereby the target analyte is transferred from one phase (e.g., a solid phase in the case of dried biomass and an aqueous liquid phase in the case of wet biomass) to a second immiscible phase (e.g., an alcohol or an alkyl halide). In other words, the analyte (i.e., lipid) molecule must dissolve into the solvent and form a solution. The solubility of the analyte in the solvent is governed by the Gibbs free energy of the dissolution process, showed in equation (10) which is directly related to the equilibrium constant governing the concentration of the analyte in either phase.

$$\Delta G° = -RT \ln \frac{[analyte]^{solvent\ phase} \cdot [solvent]^{solvent\ phase}}{[analyte]^{analyte\ phase} \cdot [solvent]^{analyte\ phase}} = \Delta H° - T\Delta S° \qquad (10)$$

As more of the analyte dissolves into the solvent phase, the natural logarithm of the quotient becomes positive and the Gibbs free energy for this reaction becomes negative, indicating that the reaction has proceeded more favourably in the direction of the analyte dissolving into the solvent. As the analyte fully dissolves into the solvent phase, the quotient approaches infinity and the equilibrium lies totally to the right, and the target analyte (i.e., lipid) is considered fully extracted into the solvent phase. The solubility of the analyte in various solvents is governed by two independent parameters (which may, or may not, work together): the enthalpy and the entropy of mixing. The solubilisation of the analyte in the solvent is therefore favoured when the dissolution process gives off energy (i.e., $\Delta H°<0$) and/or when the dissolution process increases entropy (i.e., $\Delta S°>0$). How the analyte molecule chemically interacts with the selected solvent will dictate whether the change in enthalpy is positive or negative, whether the change in entropy is positive or negative, and whether their combined sum yields a favourable Gibbs free energy of dissolution. The overall sum of these two terms is defined by the total relative contribution of all intermolecular forces that occur between the analyte and solvent molecules: Electrostatic, London forces, hydrogen bonds, and hydrophobic bonding [219]. Consequently, the development of any solvent based extraction process must comprise a choice of solvent that yields a set of chemical interactions between the analyte and solvent molecules that is more favourable than the chemical interactions both between the solvent molecules themselves (i.e., self-association), and between the analyte with the matrix it was already associated with. As a general rule analytes that strongly self-associate dissolve best in strongly associated solvents, while analytes that weakly associate dissolve best in weakly associated solvents. In other words, polar solutes will dissolve in similarly polar solvents and non-polar solutes will dissolve better in similarly non-polar solvents (i.e., ''like dissolves like'').

Organic solvents, such as benzene, cyclohexane, hexane, acetone and chloroform have shown to be effective when used on microalgae paste. A suitable solvent should be insoluble in water, preferentially solubilise the compound of interest, have a low boiling point to facilitate its removal after extraction, and have a considerably different density than water. Also, for process cost-effectiveness, it should be easily sourced, as well as inexpensive and reusable [220]. Due to these qualities, hexane is typically the solvent of choice for large scale extractions [105]. Moreover, an efficient extraction requires that the solvent be able to fully penetrate the biomass matrix in order to contact the target analytes wherever they are stored, and that the solvent's polarity must match that of the target analyte(s) (i.e. non-polar solvent such as hexane for extracting non-polar lipids) [221]. The development of any extraction process should account for the fact that the tissue structure may present formidable barriers to solvent access [222], [223]. In fact, one way to enhance the extraction is to mechanically disrupt the native structure of the biomass prior to the employment of an extraction solvent, in order to favour the continuous penetration of persistent membrane enclosed regions. Therefore, cell disruption is often necessary for recovering intracellular products from microalgae prior to exposing them to the solvent [197]. A mechanical cell disruption method that has been tested at both small and large scales is simply milling with a pestle and mortar; it is also the most used because it guarantees the best recovery of lipids [28] and is more easy to apply.

Since the hexane extraction method is used both for soybean feedstock and for microalgae, in this study the scheme of the extraction phase (and for the subsequent biodiesel) have been taken from the soybean oil separation and biodiesel conversion processes elaborated in [224]. Consequently, also final hexane-extracted algal oil is assumed to have similar composition as soybean oil [9]. The extraction phase is a multi-steps process, summarized in the schema of Figure 80. The first step is the milling of the biomass. The powder obtained is then sent to the extractors: this permit to have an alga cake, sent to the respective processing, and a mixture of oil and hexane. This mixture is sent to the step of oil recovery, after which hexane is separated (and sent back to extractors) from the crude oil. Before using this oil in the transesterification process to obtain biodiesel, a well-designed pre-treatment process is imperative for an efficient and profitable biodiesel plant. In some cases, a simple pre-clarification step using a decanter to remove insoluble impurities is all is required. But in most cases pre-clarification alone is insufficient. Typically, two main compounds need to be removed: gums and free fatty acids (FFA). Two different processing routes can be used to perform these tasks. These are called chemical refining and physical refining. The choice of route depends on the type and variety of raw materials as well as the capital and operating costs. Chemical refining first conditions the oil with acid to prepare gums for removal. Next, FFA are neutralized with caustic soda. The resulting mixture is sent to a disc stack centrifuge where the impurities are removed and the result is an oil virtually free of FFA and gums. The oil is further cleaned by mixing it with water and then performing another separation with a disc stack centrifuge. Optionally, an adsorption step with silica and/or bleaching earth can be used to complete the final step. Physical refining begins with a degumming stage. A variety of degumming processes exist, some to remove only hydratable gums and some to remove non-hydratable gums. Degumming for pre-treatment before biodiesel production removes both types of gums to the minimum level. After degumming, washing or adsorption is used to further lower gum content. Finally, de-acidification, a physical, steam stripping process, is used to remove the FFA [225]. In this study the pretreatment processes chosen are that suggested in [224]. The degumming step has been incorporated in the oil extraction process, while the removing of FFA will be described as the first step in the next section. In matter of data and power requirements a comparison has been developed between different papers. The lipid percentage of algal feedstock and working volume in the separation process were adjusted for algal feedstock obtained in the previous processes.

MILLING

Some of the steps such as hull cracking, conditioning, flaking, and other pre-extraction processing steps specific to soybean feedstock were omitted due to the differences between algal biomass and soybeans. Therefore the only energy requirement for the first step is the pure grinding of the biomass.

OIL EXTRACTION

The heart of this part of the process is the extractor. Four types of extractors are marketed in the industry: rotary or deep-bed, horizontal belt, continuous loop extractors, and a miscellaneous collection of designs. The extractor used in this model is known as a "stationary basket extractor". It falls into the rotary bed design category. Biomass is dropped into a series of baskets. The blend (the hexane/oil mixture) and hexane are pumped to the baskets to achieve a counter current scheme. It provides a counter current extraction scheme by rotating the solvent and solvent/oil mixture around the series of baskets. Each basket is washed by successively less concentrated oil/solvent mixtures, until each is ultimately washed with fresh solvent. The full mixture leaving the extractor can contain 19% to 24% oil. The full mixture is sent through a liquid cyclone that uses fresh hexane to remove entrained fines so that a clear blend can then be sent on to the recovery section of the plant. The yield of triglycerides is the same as for total oil. The rate of solvent addition is assumed to be 1.2 kg of solvent for every kg of algal powder. Most of this solvent is recovered and recycled, so that actual make-up hexane usage is only 0.0024 kg/kg of biomass.

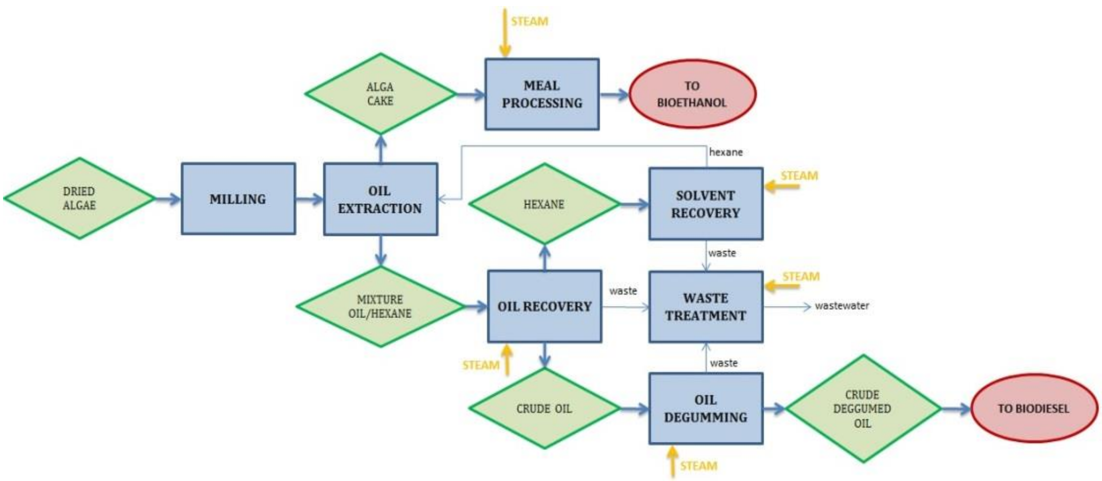
MEAL PROCESSING

Wet solvent-containing alga cake is reported to have a hexane content of 35% by weight. This is sent to a desolventizer-toaster, which does exactly what its name implies. This piece of equipment is designed to remove hexane by contacting the cake with open steam. The cake is toasted by contact with a series of high-temperature trays heated indirectly with high-pressure steam. Careful control of time, temperature, and moisture content in this unit is required to produce a meal essentially free of hexane and to inactivate urease and trypsin inhibitor enzymes present in the meal. Open steam is used to raise the moisture content to 20% on one of the trays; but final moisture content leaving the unit is 18%. Hexane concentration leaving the desolventizer-toaster is 995 ppm. The meal dryer uses indirect steam to reduce moisture content to a level of 14% and hexane to a level of 500 ppm. The meal is then cooled with air. Moisture content is further reduced at this point to 12%. Residual hexane is 400 ppm. The meal is then ground and conveyed to final storage and shipment. This part of the plant is the most energy intensive. Heating requirements for solvent removal and drying of the meal are the main contributors. All heat for this part of the plant is supplied by steam (both directly and indirectly through heat exchangers).

OIL RECOVERY

Multiple effect evaporators are used to concentrate the oil in the mixture exiting the counter current extractor. No oil losses occur in this section of the facility. It is in this part of the facility that some of the heat integration opportunities can be recognized. For example, hexane vapour coming from the first desolventizer-toaster in the meal processing section provides all the heat to the first stage evaporator. Most heat required for the extractor is, in turn, provided directly from the hexane vapour exiting the second stage evaporator.

FIGURE 80. SCHEMA OF OIL EXTRACTION PHASE.



Steam used in the second stage evaporator and the final oil stripper are the original sources of heat that drive the extraction process. Because of environmental and economic reasons, hexane is recovered extensively from vents throughout the facility. Most of these come together at the excess vapor condenser. Direct injection of steam reduces hexane content of the soybean oil to 120 ppm.

SOLVENT RECOVERY

Hexane and hexane/water condensate from the excess vapour condenser and other parts of the oil recovery system are sent to a settling tank. Solvent is continuously drawn off the top phase and pumped back to extraction section of the plant. The water phase is pumped to the waste treatment section of the facility. A small amount of hexane is pumped to the scrubber above the desolventizer-toaster, where it is used to remove entrained fines.

OIL DEGUMMING

The triglycerides are the component of the oil that can be transesterified to biodiesel. Free fatty acids are hydrolysed forms of fats that are not converted to biodiesel by transesterification and that must be removed before biodiesel production. Phosphatides in the biomass are more commonly referred to as “gums”. In our modelling of oil recovery and refining, we assume that these gums must also be removed before shipment to a biodiesel section [224]. Degumming removes phosphatides and some of the unsaponifiable matter from the crude oil. This is done simply by mixing the oil with hot water. As the gums are hydrated they swell and can be separated from the oil by the difference in density. Water is added at a rate of 75% of the level of phosphatides present in the oil. A centrifuge is used to separate the hydrated gums from the oil. The oil is vacuum dried and sent to storage for shipment. The gums collected in the aqueous phase are separated from the water and sent to meal processing, where they are added to the desolventizer-toaster. Oil losses in this section of the plant are 3.11%; but only 0.5% loss of triglycerides occurs. The final yield of crude, degummed oil is 92.5% of the oil contained in the processed algal biomass. The yield of triglycerides is 95%. Final composition of the crude, degummed oil is shown in Table 50.

WASTE TREATMENT

Waste treatment is focused primarily on recovery of residual hexane. To this end, a steam stripper is used to remove evaporated hexane that is then recovered in the excess vapour condenser in the solvent recovery section of the plant. The wastewater from the evaporator contains low levels of oil from the degumming operation. The vapour that remains from the excess vapour condenser is sent to an absorber to capture any residual hexane before venting. Hexane is recovered from the absorbent by steam stripping.

In Table 52 are listed the energy requirements and the efficiencies considered to make a comparison. In this case efficiency means the percentage of initial lipids that is successfully extracted. In particular, data reported in [224] for extraction of oil from soybean feedstock are close to data reported for extraction from microalgae, as supposed at the beginning of this section. Therefore, thanks to the fact that in that report more details are given about the usage of this energy and, in addition, that the reference reports less amount of electric energy and a higher heat requirement (more useful in this study), these data are used in this study. Furthermore, an efficiency of 92,5% has been used in the model developed here.

In Table 51, energy requirements for the extraction of oil are listed with regard to every step of this phase. In particular the heat requirement for the solvent recovery step has been calculated considering that the heat must supply power to evaporate the hexane. It is calculated assuming that 1,2 kg of hexane are added in the process per each kilogram of biomass, and using the hexane latent heat of 0,365 MJ/kg.

TABLE 50. COMPOSITION OF THE DEGUMMED OIL, SENT FOR THE NEXT ELABORATION PHASE. ONLY THE TRIGLYCERIDES CAN BE CONVERTED IN BIODIESEL.

Triglycerides	97.43%
Unsaponified Matter	1.5%
Free Fatty Acids	0.75%
Other	0.3%
Phosphatides (Gums)	0.02%

TABLE 51. ENERGY REQUIREMENTS FOR EXTRACTION PHASE, DIVIDED FOR EVERY STEP.

EXTRACTION STEP	ELECTRIC ENERGY [kWh/ton]	HEAT [kWh/ton]
grinding	4,14	-
oil extraction	3,6	-
meal processing	14,56	109
solvent recovery	0,52	122
oil recovery	0,38	24,2
oil degumming	1,69	19,02
waste treatment	0,57	10
TOTAL	**25,46**	**284,22**

D. Biodiesel Production

Biodiesel is a nonpetroleum-based fuel that generally consists of fatty acid methyl esters (FAME) or fatty acid ethyl esters (FAEE), derived from the transesterification of triglycerides (TG) with methanol or ethanol, respectively [226]. Methyl, rather than ethyl, ester production was used because methyl esters are the predominant product of commerce, because methanol is considerably cheaper than ethanol, and due to the greater ease of downstream recovery of unreacted alcohol. Also this permits to sell the whole quantity of ethanol produced in the facility here modelled. The composition of biodiesel is very similar to that one of conventional petroleum diesel, although the lower heating value (LHV) of the former is slightly lower: about 37,5 MJ/kg of biodiesel [227]. The transesterification reaction is reported in Figure 81: as it possible to see, glycerol is an important co-product of the process.

The model of biodiesel plant is broken out into six major processing sections. Crude oil is processed through a (1) caustic refining step to remove free fatty acids. The soaps generated in this step are removed by washing the oil with hot water. This wash is sent to (2) wastewater treatment. Before sending the oil to the (3) transesterification reactor, the oil is dried to remove water, which can be detrimental to yield in the reactors.

TABLE 52. ENERGY REQUIREMENTS AND EFFICIENCIES REPORTED IN LITERATURE FOR THE OIL EXTRACTION PROCESS.

EXTRACTION PROCESS	ELECTRIC POWER	HEAT	EFFICIENCY	REFERENCE
solvent	-	-	90%	[26]
solvent (hexane)	66,7 kWh/ton	211,1 kWh/ton	95%	[33]
solvent (hexane)	54,82 kWh/ton	259,5 kWh/ton	70%	[89]
solvent (butanol)	-	-	90%	[27]
solvent (hexane)	-	-	95%	[141]
solvent (hexane)	29,72 kWh/ton	201,9 kWh/ton	-	[114]
solvent (hexane)	-	-	92,5%	[9]
solvent (hexane)	25,46 kWh/ton	284,22 kWh/ton	92,5%	[224]
wet extraction	97,2 kWh/ton	486,1 kWh/ton	-	[228]
solvent (hexane)	**25,46 kWh/ton**	**284,22 kWh/ton**	**92,5%**	**This Study**

In this case efficiency means the percentage of initial lipids that is successfully extracted in the process from the biomass. In particular, two consideration have to be pointed out: firstly, in the last reference is underlined like in case of wet extraction the energy requirements are sensibly higher in comparison with normal extraction (as discussed in the previous paragraph); secondly, data reported in [224] for extraction of oil from soybean feedstock are close to data reported for extraction from microalgae, as supposed in the text. Therefore, thanks to the fact that in this report more details are given about the usage of this energy and, in addition, that the reference reports less amount of electric energy and higher heat requirement (more useful in this study), these data are used in this study. Also the efficiency of 92,5% has been used in this study. The energy requirements are reported per metric tons of alga coming from the dewatering phase.

Dry, caustic-refined oil is combined with a 2:1 stoichiometric excess of methanol and a small amount of catalyst. The reaction produces methyl ester and glycerine. Unreacted methanol is (4) recovered as extensively as possible and recycled into the reactors. A crude glycerine product (grade of about 80% glycerol) is sold as a by-product. The plant design does not include a full purification to USP-grade glycerine (about 98%), but only a (5) partial purification process. Before leaving the biorefinery the (6) methyl ester is washed to remove the residual quantities of glycerine and methanol.

OIL REFINING

The free fatty acids present in the oil are detrimental to the chemistry of transesterification because they can tie up catalyst through the formation of soaps, as shown in the reaction below.

RCOOH	+	CH_3ONa	--->	RCOONa	+	CH_3OH
(Free Fatty Acid)		(Sodium Methoxide Catalyst)		(Soap)		(Methanol)

In addition, soap formation can result in a more difficult phase separation of methyl ester and glycerol. To prevent this, caustic and water are added to the degummed algal oil before carrying out the transesterification step in a process known as alkali refining. The conditions for this process step are based on the well-established practice: the oil is heated to 70°C and mixed with 14° *Baume* (9,5 wt%) caustic solution to form soap and free fatty acids. The measured level of free fatty acids in the oil determines the addition rate of caustic. Typically, caustic is added at a rate equivalent to a 113% stoichiometric excess. Wash water, also heated to 70°C, is added at a rate of 15% of the crude oil mass flow rate. Some yield loss is assumed to result from the saponification of triglycerides according to the follow reaction:

		NaOH			
$C_3H_5(OOCR)_3$ +	$3H_2O$	--->	3RCOOH	+	$C_3H_5(OH)_3$
(Triglyceride)	(Water)	(Caustic Soda)	(FFA)		(Glycerol)

The mixture of oil, soap, and wash water is sent to a centrifuge to separate soap and water from the oil. One per cent of the oil phase is assumed to be lost with the soap and water. Of the oil entering the refining section, a total yield of 96% is sent to the transesterification section. A breakdown of the source of the losses is shown in Table 53.

FIGURE 81. THE TRANSESTERIFICATION REACTION INVOLVED TO TRANSFORM THE TRIGLYCERIDES IN BIODIESEL.

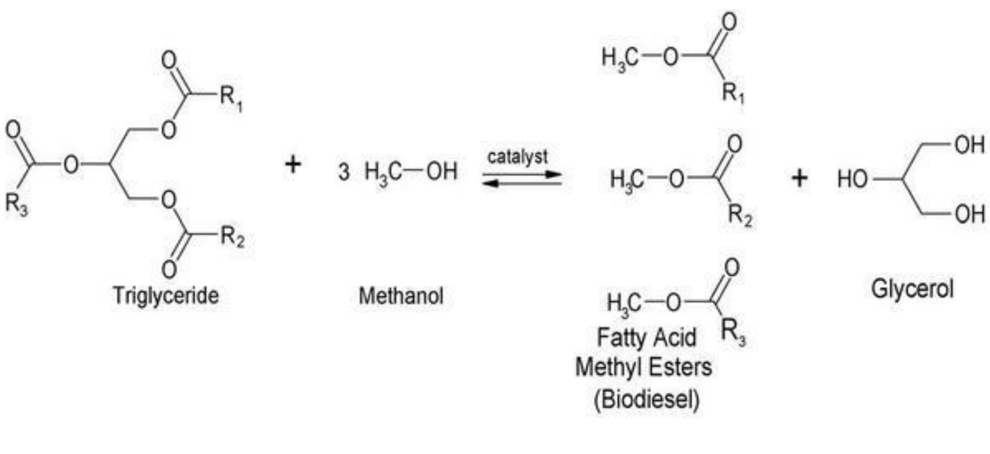


TABLE 53. LOSSES ASSOCIATED WITH ALKALINE REFINING OF THE CRUDE, DEGUMMED OIL

COMPONENT	LOSSES
Free Fatty Acids	0,719%
Triglycerides	1,796%
Unsaponifiable Matter	1,485%
Total	4,00%

A little more than half the losses are due to removal of unwanted components (the free fatty acids and unsaponifiable matter). The remainder is loss of triglyceride. About half the triglyceride loss is due to the 1% carryover of oil in the wash water, and half is due to saponification. After the oil refining the process of biodiesel production can start. This process is explained in the schema of Figure 82. The model described here is intended to be generic, and representative of contemporary industry practices. It is not meant to represent the actual biodiesel design offered by any single technology provider. The yields have been calculated using SuperPro Designer v. 8.5. This software, released by Intelligen, Inc., is free downloadable, together to a model of a biorefinery for biodiesel production from soybean oil . The capacity of the model is 10 million of gallons of biodiesel per year (MGPY). Adapting the incoming value of oil to our case, the yields were found, together with the efficiencies, of the production process.

TRANSESTERIFICATION

The stoichiometry of the reaction requires three molecules of methanol for every molecule of triglyceride reacted. On a weight basis, this corresponds to adding methanol at a rate of about 10% by weight per mass of oil processed. However, to obtain high yields and reasonable reaction times, an excess of methanol is usually used, e.g. twice the stoichiometric requirement. This translates to six molecules of methanol for every molecule of triglyceride in the oil. Excess methanol remaining after the reaction is recovered later in the process. Alkali metal hydroxides or alkoxides can be used as transesterification catalysts. Hydroxides are cheaper than alkoxides, but must be used in higher concentrations to achieve good reaction. In commercial practice, a variety of base catalysts has been used to for this reaction. These include sodium hydroxide and potassium hydroxide, as well as sodium methoxide. For consistency with [224] and [162], sodium methoxide has been selected in the model. The catalyst is present at a level of 10% in the methanol added to the reactors. Reported temperatures in the reactors vary from 50° to 120°C. A temperature of 60°C is usually chosen, because the lower end of the temperature range is typical of more modern commercial facilities. These reactors require significant mixing to achieve good contact between the methanol and oil phases. Two sequential transesterification reactions were modelled (Figure 82, REACTORs 1 and 2). The first reactor was continuously fed with oil and a solution of sodium methoxide in commercial grade and methanol. Product was removed from the reactor at a rate equal to the rate of charging with reactants and catalyst in such a manner as to give a residence time of 1 h in the reactor. Glycerol separates from the oil phase as the reaction proceeds. Following the first transesterification reaction, continuous centrifugation (Figure 82, CFUGE1) is employed to remove the glycerol-rich co-product phase, which is sent to the glycerol recovery unit. The methyl ester stream, which also contains unreacted methanol and oil, and catalyst, is fed into a second steam jacketed, stirred tank reactor (Figure 82 REACTOR 2), accompanied by the addition of sodium methoxide, 1.78% (w/w) in methanol. Again, a continuous stirred reaction is conducted at 60°C, with the crude ester product being removed from the reactor at a rate equal to that of reagent addition and in such fashion as to produce a reactor residence time of 1 h. A transesterification efficiency of 90%, well within the range of reported values [229] [230], was assumed for each of these two transesterification reactions, for an overall efficiency of 99%. The mixture of methyl esters, glycerol, unreacted substrates and catalyst exiting the second reactor was fed to a continuous centrifuge (Figure 82, CFUGE2). Typical municipal quality water is used for this, and all subsequent, washes. The glycerol-rich aqueous stream from this operation is sent to the glycerol recovery section while the impure methyl ester product goes to the biodiesel refining section for purification and dehydration (Figure 82, MIXING1 and CFUGE3). No other losses of oil or product occur in this step. Potential losses can occur in the settling tanks or in mixing tanks if some of the oil phase is carried over with the aqueous phase. It has been assumed that the settling tanks are being operated for maximum recovery of product, at the expense of some carryover of aqueous phase in the oil. Purification of the ester removes the glycerine, methanol, and water carried over. The chemistry of transesterification should yield almost exactly 1 kg of biodiesel per kg of crude algal oil entering the transesterification reactors (the correct efficiency found here is 99,4%). This model facility produces biodiesel with a whole mass yield of 95,4% of the incoming oil from the extraction phase.

METHYL ESTER PURIFICATION

The crude methyl ester stream is washed with water at pH 4.5 (thanks to the addition of HCl in 1.6% (w/w) quantity) to neutralize the catalyst and convert any soaps to free fatty acids, reducing their emulsifying tendencies (Figure 82, MIXING1). Centrifugation is then employed (Figure 82, CFUGE3) to separate the biodiesel from the aqueous phase. The latter is cycled to the glycerol recovery section. The crude, washed methyl ester product may contain presence of water. This must be lowered to a maximum of 0.050% (v/v) to meet United States biodiesel conditions. Water is removed in a vacuum dryer (Figure 82, VFLASH) from an initial value of 2.4% to a final content of 0.045%. Free glycerine and methanol are assumed to be removed by the vacuum dryer. The final composition of biodiesel is summarized in Table 54.

GLYCERINE and METHANOL RECOVERY

The glycerol liberated during transesterification has substantial commercial value if purified to USP grade. However, this process is expensive. Small and moderately sized operations, including those of the scale modelled here, often find it most cost effective to partially purify the glycerol, removing methanol, fatty acids and most of the water, and selling the product (80% glycerol by mass) to industrial glycerol refiners. In this study the production and sale of such a partially pure glycerol co-product has been included.

TABLE 54. FINAL COMPOSITION OF BIODIESEL PRODUCED AFTER TRANSESTERIFICATION AND PURIFICATION OF METHYL ESTERS.

COMPONENT	WEIGHT %
Water	0,045%
Unreacted oil	0,756%
Methyl Ester	99,180%
Glycerine	0,005%
Methanol	0,001%
NaOCH3	0,003%
Sodium Chloride	0,009%

In the model, the impure, dilute, aqueous glycerol streams exiting the transesterification reactors and the biodiesel wash process are pooled. The mixture is then treated with hydrochloric acid (Figure 82, MIXING2) to convert contaminating soaps to free acids, allowing removal by centrifugation (Figure 82, CFUGE4). This fatty acid waste is presumed to be destined for disposal as sewage, although in some contemporary industrial settings it has market value. The glycerol stream is then neutralized with caustic soda (Figure 82, MIXING3). Methanol is recovered from this stream by distillation (Figure 82, MEOH DISTILL) and is recycled into the transesterification operation (Figure 82, REMEOH). Finally, the diluted glycerol stream is distilled to reduce its water content (Figure 82, H2O DISTILL). At this point the glycerol concentration is 80% (w/w), suitable for sale into the crude glycerol market. Water recovered during drying of the ester and glycerol fractions is recycled into wash operations (Figure 82, RWATER). The model includes maximum recovery of the heat present in condensates, transferring it via heat exchangers to the material feed streams entering reactors.

WASTE TREATMENT

Wastes collected from alkali refining, methyl ester purification, and methanol recovery are sent to clarifiers for removal of oil and grease. The oil and grease skimmed off the wastewater is land filled; the remaining wastewater is sent to the municipal sewer system.

To calculate the final yields of the process, an efficiency of 96% has been used for the oil refining step. Then, according to the simulation done with SuperPro Designer software v. 8.5 the following final efficiencies were obtained: 0,093% (w/w) for glycerol and 99,4% (w/w) for biodiesel, referred to the incoming oil from the refining step.

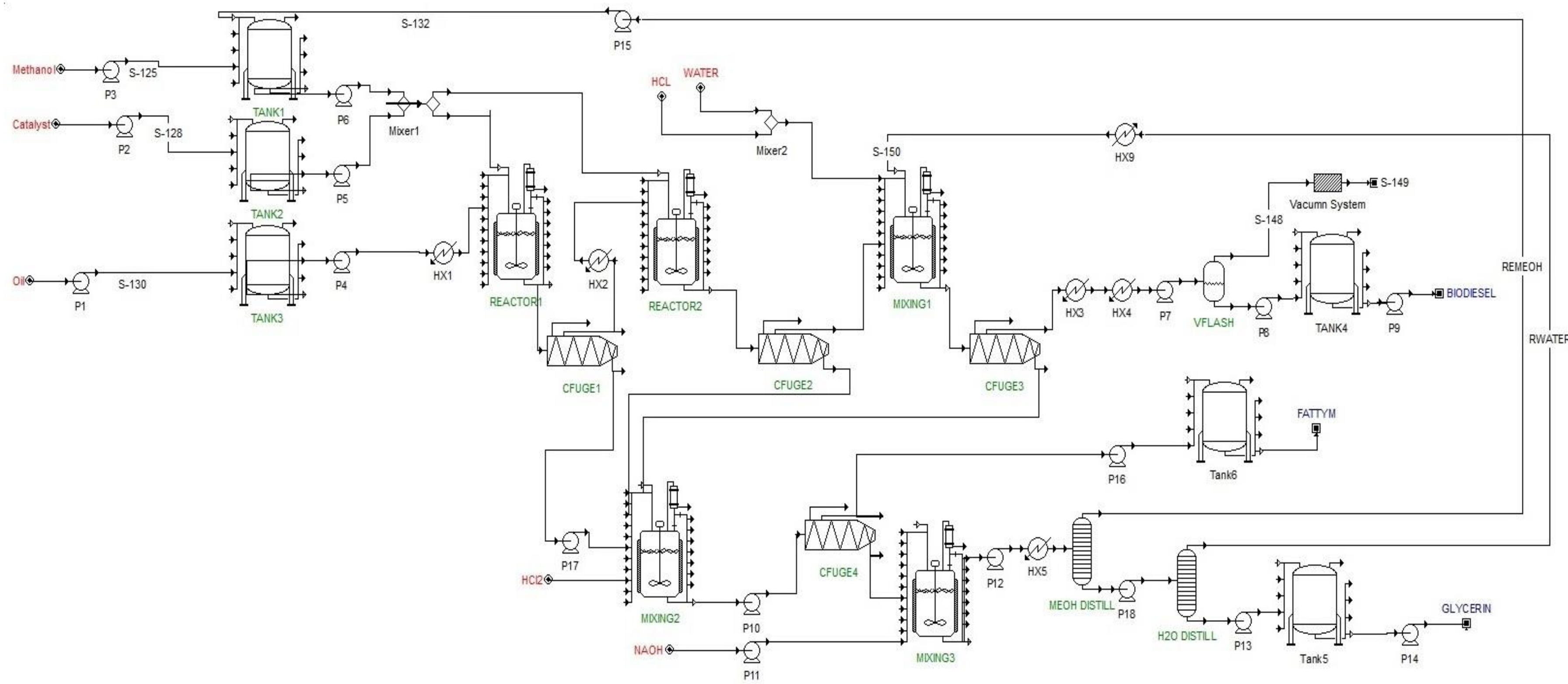
For this study, the power requirements were calculated similarly to the previous steps of the production chain. A comparison between different references was done in Table 55 and a value was chosen.

In addition, a simulation with SuperPro Designer was done. The data supplied in the on-line model for a biorefinery with a capacity of 10 MGPY (million US gallons per year) report an electric energy and heat requirements of 23,2 kWh/ton of biodiesel produced and of 444,7 kWh/ton, respectively. Adapting the incoming oil flow and the addition of other materials (like methanol, catalyst, etc.) to simulate a biorefinery of 8 MLPY (million litres per year), similar to the final yields found in this study, energy requirements were the following: 89,3 kWh/ton for electricity and 748 kWh/ton for heat. Because of the averages of values reported in literature are lower, in this study were used values reported in [98], that represent a good compromise between the simulation and the literature.

E. Bioethanol Production

Ethanol production is based on a fermentation process that is a biological process where sugars contained in biomasses are converted to alcohol. As for biodiesel, nowadays an established and well-known technology doesn't exist for a commercial-scale plant for production of bioethanol from algae. For this reason, the content and composition of microalgae's carbohydrates were studied to verify which pretreatment process could be used to hydrolyse polysaccharides, in order to obtain fermentable monomers of glucose. This analysis is done to verify if it is possible to use the same conversion schema typical for fermentation of other feedstock, such as corn. Carbohydrates are polysaccharides, i.e. they are the polymers of monosaccharides.

FIGURE 82. THE SCHEMA OF THE BIODIESEL PRODUCTION PROCESS.



Oil, methanol and catalyst enter in the 2 transesterification stirred reactors. Then the mixture is separated via usage of centrifuges. then biodiesel enter in the mixing tank to be washed by water and be purified. Whereas methanol and glycerol follow their recovery and washing processes, meanly by distillation.

TABLE 55. ENERGY REQUIREMENTS AND YIELDS FOR THE CONVERSION OF CRUDE OIL TO BIODIESEL.

PROCESS	BIODIESEL YIELD (w/w)	GLYCEROL YIELD (w/w)	ELECTRIC ENERGY [kWh/ton[32]]	HEAT [kWh/ton]	REFERENCE
Transesterification	-	-	45	227	[33] base case
transesterification	-	-	98,5	495,8	[33] best case
transesterification	-	-	-	250	[89]
base catalyzed transesterification	1	0,109	-	-	[141]
transesterification	0,95	0,2	28	348,9	[227]
oil refining	-	-	27,8	166,7	[98]
Transesterification	-	-	55,6	444,4	
base catalyzed transesterification	0,99	0,1	23,2	444,7	Software SuperPro Design (capacity 10 MGPY[31])
base catalyzed transesterification	0,994	0,09	89,3	748	Software SuperPro Design (capacity 8 MLPY)
-	-	-	11,9	208,3	[228]
base catalyzed transesterification	1	0,1	28,9	381,4	[224]
-	0,96	0,213	-	-	[231]
Transesterification	0,98	0,14	51	413	AVERAGE
	0,994	**0,09**	**83,4**	**611,1**	**This study**

In this study, the yields were calculated with SuperPro Desinger, while the energy requirements used are from [98], because it represents a fair compromise between the averages found in literature and what was calculated with software (case of 8 MLPY capacity).

[31] MGPY = million of gallons (US gallons) per year; MLPY = million of litres per year

[32] Electricity and heat requirements are expressed per metric tons of biodiesel produced

There are many types of carbohydrates, among which the most abundant in nature are starch and cellulose (the latter usually compares even with hemicellulose). All these 3 kinds of carbohydrates are polymers of glucose; the feature that mainly characterizes each polysaccharide is the type of bonds that keep together the molecules. In the following is inserted a brief description of these carbohydrates, that are present in microalgae.

STARCH

Starch is a glucose polymer, that is a glucose molecular complex joined together by glycosidic bonds. Pure starch is a white, tasteless and odourless powder that is insoluble in cold water or alcohol. It consists of two types of molecules: the linear and helical amylose and the branched amylopectin. Depending on the plant, starch generally contains 20 to 25 % amylose and 75 to 80 % amylopectin. Amylose is not branched polysaccharide of glucose with a helical structure. Glucose molecules are joined together by glycosidic bonds α (1→4). Amylopectin has a slightly more complex structure, because the glucose chains are branched. In the linear chains there are always α (1→4) bonds type, whereas in the branch point there are α (1→6) bonds type (see Figure 83). Concerning to the ethanol production, starch must be converted with enzymes (or acids) to dextrins[33] first, and then to sugars through a hydrolysis process. This permits to make accessible the simple glucoses for fermentation with yeast.

CELLULOSE

The largest fraction of carbon that can be found in biomass is cellulose. Cellulose is a glucose polymer, too, but in this case glucose is joined together by β-glycosidic bonds. These bonds create linear chains very stable and much more resistant towards the chemical attacks because of the high strength of hydrogen bonds that are created among the cellulosic chains, as shown in Figure 84. The hydrogen bonds make the polymer more rigid thus inhibiting the molecular flexibility that has to be guaranteed to break the glycosidic bonds. Moreover, the conversion of cellulose to simple sugars is usually complicated by the presence of lignin. Lignin is a heterogeneous aromatic polymer that forms a physical seal and supports cellulose fibrils in plant cells [232]. This evolutionary trait allowed terrestrial plants to adapt to land ecosystem and made plant cell walls very resistant to microbial degradation. From biomass conversion point of view however, it is a problematic trait as it requires introducing expensive pre-treatment steps to liberate cellulose fibres for hydrolysis. Despite years of research and development it is still an active area of research for optimisation and improvement [233]. However, the key difference between microalgae and common terrestrial plants is absence of lignin. The crystalline structure of algal cellulose can be advantageous when selecting appropriate mixture of enzymes [21]. The particular class of enzymes that attacks the cellulose is called cellulase. It should be noted, however that cellulose-cellulase interactions are complex and various cellulases of the same organism often exhibit preference towards different forms of cellulose to ensure efficient hydrolysis. It is therefore essential to select appropriate enzymatic cocktail for particular feedstock and carefully select method assessing cellulase performance that can differ significantly with methodology used [234].

33 Dextrins are a group of low-molecular-weight carbohydrates produced by the hydrolysis of starch or glycogen. Dextrins are mixtures of polymers of D-glucose units linked by α-(1→4) or α-(1→6) glycosidic bonds. They are a short chain of glucoses, shorter than the original polysaccharide. Dextrins can be produced from starch using enzymes like amylases, as during digestion in the human body.

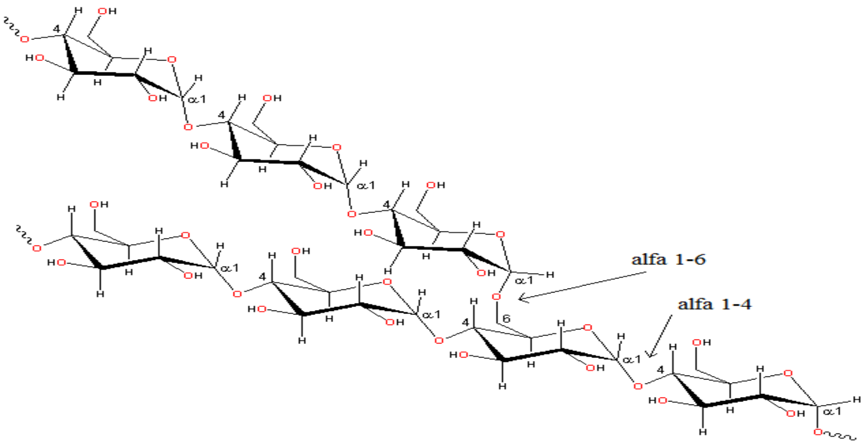


FIGURE 83. THE STRUCTURE OF THE STARCH. IT IS POSSIBLE TO RECOGNIZE THE SEQUENCE OF MONOMERS OF GLUCOSE THAT IS REPEATED WITHIN THE MOLECULE. ALSO, THERE ARE UNDERLINED THE TWO DIFFERENT BOUNDS OF THIS POLYMER: α (1-6) AND α (1-4).

HEMICELLULOSE

Hemicellulose consists of a short and much branched chain of sugars (from 10 to 200 molecules). It is constituted by sugars with five carbon atoms (generally xylose and arabinose) and sugars with six carbon atoms (glucose, mannose and galactose). Sugars are often substituted by acetic acid. Hydrolysis of hemicellulose creates products with high content of xylose if raw material comes from hard wood, while it produces more six carbon atoms sugars if coming from tender wood and plants. Hemicellulose is present in almost all cellulosic biomasses, with percentages that go from 10% to 40%.

Carbohydrates of microalgae are composed both by starch (it is within the cells and it supplies the energetic reserves for the organism) and by cellulose and hemicellulose, to form the walls of the cells [106]. Accordingly to [9], it is assumed that hemicellulose is not converted in ethanol. Many studies have been done about the most suitable pretreatment process to use for hydrolysing the carbohydrates of algae. Both acid pretreatment [235], alkaline pretreatment [236], and enzymatic pretreatment [32] were considered here. Since the first application of microbial enzyme in the food industry in the early 1960s, a great deal of effort has been made to replace traditional acid hydrolysis with enzymatic hydrolysis in almost all glucose production due to higher yields under mild conditions, less by-products, and no corrosion issues [237]. Moreover, successful results have been reported for enzymatic pretreatment of *Chlamydomonas reinhardtii* by [32].

This strain is a "green alga", the same algal subfamily of *Chlorella* species. In that study, authors investigated the feasibility to use α-amylase for liquefaction phase and amyloglucosidase for the subsequent saccharification phase. They are the same enzymes used even to hydrolyse starch of corn feedstock for fermentation in ethanol. Since algae do not contain lignin, it is anticipated that the ethanol conversion process involving algae cake will not require a harsh lignocellulosic pretreatment. Based on preliminary studies [238], algae pretreatment would consist of a process similar to corn dry grind liquefaction process. Moreover in this latter paper, the carbohydrate converted via enzymatic pretreatment was cellulose.

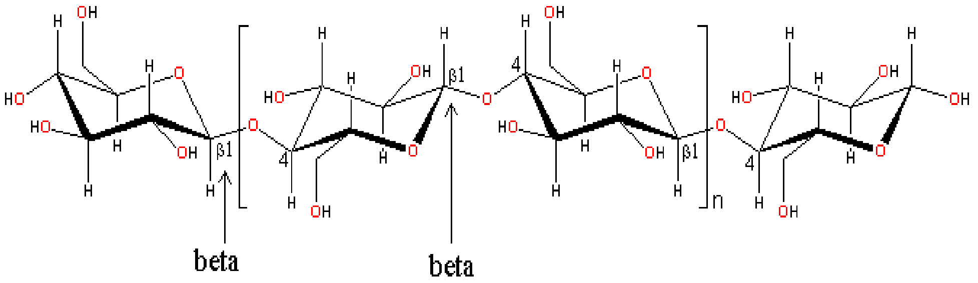


FIGURE 84. THE STRUCTURE OF THE CELLULOSE. THE SAME MONOMER (GLUCOSE) OF FIGURE 83 IS RECURRING INSIDE THIS MOLECULE. IN COMPARISON TO STARCH, HERE GLUCOSE IS BOUNDED WITH β (1-4) BONDS.

Hence, a research in matter of which kind of carbohydrate was hydrolyzable has been done, to understand the percentage of carbohydrates (or glucose) that could be converted in ethanol. This permits to simulate the yields of ethanol with more accuracy. After discussion with experts, and careful analysis of literature, it emerged that both cellulose and starch are hydrolyzable via enzymatic process. In particular, as reported in [19], *"a self-made enzyme mixture from the bacterium Pseudomonas sp."* is able *"to hydrolyze carbohydrate components (mainly starch and cellulose) in the microalgae feedstock"*. The results of study assert that:

1. *Chlorella vulgaris* was used to test the enzymatic process
2. *Chlorella vulgaris* was containing 51% (w/w) of carbohydrates (as in the base case scenario of this work)
3. Of these carbohydrates, the 90,4% is glucose (46,1% of the total). Numbers are also confirmed in [106] and very close to that found in [32], that used *Chlamydomonas reinharditii*, a different microalga of the same subfamily (green algae)
4. The SHF[34] and SSF[35] processes converted the enzymatic microalgae hydrolysate into ethanol with a 79,9% and 92,3% theoretical yield, respectively
5. The final yields obtained for enzymatic pretreatment are 17,8% and 21,4% of grams of ethanol on grams of algae, very similar to acid pretreatment (with sulfuric acid). These represent about the half of the initial glucose content, since the other half is converted in CO_2, during the reaction, see equation (11).

The whole process to obtain ethanol is therefore described in this section. For production of ethanol from corn, two different technologies exist: a dry mill route and a wet mill route. Regarding what discussed earlier, the dry mill production process has been adopted and the wet mill is not further investigated. To find the final yield of ethanol, were used yields discussed in the point number 5 of the previous list. Yields were adapted, in order to consider the fact that in this work the biomass that enters in the enzymatic process has not got more lipids, and therefore the carbohydrates content is higher in percentage. This affect the final yield in percentage. The entire process can be divided in six steps (see Figure 86): milling, cooking, fermentation, distillation, dehydration, distillers grains and soluble.

[34] Separate Hydrolysis and Fermentation
[35] Simultaneous Saccharification and Fermentation. Complete explanation of these processes will be done in the next pages

MILLING

Milling usually is the initial preparation of biomass. Its aim is to make the downstream processing more efficient, easier, and quicker. Because of the algal biomass is already be grounded before the oil extraction phase, the milling process is not be considered for this part of the production chain.

COOKING

To pre-treat the starch of a biomass in general (such as algae or corn) the following process has to be done. The powder of algae is mixed with water. During the cooking process two different enzymes are introduced. The first one, the endoenzyme *α-amylase*, does a chemical fragmentation of the starch polymer into short section. The second enzyme, the esoenzyme *glucoamylase*, chemically breaks the shorts sections in simple sugars. Cooking process can be divided in three continuous phases: liquefaction, sterilization and saccharification. In the liquefaction phase, cake is initially mixed with hot water (usually 88°C) for a period that ranges from 20 minutes to one hour, followed by a high temperature steam injection for a few minutes. While mixed with water the microcrystalline structure of algae adsorbs humidity, swell up and become fragile losing their crystalline structure. This structure is then easily broken when steam is added. Physical breakdown of the molecules allows the introduced enzymes to attack all the available starch that otherwise would not be converted to ethanol in the downstream fermentation phase. Meal is also mixed with the *α-amylase* enzyme. This enzyme attacks randomly the α-(1-4) bonds producing molecule of variable length called dextrins. They are oligosaccharides (short polysaccharides) with 3 or more glucoses bounded. Together with this enzyme other products are introduced to maintain the ph around 6 in order to guarantee the maximum conversation velocity, calcium as enzyme nourishment and urea that supplies nitrogen necessary for the downstream fermentation phase. The process is called liquefaction because, with increasing incubation time, shorter and shorter dextrins will be produced, leading to a lower viscosity of the slurry. This aspect is also useful to evaluate the enzymatic efficiency. In the sterilization phase temperature rises to 110°C for about 20 minutes, in order to avoid bacterial contamination and to complete the previous crystalline structure breaking. During saccharification phase, mixture temperature is lowered to 80°C and the *glucoamylase* enzyme, necessary to prepare the starch for the fermentation process, is introduced. This enzyme does not attack randomly the chain, but only attacks one side of dextrins thus producing only one glucose molecule per dextrin at a time. It follows that *glucoamylase* enzyme requires about 2 hours to achieve a complete conversation; two times the time required by *α-amylase*. Also this enzyme must work with restricted ph values (about 4.4), thus requiring the addition of acids. *α-amylase* enzyme cannot hydrolyse α (1-6) bonds, whereas *glucoamylase* is not efficient towards them. Normally this is not a big problem because α (1-6) bonds are much less frequent than α (1-4) bonds.

However, this process must be adapted to hydrolyse also the cellulosic content of the carbohydrates. To enter in further chemical and biological details of the enzymatic process is not the aim of this work. Here it is only said that the right enzymatic mixture to hydrolyse as much as glucose possible consist of: endoglucanase, β-glucosidase and amylase [19]. This mixture has given a proof to attack both starch and cellulose, giving a final glucose yield of 90,4% of carbohydrate content in *Chlorella vulgaris*.

FERMENTATION

The liquid slurry coming from the previous phase of the process is pumped inside vessels together with a large quantity of yeast. In this way yeast can convert simple sugars (mainly glucose and maltose) to ethanol, CO_2 and heat. Fermentation time range depends on the yeast type, temperature and final ethanol concentration desired. Several types of microorganisms can be used for the fermentation process: yeasts, bacteria, fungus. Historically the most commonly used is the *Saccaromyces cerevisiae* yeast that can produce ethanol with a maximum concentration of 18%. Yeast is an anaerobic facultative enzyme, meaning it can work with the presence of oxygen and also when oxygen has been totally consumed. When working with oxygen it converts glucose to water and CO_2; if it works without oxygen it produces ethanol, CO_2 and heat. Fermentation is a complex series of chemical reactions that can be summarized by equation (11):

$$C_6H_{12}O_6 \rightarrow 2C_2H_5OH + 2CO_2 + heat \qquad (11)$$

Therefore, of the incoming amount of glucose, about half is converted in ethanol, the other half is converted in CO_2. Fermentation is a slow process that requires from 50 to 60 hours to be completed. Modern batch systems require a big supply of yeast to maintain a high output level. Yeast is introduced in a small batch some hours before the beginning of fermentation where a mixture of water, malt, enzymes and nourishments is added in order to guarantee the growth and the anaerobic reproduction of yeast. In this way, once introduced inside the fermentation batch, yeast will already be at its maximum activation level and fermentation can start immediately. Most important aspects that must be controller during the fermentation process are sugar concentration, ethanol concentration, temperature (that must be maintained under 32°C), bacterial infections and nourishment levels. These aspects are also called yeast stress factors. A higher or lower level than nominal of just one of these factors would not create problems but only the reduction of yeast working efficiency. The problem would get more critical when simultaneous stress factors occurred, because this situation could lead to the death of yeast. Furthermore, even without the death of yeast, its conversion velocity could not return to its optimal levels and it will not be able to complete the fermentation and starch unfermented would remain. Concerning the fermentation process, some innovations are identified. Their aim is always the reduction of energy needs and of conversion time [239]. The most important innovation regards the combination of saccharification and fermentation processes: realizing a saccharification process with high sugar levels helps the reduction of fermentation times, but osmotic stress due to this high percentage can create yeast stress in the batch type fermentation. The introduction of "*Simultaneous Saccharification and Fermentation process*" (SSF) has allowed the generation of sugars directly inside the fermenters, leading to a higher saccharification velocity guaranteed by the removal of sugar in the same time it is generated. Some plant layouts have introduced also the yeast propagation, thus creating the "*Simultaneous Saccharification Yeast Propagation and Fermentation process*" (SSYPF). There are three potential advantages due to the combination of these three phases: higher conversion efficiency, anticipated ethanol production thus reducing the possibility of bacterial contamination, the reduction of storage of big quantity of yeast.

DISTILLATION

Fermented material is sent to distillation columns where added heat allows the boiling of ethanol and its separation from water and other components. When fermentation is completed, the next step is the purification of ethanol in order to separate ethanol from other compounds, mainly water. Distillation is the thermal process more convenient to separate ethanol and water, thanks to their different boiling points: 77.7°C for ethanol, 100°C for water (at p = 1 atm). Heating the water-ethanol mixture inside a vessel, liquid water would remain at the bottom. Separation cannot occur in one single heating for the presence of intermolecular interaction because the difference between the two boiling points is not infinite: hence distillation columns are used. Distillation process cannot completely separate water from ethanol because at a volume concentration of 97.2% the compound becomes azeotropic, as strong intermolecular bonds between water and ethanol are present. At the azeotropic concentration the boiling vapour has the same composition as the liquid it comes from, therefore even more distillation steps would not further separate the two compounds. Gases extracted to decrease pressure inside the distillation columns are sent, together with CO_2, to a water scrubber. Water mixed with ethanol exiting from the scrubber is sent back to distillation columns, whereas CO_2 is usually vented to atmosphere or sometimes can be sold [240], or used to grow algae in open ponds.

DEHYDRATION

At the bottom of distillation column it's produced ethanol with a 3-5% v/v of water concentration. Remaining water can be eliminated through a further process called dehydration. At the end of this phase the final product is nominally 100% ethanol. Dehydration is a necessary step for the utilization of ethanol as fuel, because presence of water inside ethanol and gasoline blends (3% represents a significant quantity) increases the risk of phase separation. Industrial dehydration processes utilize two methods: azeotropic distillation and molecular sieves. Azeotropic distillation requires the introduction of a third compound, during the distillation phase, that *interacts* with both initial compounds creating an azeotropic ternary mixture. In ethanol industry benzene is used as third compound: when it is introduced inside the distillation columns three distinct regions with different concentration of the three compounds will appear. At the bottom remains a mixture of mainly water, in the middle an azeotropic ternary mixture of water, benzene and alcohol and at the top of the column dehydrated ethanol. Disadvantages of this process are the following:

- Large amount of energy required
- A complex system of columns to allow the benzene to recover
- The amount of ethanol and benzene contamination when at optimal working values are not guaranteed
- Security implication of benzene use and storage, as it is a highly inflammable and carcinogenic compound.

For these reasons the azeotropic distillation is not frequently used. The other and most common dehydration method is to separate the azeotropic mixture through molecular sieves. Molecular sieve are zeolites made in small spheres that have a very precise and small size of pores (Angstrom or nanometres order of magnitude). For ethanol industry sieves with 3 A diameter pores are used, since ethanol molecules have a diameter of 4 A and those of water are equal to 2,8 A. Therefore, water molecules can pass and enter the sieves, whereas the ethanol ones remain outside. Besides pore feature, the superficial absorption is an important aspect of separation efficiency. In fact, when water is absorbed by the surface of sieve, the sieve itself retains the molecules until a sufficient amount of external energy is introduced to remove water. Methods for sieve regeneration are two: sieves heating to let the water leave the sieve as steam, or a pressure decrease. Heat is normally supplied through external steam. Ethanol leaving this phase contains 99% weight ethanol. The final product is condensate, cooled and usually denatured with a 5% gasoline introduction.

DISTILLERS GRAINS AND SOLUBLES

This phase of the process is not directly linked to ethanol production. Its aim is to realize commercial co-products that can be sold. Waste material coming from oil extraction and fermentation contains still water, fibers, proteins, mineral, and not extracted lipids. Furthermore not the whole quantity of starch has been successfully converted to ethanol and a little amount is still conserved in the "waste". Actually, it is not definitively know at present whether residual algae meal can replace DDGS or other protein rich products [9]. In this study, it has been assumed that in the near future some improvements could be realized to the production cycle or to the algal strain. Then this worthwhile co-product, mineral and proteins rich, with also some residual content of lipids and carbohydrates, has been considered "sealable", and therefore a worth has been added in the revenues of the economic analysis. Water represents the highest percentage of the waste material and this aspect leads to 3 significant problems:

- Selling water together with fodder is not economically convenient
- Only a few animals eat wet fodder
- Presence of water can lead to growth of bacteria and thus a reduction of available time of selling animal fodder.

For these reasons a low cost process to reduce the amount of water has always been searched. Nowadays several methods are used, but almost all of them have an initial centrifuge phase that divides the byproduct stream into *stillage* (liquid) and *wet cake* (solid). The solid grains (35% in weight) leaves the centrifuge quite quickly. The thin stillage that leaves the centrifuge at about 95°C is partially reintroduced in the previous liquefaction phase, while the remaining part is concentrated inside an evaporator that eliminates water until a 25-50% solid is obtained, starting from 5-10%. This resulting material can already be sold but, due to its low commercial value and considering the high transport cost incidence, it is normally mixed with the wet cake in order to produce *Wet Distillers Grain with Solubles (WDGS).* WDGS are then dried becoming *Dried Distillers Grains with Solubles (*DDGS), material that contains about 9% weight of water. The WDGS would already have good nutritive features but, due to the still high water percentage, life of fodder is significantly reduced and cost of transport would have a very high incidence. In Figure 85 and Figure 86, two schemas are reported for the ethanol production cycle. Finally, to complete the model for the energetic evaluation, power consumptions and yields for ethanol production process have been considered. The power consumptions were taken by [125]: electric needs range from 0.75 to 1.20 kWh/gal and for this work a value of 1.00 kWh/gal (that corresponds to about 1 MJ/L) has been assumed; while for the thermal energy requirement a value of 7,4 kWh/gal (that corresponds to 7,05 MJ/L) is used in the model. The final yield of ethanol were calculated using a yield of 0,5 of grams of ethanol for grams of glucose content, after having subtracted the extracted lipids content.

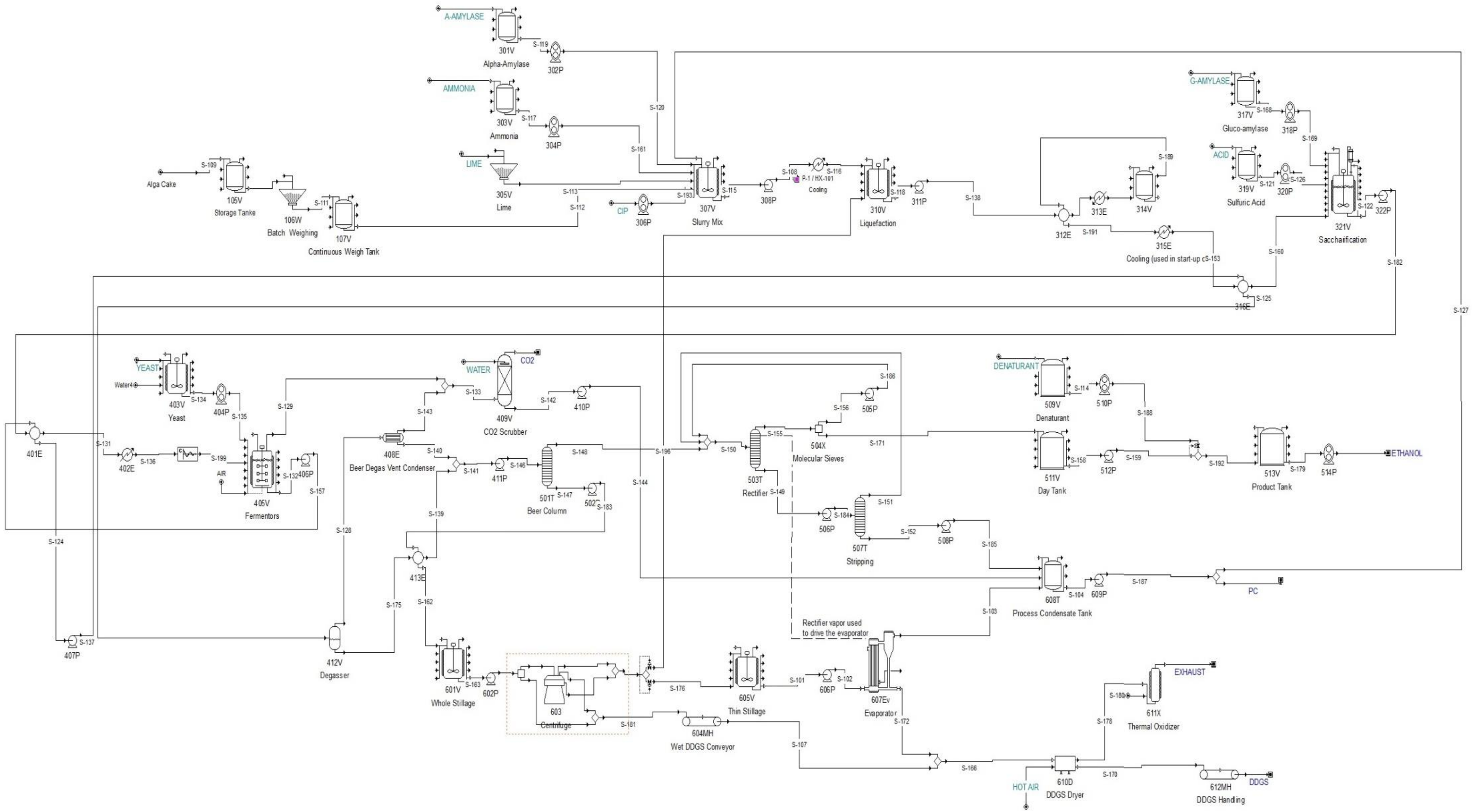


FIGURE 85. PLANT SCHEME FOR PRETREATMENT OF BIOMASS, FERMENTATION IN ETHANOL AND PURIFICATION AND WASHING OF OBTAINED PRODUCTS.

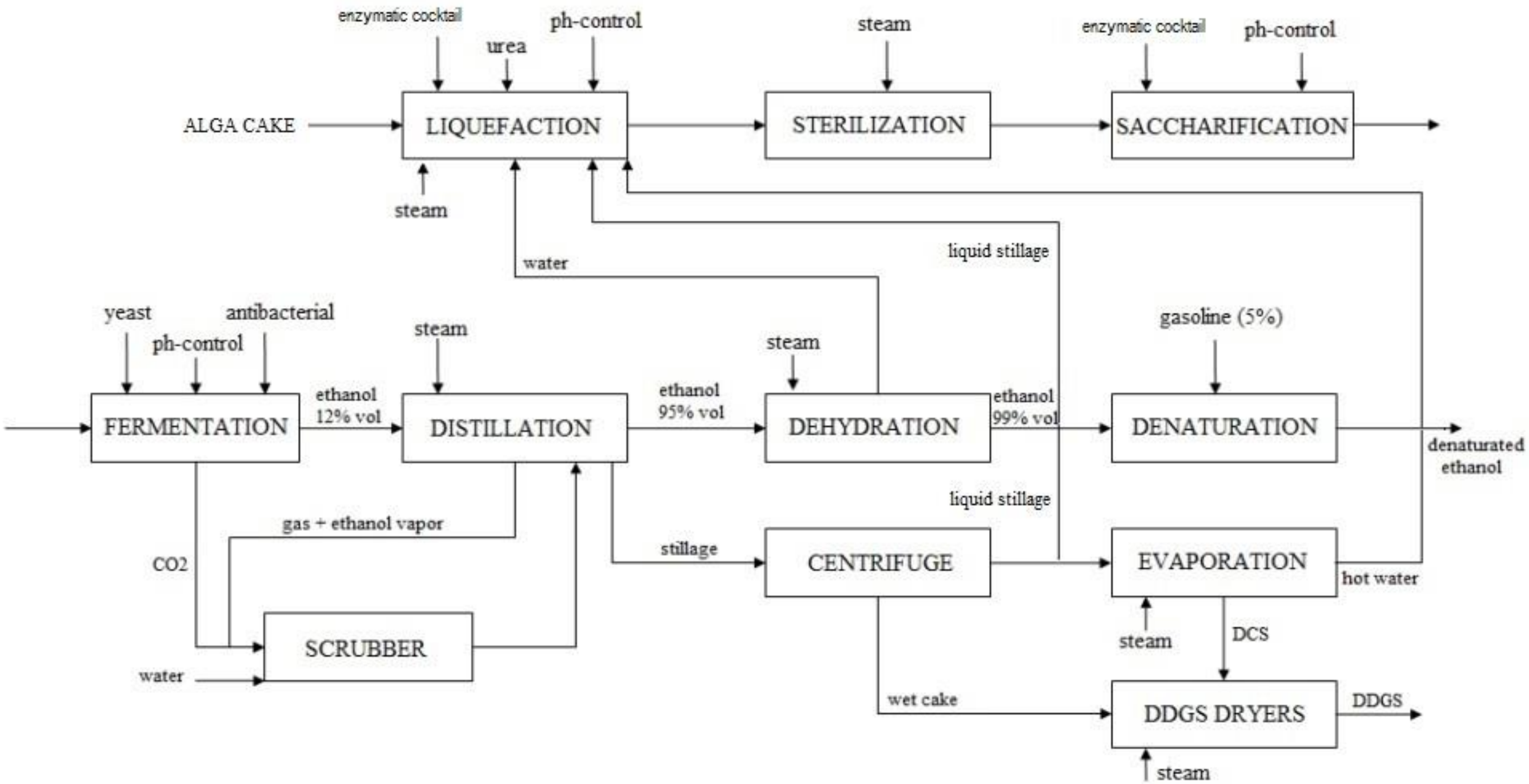


Figure 86. Schematic diagram of processes involved for production of ethanol from a biomass rich in carbohydrate. The alga cake coming from the extraction phase enter the process of liquefaction. Firstly sugar are hydrolysed, then they are fermented by yeast. After the fermentation process, products are sent to purifications and washing steps.

APPENDIX B.

CONVERSION OF REFERENCES DATA

In Table 46 is presented a list of power requirements for dewatering processes. For the consumptions supplied by [33], data were indicated in kWh/ton of algae cultivated. A simple conversation is done to indicate the consumptions in kWh/m^3, following the next steps.

Process	Energy requirements	Result of process in ref.
Centrifugation	26,5 kWh/ton	From 2% to 16%
Filtration	60 kWh/ton	From 16% to 30%
Thermal drying	1198 kWh/ton	From 30% to 85%

Because of data have to be expressed in function of m^3 of water, a correlation must be found between 1 ton of alga and 1 m^3 of water. Assuming that the density of algae is the equal to the density f water (i.e. rho_algae = 1 g/mL) [9]; using the results for every dewatering step indicated in the paper, it is found:

%	ton_of_algae	ton_tot	ton_H_2O
85	1	1/0,85 = 1,176	1,176-1=0,176
30	1	1/0,30 = 3,333	3,333-1=2,333
16	1	1/0,16 = 6,250	6,250-1=5,250
2	1	1/0,02 = 50,0	50-1=49

Thermal drying: from 30% to 85%; the slurry loses 2,333-0,176=2,157 ton of water.

Now the number are simply converted:

Process	Energy requirements	Result of process in ref.	Water [m^3]	Data converted
Centrifugation	26,5 kWh/ton	From 2% to 16%	49 (inlet)	26,5/49=0,54 kWh/m^3
Filtration	60 kWh/ton	From 16% to 30%	5,25 (inlet)	60/5,25=11,43 kWh/m^3
Thermal drying	1198 kWh/ton	From 30% to 85%	2,157 (removed)	1198/2,157=555 kWh/m^3

555 kWh/m^3 are equal to 1,998 MJ/kg, data suggested by [206]

Similarly, data by [89] is converted from kWh/kg of diesel produced to kWh/m^3. In this paper, it is reported that 5,93 kg of algae are necessary to produce 1 kg of biodiesel. Considering that here the dewatering phase is characterized only by filtration until 20% in dry biomass and, subsequently by a thermal drying until 90%, following the same method, it is found that, during the thermal drying, 3,89 (i.e. 4-0,11) kg of water are lost for every kilo of algae processed.

Because of the energy consumption is 22,72 kWh/kg of diesel, it corresponds to 3,83 kWh/kg of algae. Then it is necessary to adapt this consumption to kg of water, simply making a ratio: 3,83 kWh/kg_algae divided for 3,89 kg of water evaporated. It is found that the energy consumption is 0,98 kWh/kg of water removed, that corresponds to 984,6 kWh/m^3, that is 3,53 MJ/kg.

APPENDIX C.

FINANCIAL AND REAL OPTIONS

Some few examples of options.

Example 1. Financial Option [149]

A company (say HiTech) is a publicly held technology company whose stock is selling at $20 per share. The stock price is expected to rise significantly in the near future because of the company's innovative products and market demand. At the same time, however, there is also market uncertainty that indicates a possible sharp drop in the stock price. As an investor, you can buy shares of this stock today at $20 per share or instead buy options to buy or sell the stock in the future. For instance, let us say that you buy one option on the underlying stock of HiTech today at the market price of $2 that gives you a right to buy the stock one year from now at a price of $25. One year from now, if HiTech's stock price drops below $25, you can walk away with no obligation to buy the stock and lose the $2 that you paid to acquire the option. On the other hand, if the stock price goes above $25, to say $35, as a rational investor, you will exercise your option and buy one share of the stock. This would be worth $35, but you pay only the agreed-upon price of $25, thus making a gross profit of $10. Accounting for the initial price of $2 paid to buy the option, your net profit is $8. Thus, using the options approach, you would exercise your option (i.e., buy the stock) only if it goes above your exercise price; otherwise, you would walk away and take your up-front fee as a loss. In another scenario, you may acquire an option to *sell*. If you believe that the stock price of HiTech will be below $25 a share one year from now, you may buy one option of the stock at the market price of $2 that gives you a right to sell the stock one year from now at a price of $25 per share. If the stock price is above $25 per share on that day, you will not exercise the option, which expires and becomes worthless. However, if the stock price drops below $25, to say $15, you will exercise your option to sell one share of the stock worth $15 for a price of $25, making a gross profit of $10 and a net profit of $8 after accounting for the initial option price of $2. In both of the above scenarios, the options approach allows you to take advantage of the payoff when it is positive while limiting the downside risk.

Example 2. Real Option

Let's analyze an investment as an American call option. The firm can proceed with the investment at any time. In the analogy with real option, it represents an option to invest. The expected payoff is the sum of the expected future net cash flows conditional on undertaking the project (with present value V_t), minus the needed payment of the investment F_t. For example, today the investment cost Ft is 2,8 $ and the value Vt is 3$. This means that if the investment is made today the NPV will be

$$P_t = V_t - F_t = 3 - 2{,}8 = 0{,}2\$$$

Let's now suppose that the value of the project at time t+1 is uncertain (for example because the value of the output is uncertain) and will have equal probability to go to 2$ than to go to 4$. The average value remains still 3$, but the NPV of waiting is quite different since:

Time t		Time (t+1)			Decision
	p^{up}=0,5	V_{t+1}=4	$\longrightarrow$	NPV_{t+1}=1,2$	The investment is done
V_t=3					
	p^{down}=0,5	V_{t+1}=2	$\longrightarrow$	NPV_{t+1}=0$	The investment is not done

Indeed, in case the value became V=4, the NPV will be then

$$NPV_{t+1} = p^{up} \cdot \left(V_{t+1}^{up} - F_t\right) + p^{down} \cdot (V_{t+1}^{down} - F_t) = 0{,}5 \cdot (4 - 2{,}8) + 0{,}5 \cdot 0 = 0{,}6\$$$

where up and down represent simply the values in case the worth of V increases or decreases, respectively. In this example of “equal probability” both p^{up} and p^{down} are set to 0,5.

Therefore, the option to wait has a value that increases the standard NPV. The expanded NPV, which considers the value of the flexibility is then:

Expanded NPV		Standard NPV		Value of the option
0,6$	=	0,2$	+	0,4$

However, in general, differently from the financial options, there is uncertainty not only about the value of the underlying but even on the value of the corresponding strike price.

APPENDIX D.

THE DYNAMIC PROGRAMMING METHOD

The general problem of an American options, that can be exercised at any time prior to expiration time T, is to decide the more profitable exercise moment τ<T. The value of these options is defined as:

$$F(t, X_t) = \max_{\tau<T}\{E_t^*[e^{-r(\tau-t)}\Pi(\tau, X_t)]\} \quad (2)$$

in which $F(t, X_t)$ is the value of the option at time t and with state variables with value X_t, $\Pi(t, X_t)$ the payoff at time t and state variables X_t, and E_t* [ ] is the expectation conditional to the information available at time t [241]. This value is achieved by determining the optimal stopping time τ within the interval [t, T] when one should exercise the option to maximize it. At expiration time, the value of the option is equal to the payoff:

$$F(T, X_t) = \Pi(T, X_t) \quad (3)$$

Dynamic programming splits the problem of deciding when to exercise some option into multiple sub-problems, one for every time step. This is a necessary but not sufficient condition for optimality. These sub-problems consist in deciding at every time step if it's better to exercise the option in that time step or to wait the next one, comparing the payoff from immediate exercise $\Pi(t, X_t)$ with the expected payoff from continuation $\varphi(t, X_t)$.

Rationally the owner of the option will exercise the option when the immediate payoff is higher. Using the Bellman equation, equation (3), the value of an option F (t, X_t), depending on the time t and on the value of the state variables X_t is the maximum between the payoff $\Pi(t, X_t)$ and the value of continuation $\varphi(t, X_t)$, that is the value to wait until next time t+1.

$$F(t, X_t) = max\{\Pi(\tau, X_t), e^{-r}E_t^*[F(t+1, X_{t+1})]\} \quad (4)$$

$$\varphi(t, X_t) = e^{-r}E_t^*[F(t+1, X_{t+1})] \quad (5)$$

The continuation value is then the actualized expectation of the value of the option at the next time. The path-wise optimal policy will then be to exercise the option only if $\Pi(t, X_t)$> $\varphi(t, X_t)$.The equation (3) is resolved backward.

APPENDIX E.

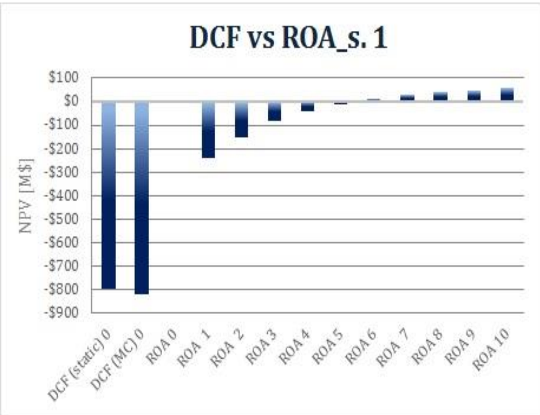
RESULTS & SENSITIVITY ANALYSIS

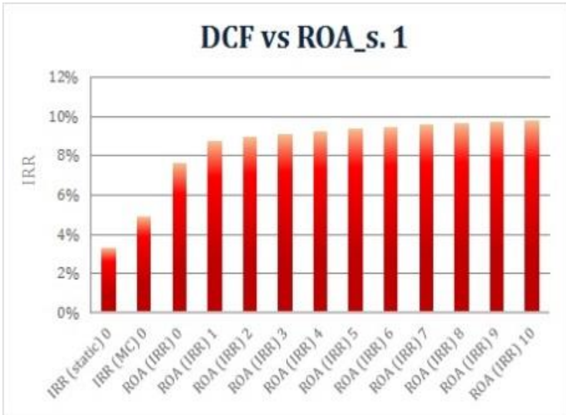
OPTION TO BUILD: BIOREFINERY

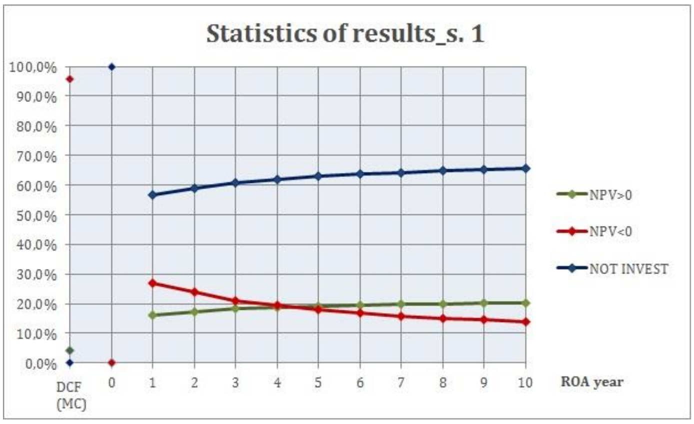
For every scenario, the following items are reported: input values that characterize the scenario, graphs of NPV (blue) and IRR (red) in relation to the method used (DCF or ROA at year n), statistics for the investment. In particular the latter underlines the probability that the investment returns back a positive NPV (green line), a negative NPV (red line) or that the manager decides to do not invest (blue line).

Scenario 1 – cheap diesel

BIODIESEL [$/L]			BIOETHANOL [$/L]			ELECTRICITY [$/kWh]			DRIFT		
min	av	max	min	av	max	min	av	max	min	av	max
1,235	1,3	1,365	0,665	0,7	0,735	0,076	0,08	0,084	0%	2%	4%

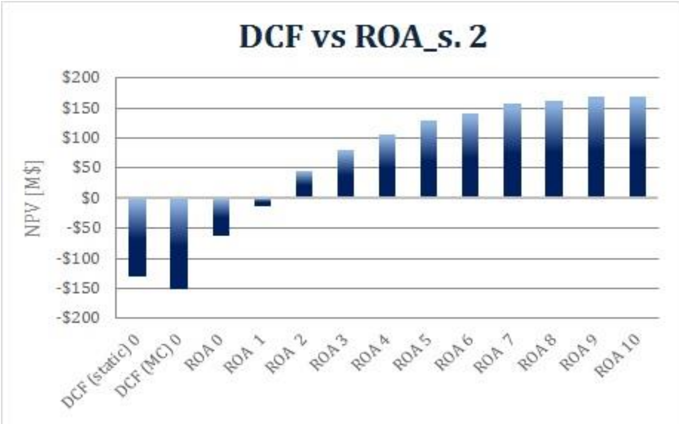


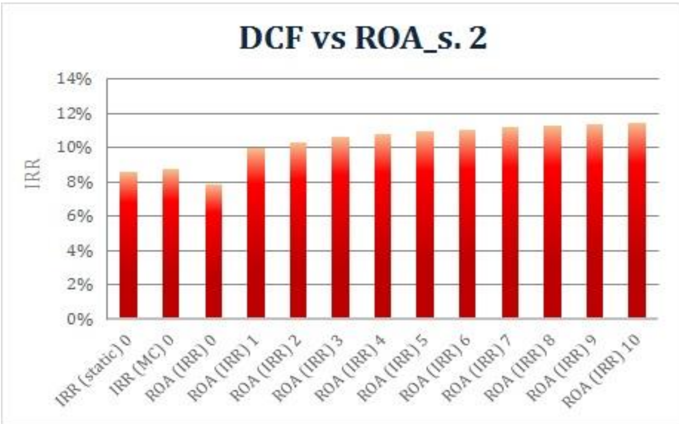


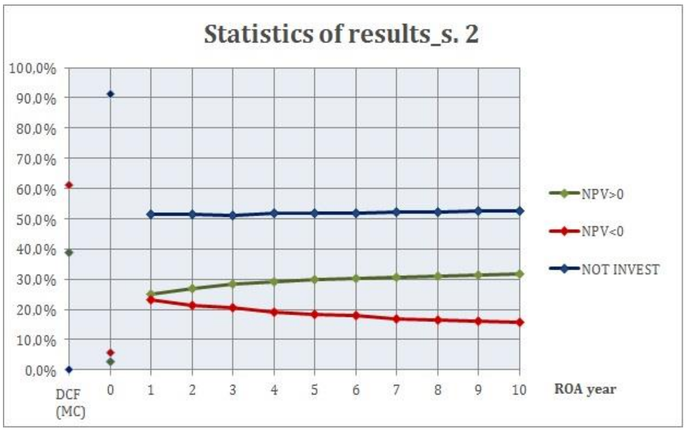


Scenario 2 – standard diesel

BIODIESEL [$/L]			BIOETHANOL [$/L]			ELECTRICITY [$/kWh]			DRIFT		
min	av	max	min	av	max	min	av	max	min	av	max
1,425	1,5	1,575	0,665	0,7	0,735	0,076	0,08	0,084	0%	2%	4%

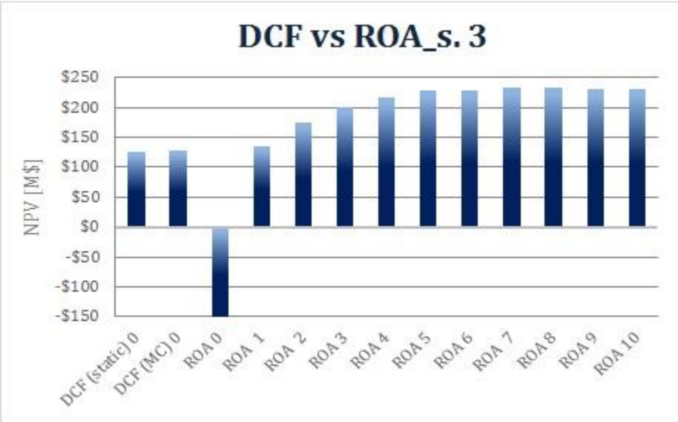


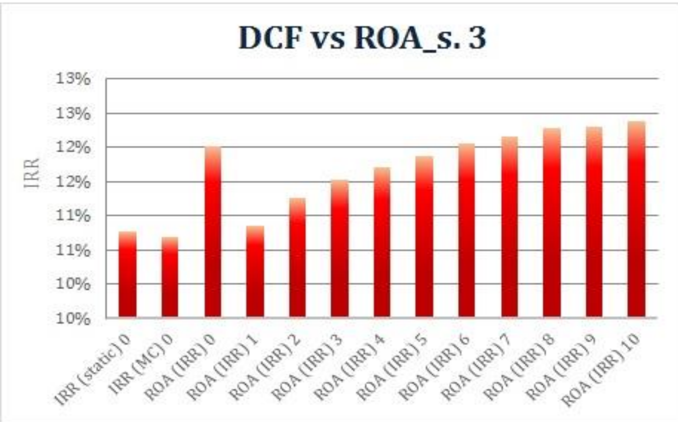


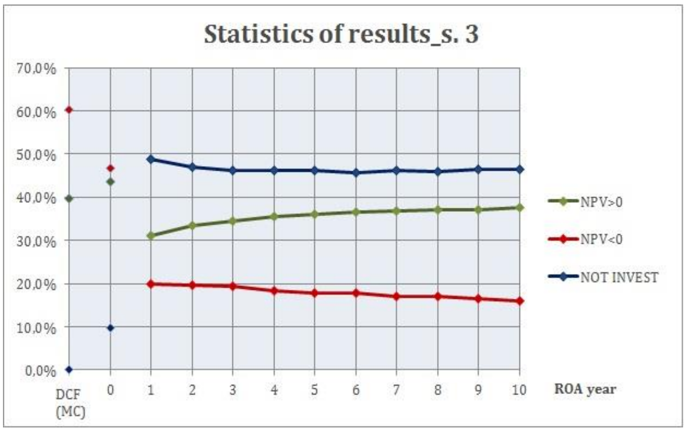


Scenario 3 – break-even price

BIODIESEL [$/L]			BIOETHANOL [$/L]			ELECTRICITY [$/kWh]			DRIFT		
min	av	max	min	av	max	min	av	max	min	av	max
1,52	1,6	1,68	0,665	0,7	0,735	0,076	0,08	0,084	0%	2%	4%

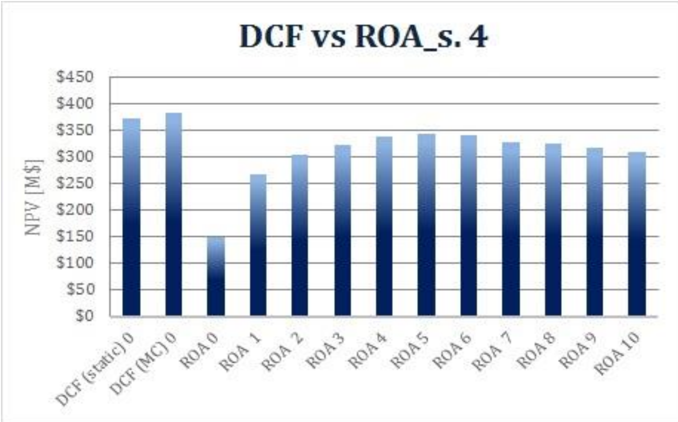


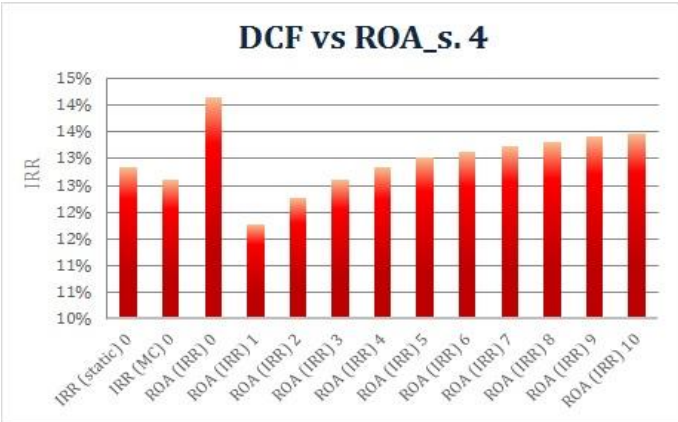


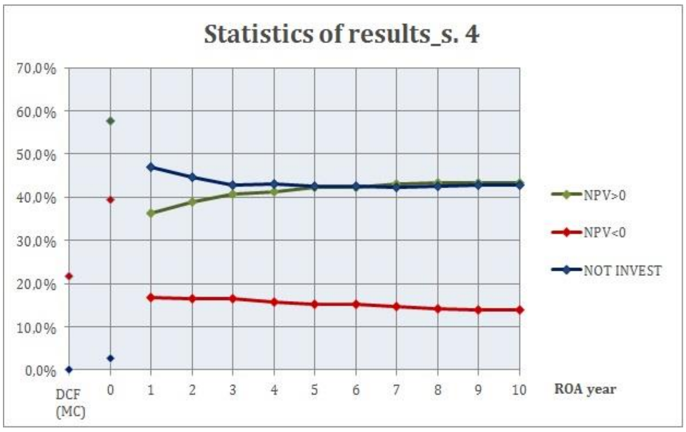


Scenario 4 – expensive diesel

BIODIESEL [$/L]			BIOETHANOL [$/L]			ELECTRICITY [$/kWh]			DRIFT		
min	av	max	min	av	max	min	av	max	min	av	max
1,615	1,7	1,785	0,665	0,7	0,735	0,076	0,08	0,084	0%	2%	4%

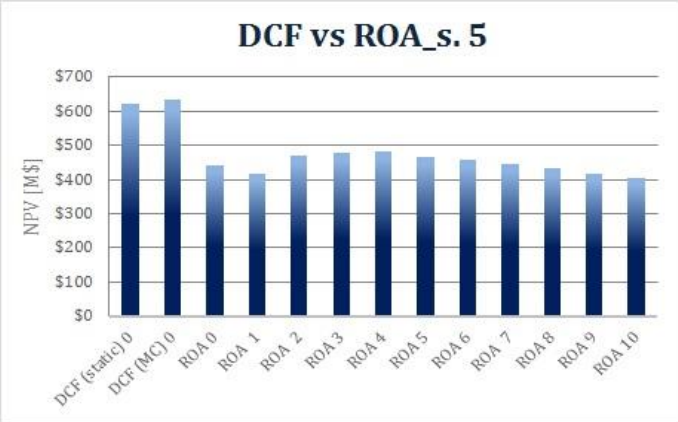


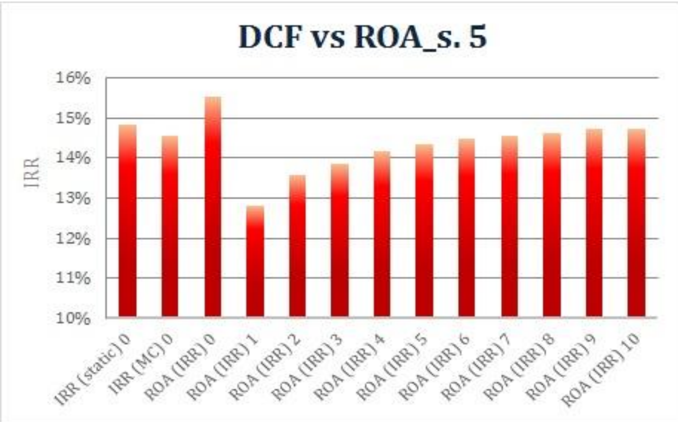


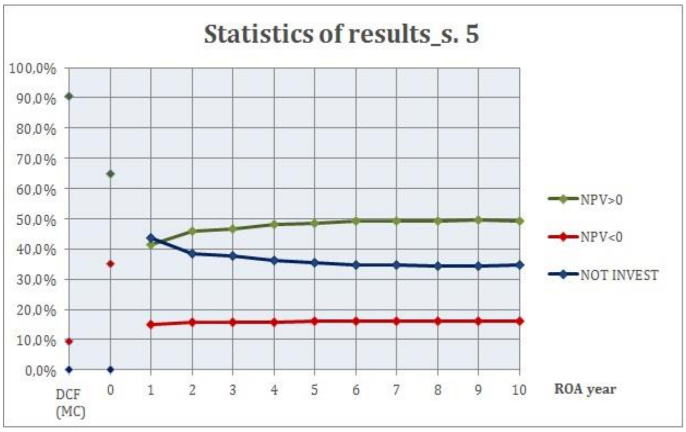


Scenario 5 – incentive

BIODIESEL [$/L]			BIOETHANOL [$/L]			ELECTRICITY [$/kWh]			DRIFT		
min	av	max	min	av	max	min	av	max	min	av	max
1,71	1,8	1,89	0,665	0,7	0,735	0,076	0,08	0,084	0%	2%	4%

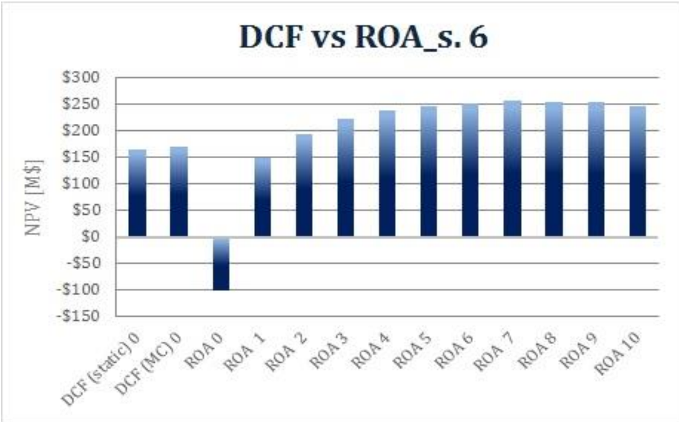


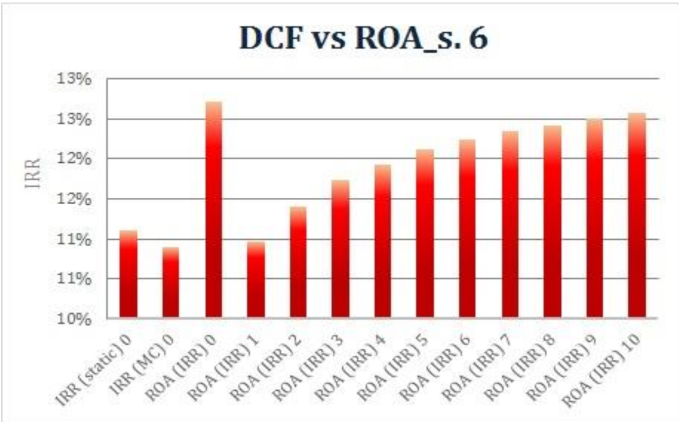


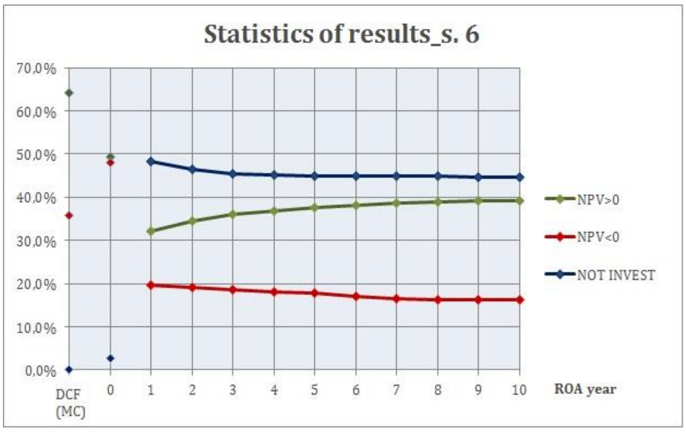


Scenario 6 – cheap electricity

BIODIESEL [$/L]			BIOETHANOL [$/L]			ELECTRICITY [$/kWh]			DRIFT		
min	av	max	min	av	max	min	av	max	min	av	max
1,425	1,5	1,575	0,665	0,7	0,735	0,067	0,07	0,074	0%	2%	4%

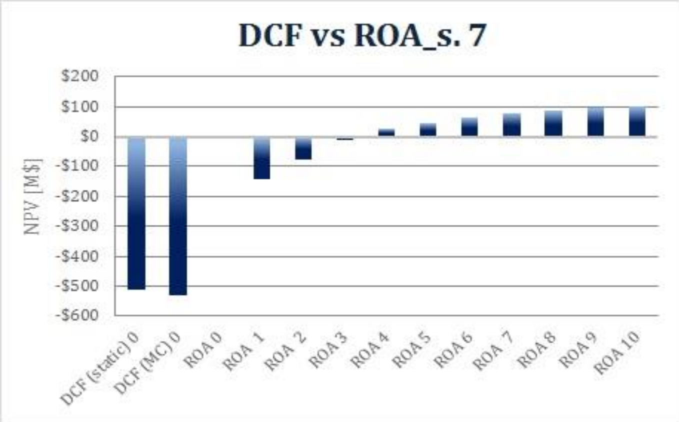


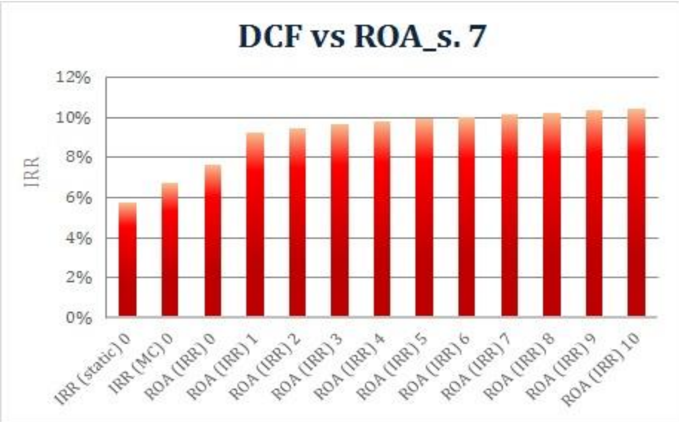


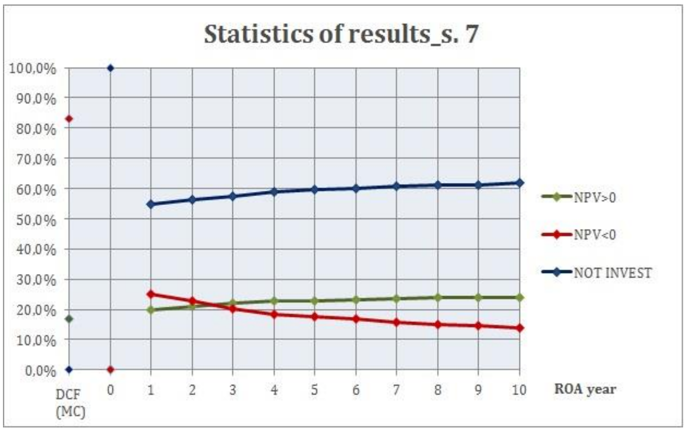


Scenario 7 – expensive electricity

BIODIESEL [$/L]			BIOETHANOL [$/L]			ELECTRICITY [$/kWh]			DRIFT		
min	av	max	min	av	max	min	av	max	min	av	max
1,425	1,5	1,575	0,665	0,7	0,735	0,086	0,09	0,095	0%	2%	4%





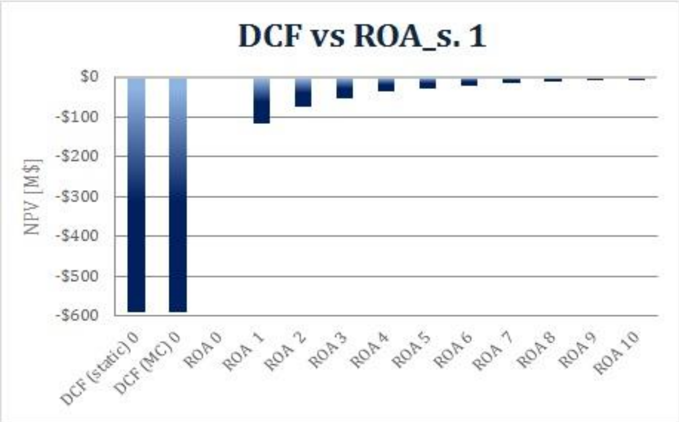


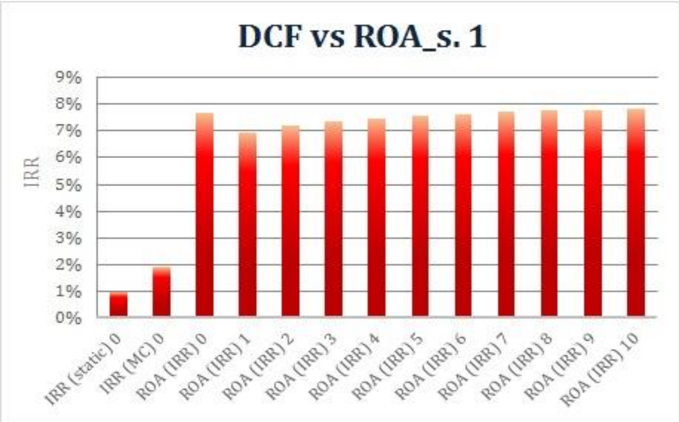
OPTION TO BUILD: DESALINATION

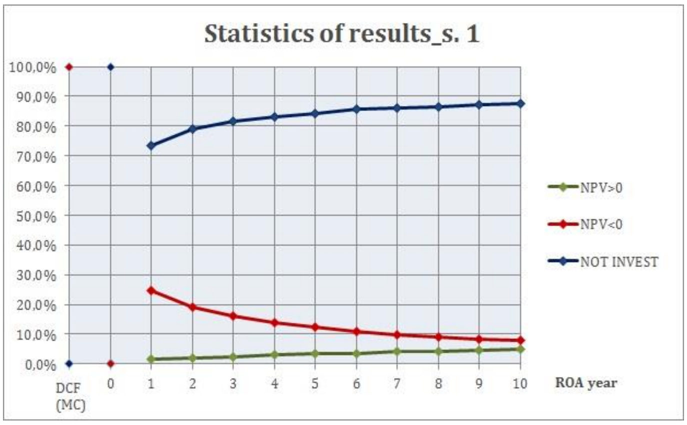
The same graphs of the Biorefinery case are reported even for the 7 scenarios hypothesised for the desalination case.

Scenario 1 – cheap water

WATER			ELECTRICITY. Night window		
min	av	max	min	av	max
1,14	1,2	1,26	0,038	0,04	0,042

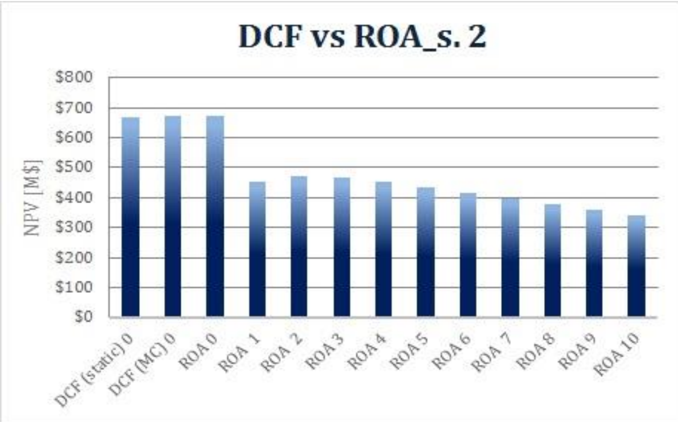


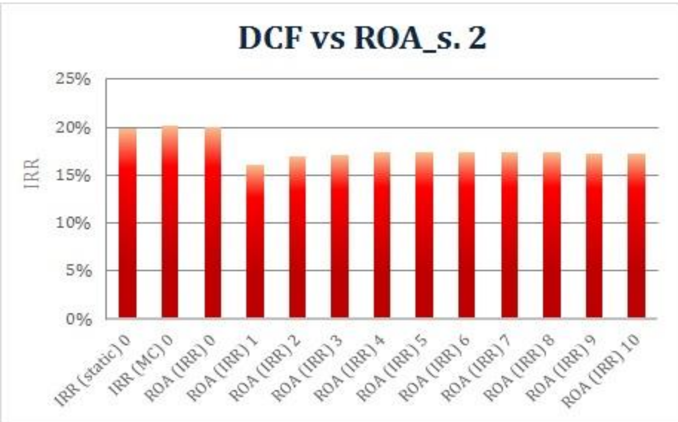


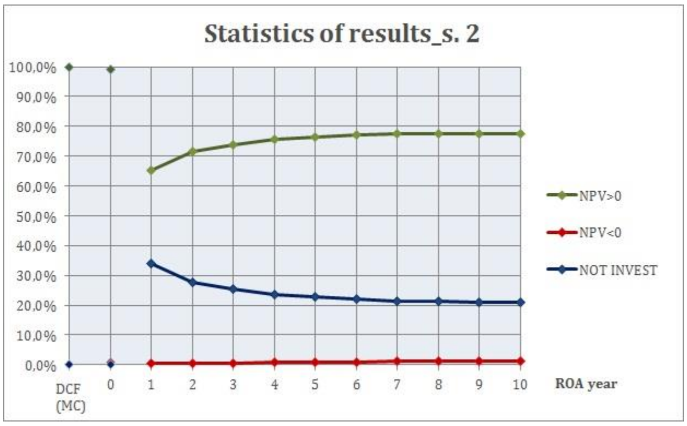


Scenario 2 – expensive water

WATER			ELECTRICITY. Night window		
min	av	max	min	av	max
2,38	2,5	2,63	0,038	0,040	0,042

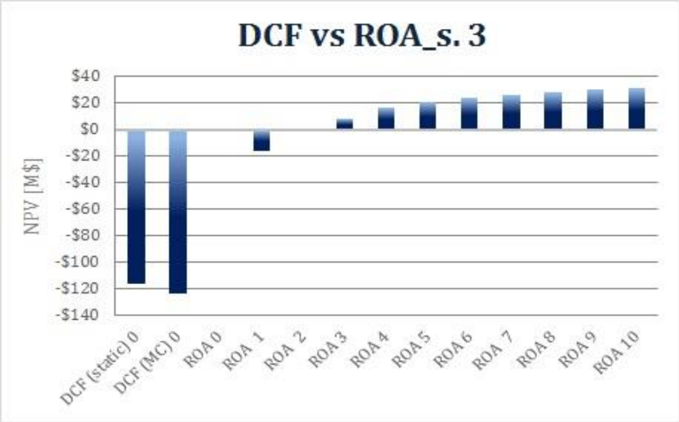


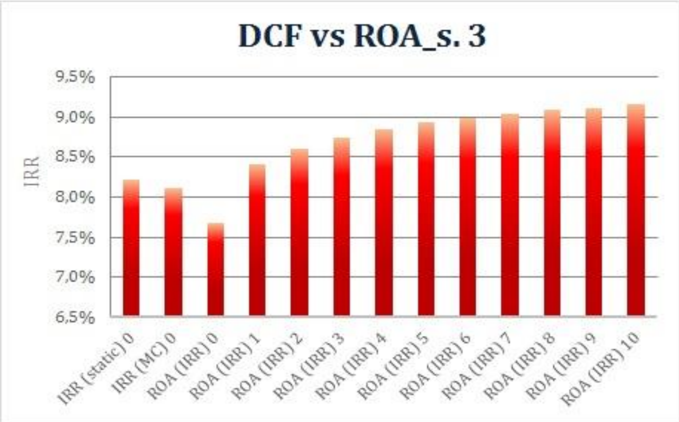


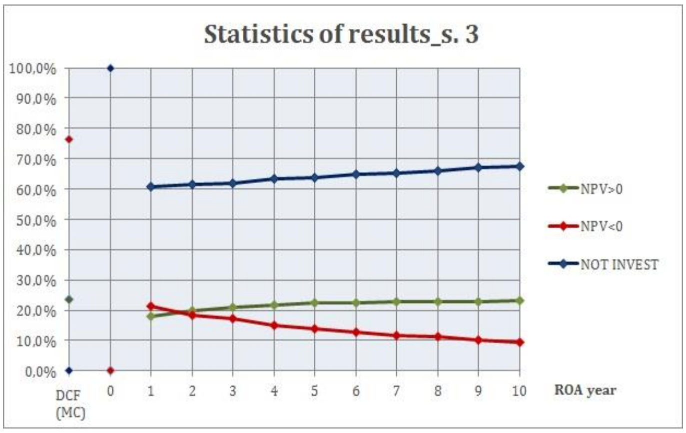


Scenario 3 – standard case

WATER			ELECTRICITY. Night window		
min	av	max	min	av	max
1,52	1,6	1,68	0,038	0,040	0,042

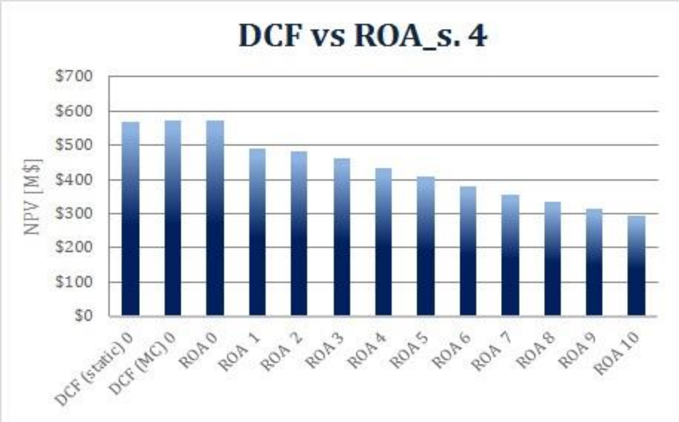


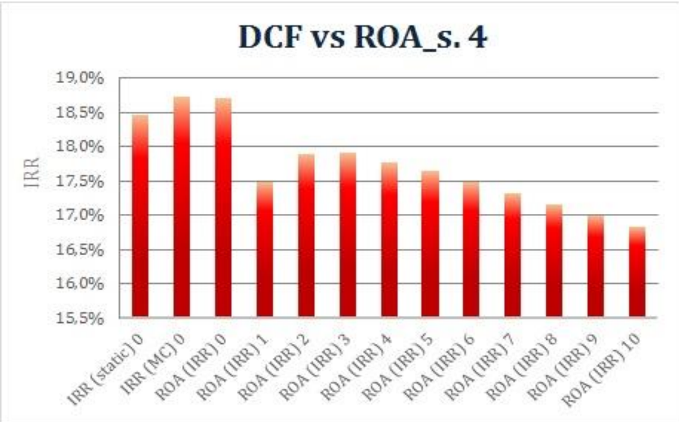


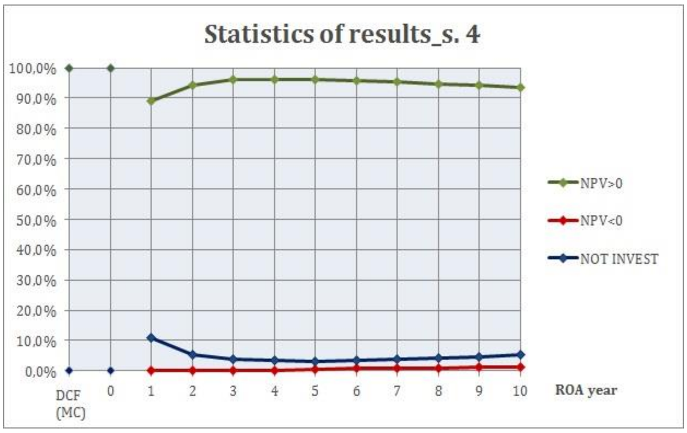


Scenario 4 – pure load following

WATER			ELECTRICITY. Night window		
min	av	max	min	av	max
1,52	1,6	1,68	0	0	0

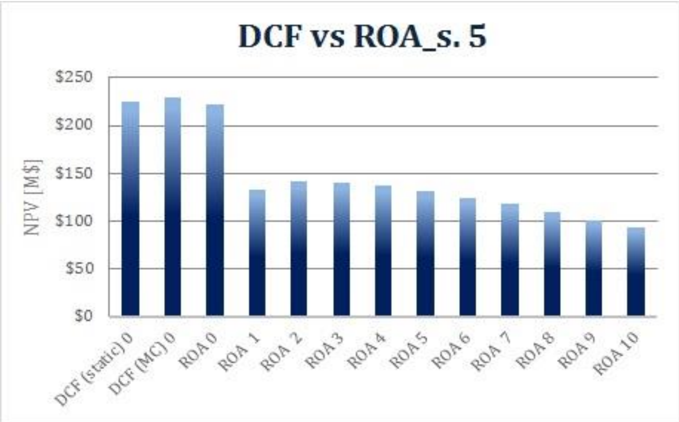


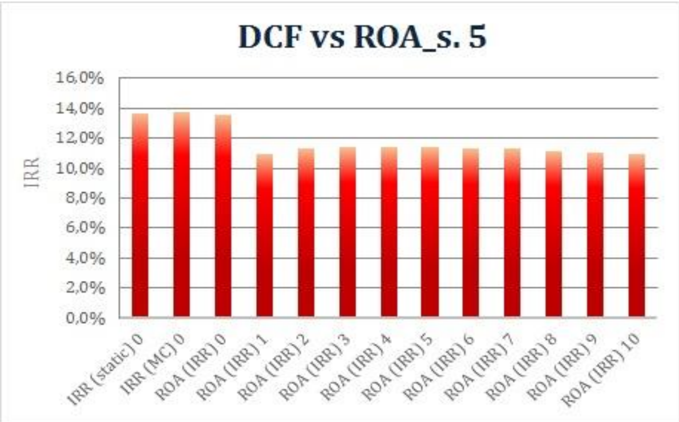




Scenario 5 – cheap electricity

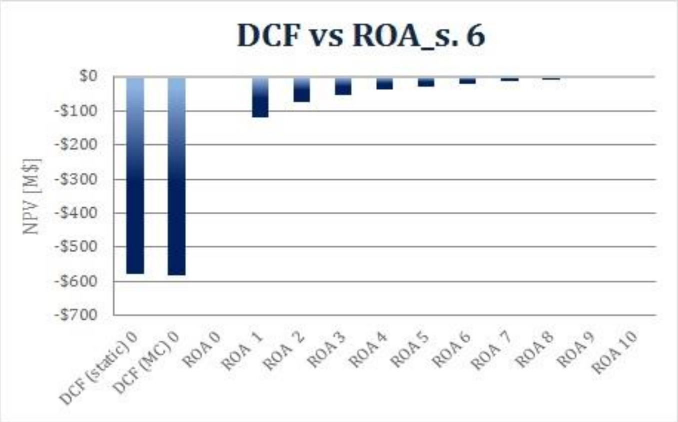
WATER			ELECTRICITY. Night window		
min	av	max	min	av	max
1,52	1,6	1,68	0,019	0,020	0,021

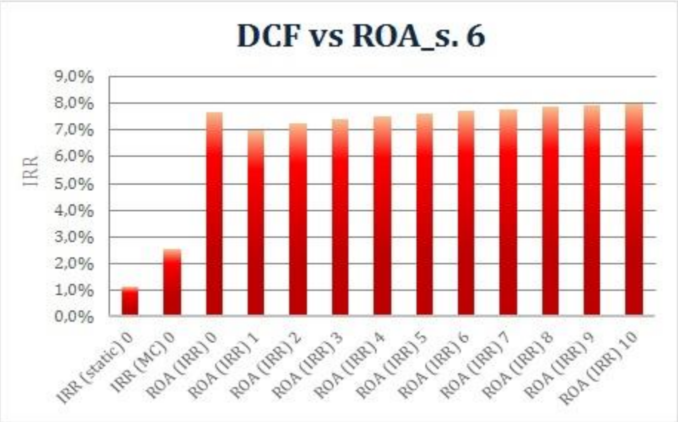


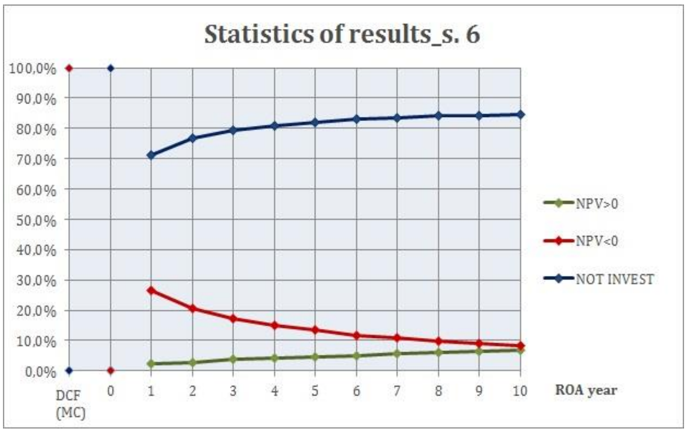


Scenario 6 – night price

WATER			ELECTRICITY. Night window		
min	av	max	min	av	max
1,52	1,6	1,68	0,057	0,060	0,063

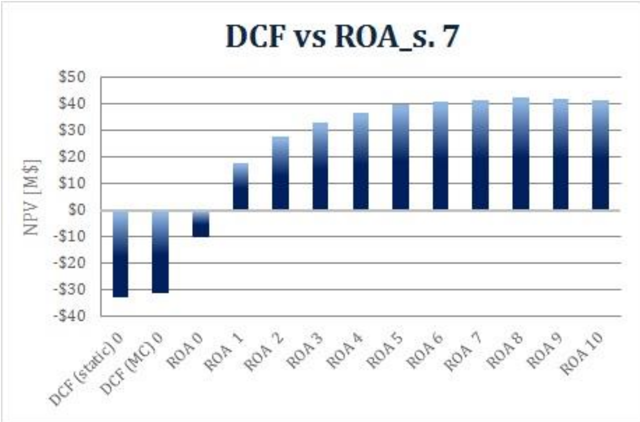


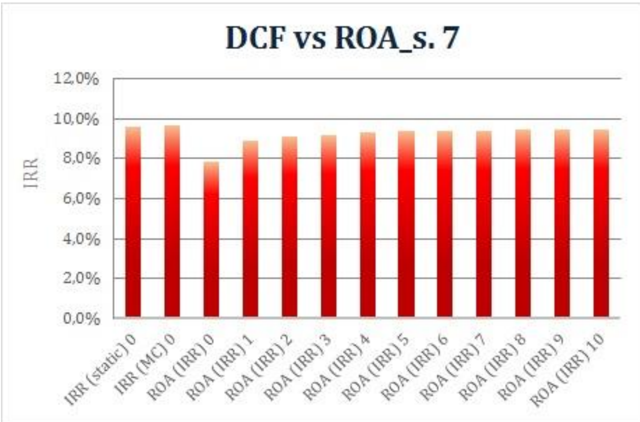


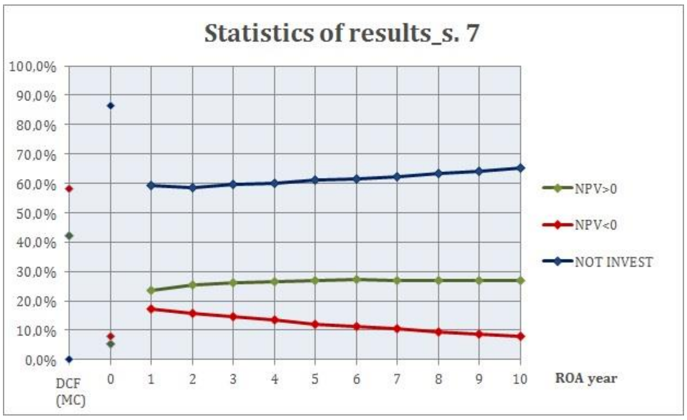


Scenario 7– breakeven case

WATER			ELECTRICITY. Night window		
min	av	max	min	av	max
1,43	1,50	1,58	0,029	0,030	0,032







OPTION TO SWITCH: DESALINATION

The 4 completed tables for the 100 simulations for the option to switch are inserted in this section of the appendix..

flexi-static	electricity price	water price [$/m3]																			
	@2.00 [$/kWh]	0,4	0,6	0,8	1	1,1	1,2	1,3	1,35	1,4	1,45	1,5	1,55	1,6	1,8	1,9	2	2,2	2,4	2,6	2,8
France	0	$448	$448	$451	$464	$478	$499	$528	$546	$568	$593	$621	$653	$687	$858	$966	$1.089	$1.380	$1.723	$2.096	$2.485
cheap	0,02	$285	$279	$282	$295	$308	$329	$359	$376	$398	$423	$451	$483	$518	$690	$797	$ 918	$1.211	$1.554	$1.928	$2.316
standard	0,04	$282	$203	$126	$125	$139	$159	$188	$207	$229	$254	$283	$314	$349	$520	$627	$ 750	$1.041	$1.384	$1.757	$2.147
night	0,06	$279	$200	$123	$ 56	$ 30	$ 11	$ 19	$ 38	$ 59	$ 84	$112	$144	$178	$351	$458	$ 581	$ 872	$1.215	$1.588	$1.977
market	-	$280	$201	$127	$ 73	$ 60	$ 64	$ 85	$101	$121	$146	$173	$204	$239	$411	$517	$ 640	$ 932	$1.275	$1.648	$2.037

ROA value	electricity price	water price [$/m3]																			
	@2.00 [$/kWh]	0,4	0,6	0,8	1	1,1	1,2	1,3	1,35	1,4	1,45	1,5	1,55	1,6	1,8	1,9	2	2,2	2,4	2,6	2,8
France	0	$160	$240	$322	$415	$469	$530	$598	$637	$678	$723	$735	$667	$602	$373	$282	$ 205	$ 97	$ 41	$ 16	$ 6
cheap	0,02	$ -	$ 73	$156	$249	$302	$362	$432	$470	$512	$557	$605	$657	$601	$375	$282	$ 204	$ 99	$ 42	$ 17	$ 6
standard	0,04	$ -	$ -	$ 3	$ 83	$136	$196	$265	$303	$346	$390	$439	$490	$545	$374	$282	$ 205	$ 97	$ 41	$ 16	$ 6
night	0,06	$ -	$ -	$ 3	$ 17	$ 30	$ 51	$ 99	$138	$179	$223	$271	$323	$378	$375	$282	$ 206	$ 98	$ 42	$ 17	$ 6
market	-	$ -	$ 1	$ 6	$ 32	$ 59	$103	$164	$200	$239	$284	$332	$382	$437	$374	$280	$ 205	$ 99	$ 42	$ 16	$ 6

NPP used	electricity price	water price [$/m3]																			
	@2.00 [$/kWh]	0,4	0,6	0,8	1	1,1	1,2	1,3	1,35	1,4	1,45	1,5	1,55	1,6	1,8	1,9	2	2,2	2,4	2,6	2,8
France	0	79,8%	79,7%	79,2%	77,8%	76,8%	75,4%	73,6%	72,6%	71,4%	70,2%	69,0%	67,8%	66,6%	61,9%	59,2%	56,4%	50,9%	47,2%	45,3%	44,4%
cheap	0,02	86,9%	79,7%	79,2%	77,8%	76,8%	75,4%	73,6%	72,6%	71,4%	70,2%	69,1%	67,8%	66,6%	61,9%	59,2%	56,3%	51,0%	47,2%	45,3%	44,4%
standard	0,04	86,9%	86,9%	86,4%	77,8%	76,7%	75,4%	73,6%	72,6%	71,4%	70,2%	69,0%	67,8%	66,6%	61,9%	59,2%	56,4%	50,9%	47,2%	45,3%	44,5%
night	0,06	86,9%	86,9%	86,4%	85,0%	83,9%	82,6%	73,6%	72,6%	71,4%	70,2%	69,1%	67,8%	66,7%	61,9%	59,2%	56,4%	50,9%	47,2%	45,3%	44,5%
market	-	86,9%	86,8%	85,9%	83,1%	80,6%	77,5%	74,5%	73,2%	71,8%	70,5%	69,2%	67,9%	66,7%	61,9%	59,2%	56,3%	50,9%	47,2%	45,3%	44,4%

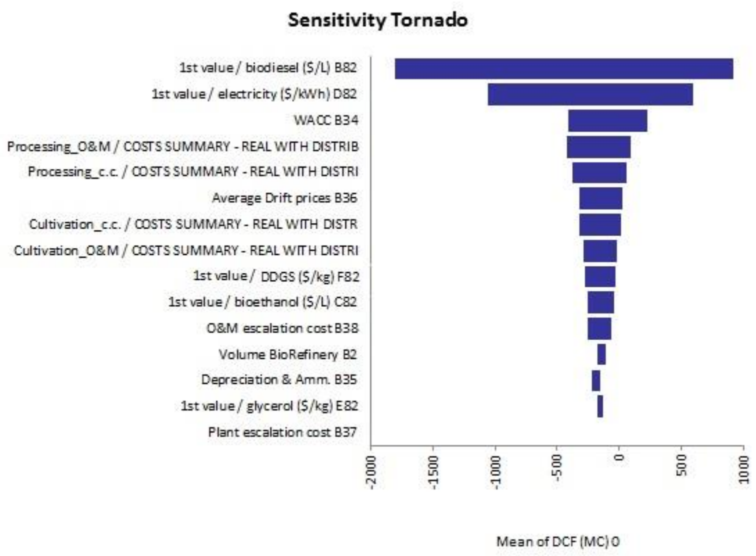
MED used	electricity price	water price [$/m3]																			
	@2.00 [$/kWh]	0,4	0,6	0,8	1	1,1	1,2	1,3	1,35	1,4	1,45	1,5	1,55	1,6	1,8	1,9	2	2,2	2,4	2,6	2,8
France	0	37,5%	37,6%	38,5%	40,9%	42,7%	45,2%	48,2%	50,1%	52,1%	54,1%	56,2%	58,4%	60,4%	68,7%	73,3%	78,3%	87,8%	94,2%	97,5%	99,0%
cheap	0,02	25,0%	37,6%	38,5%	40,9%	42,8%	45,1%	48,2%	50,0%	52,1%	54,2%	56,2%	58,4%	60,4%	68,7%	73,3%	78,3%	87,7%	94,2%	97,5%	99,1%
standard	0,04	25,0%	25,1%	26,0%	40,9%	42,8%	45,1%	48,2%	50,0%	52,1%	54,1%	56,2%	58,3%	60,4%	68,7%	73,2%	78,3%	87,8%	94,2%	97,5%	99,0%
night	0,06	25,0%	25,1%	26,0%	28,4%	30,3%	32,6%	48,2%	50,1%	52,1%	54,1%	56,2%	58,3%	60,3%	68,6%	73,2%	78,3%	87,8%	94,2%	97,5%	99,0%
market	-	25,0%	25,2%	26,9%	31,7%	36,0%	41,5%	46,6%	49,0%	51,3%	53,7%	55,9%	58,2%	60,2%	68,6%	73,3%	78,3%	87,8%	94,2%	97,6%	99,0%

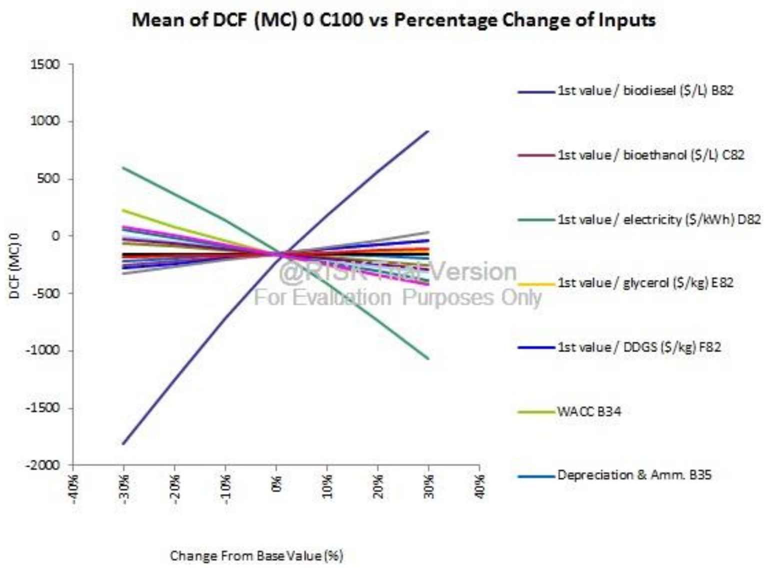
SENSITIVITY ANALYSIS

In this section of the Appendix all the graphs for the sensitivity analysis of the model are reported. At first the graphs regarding the biorefinery case are inserted, then there are the same graphs regarding the desalination. For both cases, 3 different sensitivity analysis are done: 2 are for the DCF Monte Carlo model, one for the Real Option at year 5.

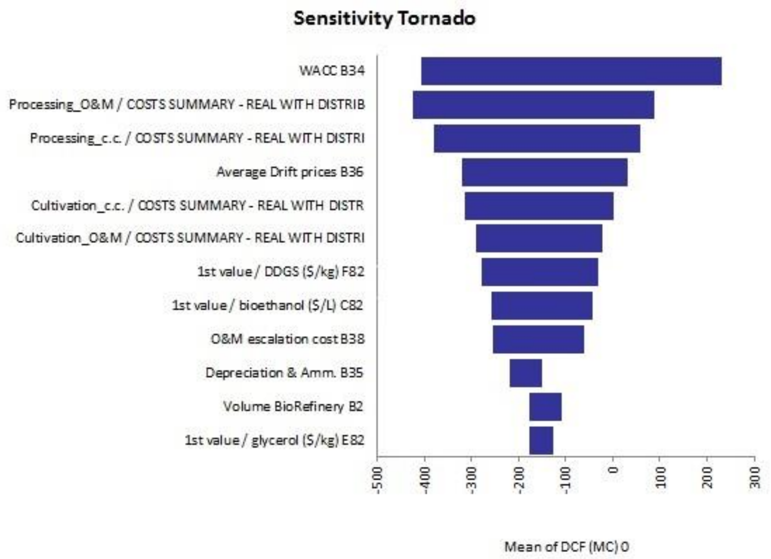
Biorefinery

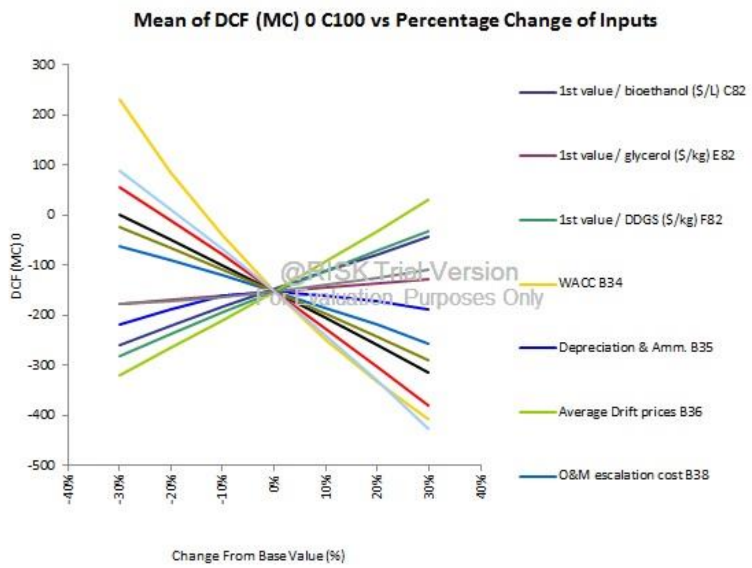
DCF MC, General



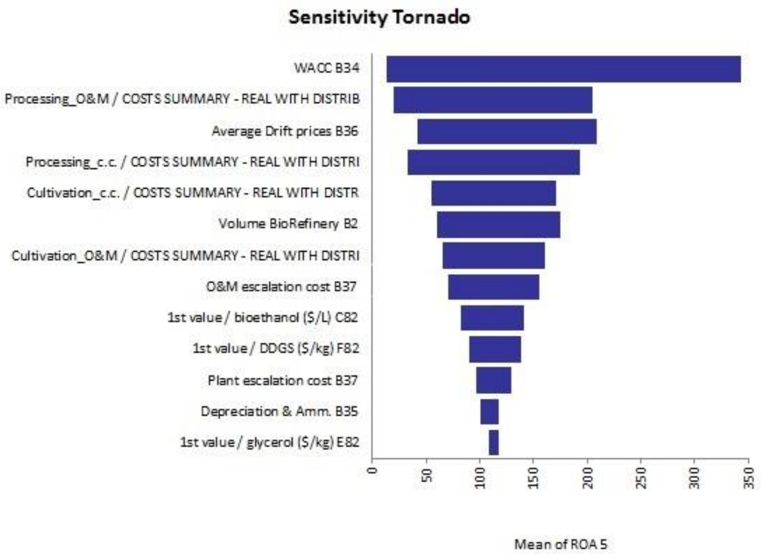


DCF MC, Zoom without biodiesel and E.E.

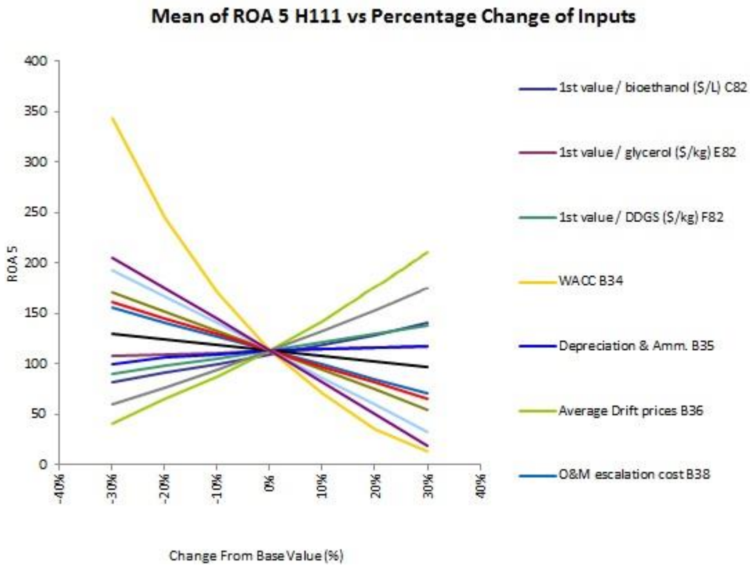




ROA year 5



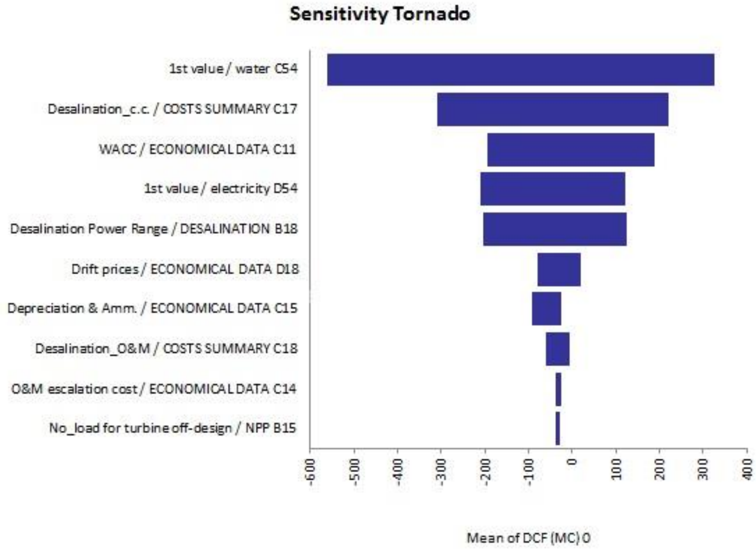
Sensitivity Tornado
WACC B34
Processing_O&M / COSTS SUMMARY - REAL WITH DISTRIB
Average Drift prices B36
Processing_c.c. / COSTS SUMMARY - REAL WITH DISTRI
Cultivation_c.c. / COSTS SUMMARY - REAL WITH DISTR
Volume BioRefinery B2
Cultivation_O&M / COSTS SUMMARY - REAL WITH DISTRI
O&M escalation cost B37
1st value / bioethanol ($/L) C82
1st value / DDGS ($/kg) F82
Plant escalation cost B37
Depreciation & Amm. B35
1st value / glycerol ($/kg) E82
0
50
100
150
200
250
300
350
Mean of ROA 5

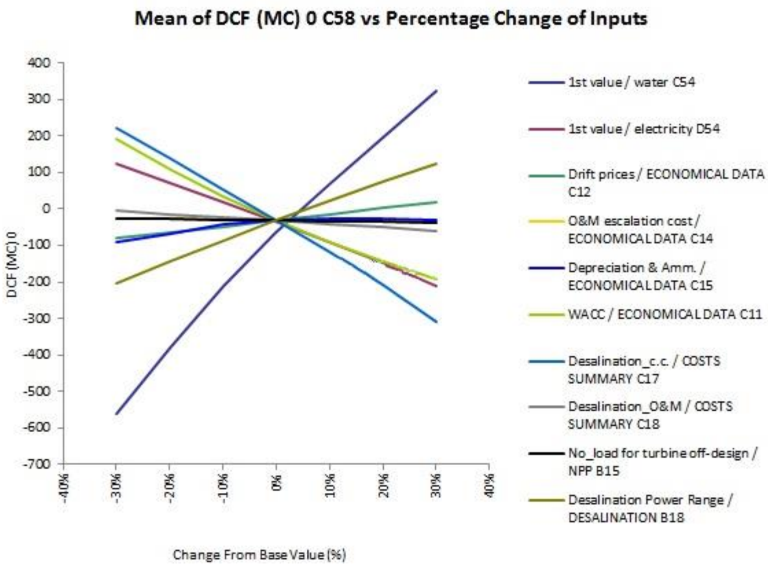


Mean of ROA 5 H111 vs Percentage Change of Inputs
400
350
300
250
200
150
100
50
0
ROA 5
-40%
-30%
-20%
-10%
0%
10%
20%
30%
40%
Change From Base Value (%)
1st value / bioethanol ($/L) C82
1st value / glycerol ($/kg) E82
1st value / DDGS ($/kg) F82
WACC B34
Depreciation & Amm. B35
Average Drift prices B36
O&M escalation cost B38

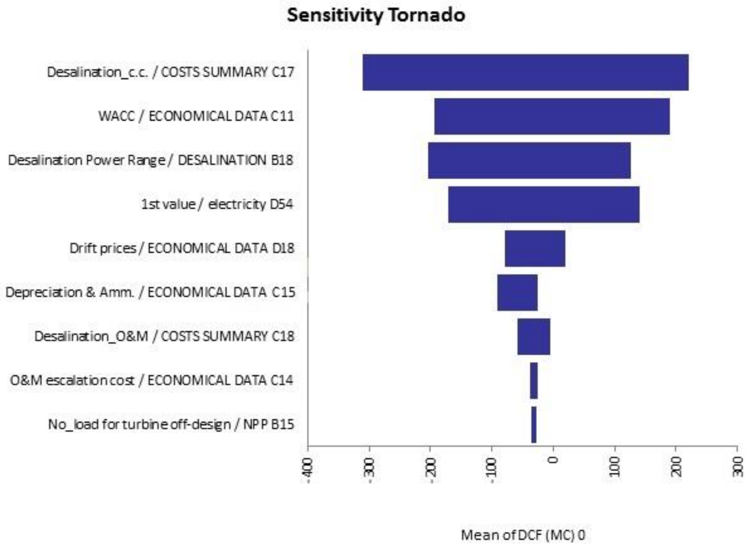
Desalination

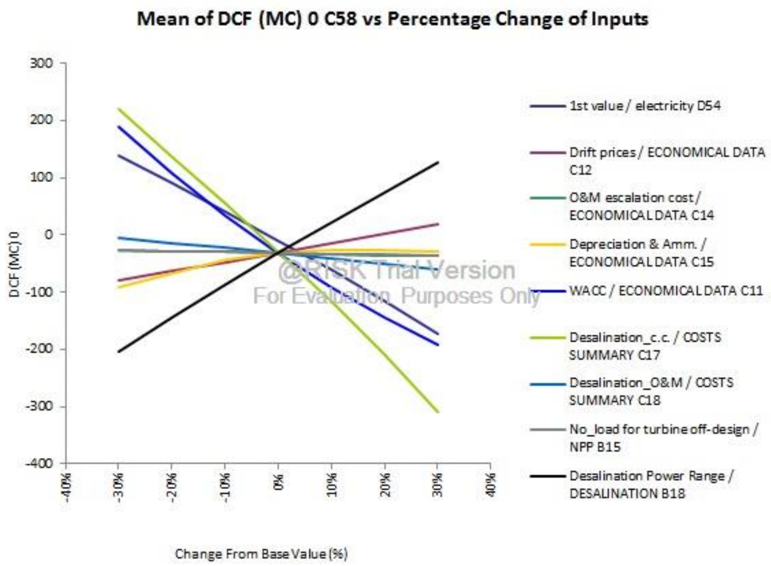
DCF MC, General



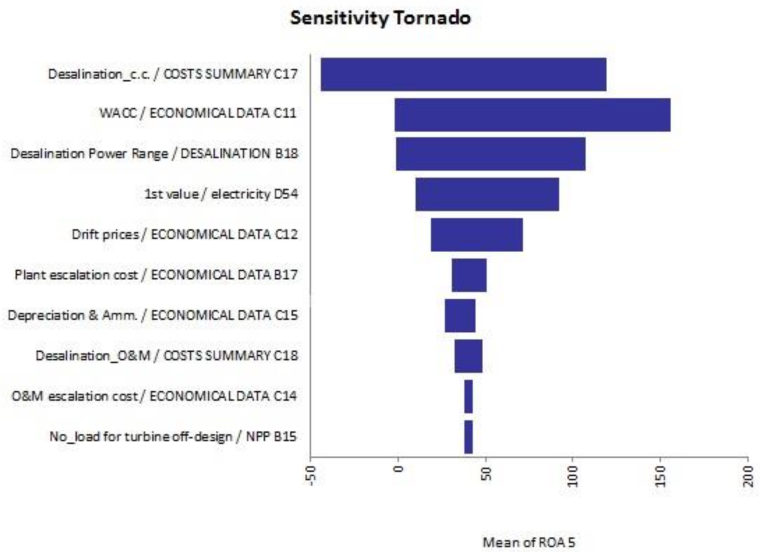


DCF MC, Zoom without water

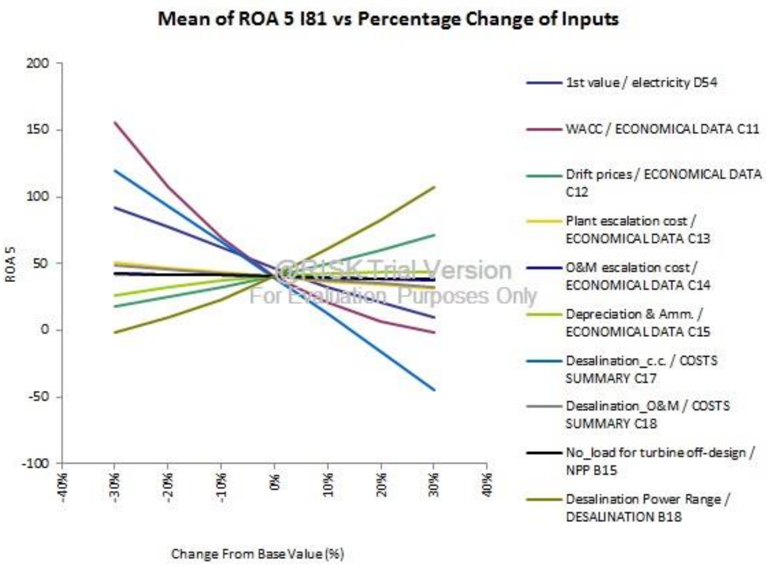




ROA year 5



Sensitivity Tornado
Desalination_c.c. / COSTS SUMMARY C17
WACC / ECONOMICAL DATA C11
Desalination Power Range / DESALINATION B18
1st value / electricity D54
Drift prices / ECONOMICAL DATA C12
Plant escalation cost / ECONOMICAL DATA B17
Depreciation & Amm. / ECONOMICAL DATA C15
Desalination_O&M / COSTS SUMMARY C18
O&M escalation cost / ECONOMICAL DATA C14
No_load for turbine off-design / NPP B15
-50
0
50
100
150
200
Mean of ROA 5



Mean of ROA 5 I81 vs Percentage Change of Inputs
200
150
100
50
0
-50
-100
ROA 5
-40%
-30%
-20%
-10%
0%
10%
20%
30%
40%
Change From Base Value (%)
@RISK Trial Version
For Evaluation Purposes Only
1st value / electricity D54
WACC / ECONOMICAL DATA C11
Drift prices / ECONOMICAL DATA C12
Plant escalation cost / ECONOMICAL DATA C13
O&M escalation cost / ECONOMICAL DATA C14
Depreciation & Amm. / ECONOMICAL DATA C15
Desalination_c.c. / COSTS SUMMARY C17
Desalination_O&M / COSTS SUMMARY C18
No_load for turbine off-design / NPP B15
Desalination Power Range / DESALINATION B18

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