Extraction of astaxanthin from microalgae: process design and economic feasibility study

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Abstract. In this work, the process design and the economic feasibility of natural astaxanthin extraction from Haematococcus pluvialis species have been reported. Complete process drawing of the process was first performed, and then the process was designed including five main steps being the harvesting process, the cell disruption, the spray drying, the supercritical CO\textsubscript{2} extraction and the anaerobic digestion. The major components of the facility would include sedimentation tanks, a disk stack centrifuge, a bed miller, a spray dryer, a multistage compressor, an extractor, a pasteurizer and a digester. All units have been sized assuming a 10 kg/h of dried biomass as a feedstock to produce nearly 2592 kg of astaxanthin per year. The investment payback time and the return on investment were all estimated for different market prices of astaxanthin. Based on the results the production process was found to become economically feasible for a market price higher than 1500$/Kg. Also, a payback period of 1 year and an ROI equal to 113% was estimated for an astaxanthin market price equal to 6000$/Kg.

1. Introduction
The potential of photosynthetic microorganisms is presented in different domains such as food, feed, pharmaceutical, cosmetic and fuel industries. Nowadays, microalgae are under the spot since they constitute a renewable resource. Cultivating this bioresource would solve some of the limitations in the future of traditional biomass production and markets. These microorganisms are known for their high growth rate and biomass productivity compared to vegetables [1]. Recently, microalgae have become the focus of the research on photosynthetic microorganisms due to several reasons. Firstly, these microorganisms are tiny cells that possess different types of pigments responsible of the photosynthesis process. Throughout their growth, pigments, proteins, lipids, and carbohydrates are formed and accumulated in their cells. To grow, they need light, carbon dioxide, nutrients and water or waste water. Consequently, they can be used to reduce the carbon dioxide present in the atmosphere and to lower the concentrations of nitrogen and phosphorus in waste water. Further, due to the limitation in fossil fuels that the world could face, microalgae could be a promising solution for the production of biofuel [2]. The main feature of growing microalgae is the natural commercial extracts obtained after processing them. Of the natural extracts produced from the microalgae Haematococcus pluvialis, astaxanthin have received the most attention. The most common use of commercially produced astaxanthin is in fish farming where it imparts coloration to farmed salmonids and crustaceans [3]. Also, due to its antioxidant activity and fortification astaxanthin has been widely used...
in the pharmaceutical and nutraceutical industry [4]. Nowadays, the commercial astaxanthin is dominated by the synthetic one since the market prices for natural astaxanthin are higher compared to the synthetic version [5, 6]. However, growing consumer demand in the health food sector for natural products has given the opportunity for production of astaxanthin from natural sources. The production of natural astaxanthin being an expensive process, the objective of this work is to develop a new bioprocess to produce natural astaxanthin in a low cost location such as Lebanon and using the available technologies. In this study, we will not include the cultivation process of the biomass rather we will focus on the astaxanthin extraction process. We will assume that the biomass suspension rich in astaxanthin is supplied from a platform producing the microalgae *Haematococcus pluvialis* on a large scale. The purpose of this paper is to explore whether natural astaxanthin production in Lebanon is economically feasible and if the application of the process on an industrial scale could be recommended. This study will be divided into three parts: process synthesis, process design and economic analysis.

2. Description of the production process

The production facility will be theoretically located in Bekaa (Lebanon), a place with considerable sunshine time. The biomass suspension flow rate was estimated to be 10 kg/h of dried biomass as a feedstock to produce nearly 2592 kg of astaxanthin per year. The most important mass inflow refers to freshwater and the suspended microalgae concentration in the inlet flow mixture was considered to be 1 kg/m³. The main processing units would include two sedimentation tanks, a disk stack centrifuge, a bed miller, a spray dryer, a multistage compressor, an extractor, a pasteurizer and a digester. A continuous process flow diagram of the astaxanthin production was developed as shown in figure 1. According to this study, the harvesting process was performed in a two-step approach employing two gravity sedimentation tanks V-101/102 as a first step followed by a disk-stack centrifugation V-103. Large amount of water was eliminated by sedimentation followed by centrifugation where the concentrated slurry was further dewatered to a cake consisting of 21% of the total suspended solid (TTS). Then, the cake was introduced into a bead miller V-104 in order to disrupt the cell and release the compounds. After harvesting and cell disruption a spray dryer V-105/106 was used to remove the moisture left in the biomass and to increase its shelf life. During this process the algal slurry is dried to 5% moisture content. For the recovery of astaxanthin from *Haematococcus pluvialis*, supercritical CO₂ extraction was used. Throughout this process, the biomass or the paste was entered into the extractor column T-101 where it was mixed with the supercritical CO₂ fluid and the ethanol as a co-solvent at a temperature of 60°C and a pressure of 30 MPa. Since the gas holder tank TK-101 would contain CO₂ gas at 45 bars and 20°C a two-stage compressor C-101 and C-102 and an intercooler E-101 were used to increase the pressure of the CO₂ to 30 MPa. Then, a cooler E-102 was used to cool the CO₂ supercritical fluid to 60°C before entering the extraction column T-101. The outlet extract stream was passed through a cyclone V-107 where the astaxanthin was separated from the CO₂ and collected in the tank TK-103. The fluid CO₂ that leaves the cyclone was recycled back to the multistage compressor after being compressed C-103 to 45 bars. The residual biomass (Stream 22) left in the extraction chamber was heated in exchanger E-201 to 65°C before entering the pasteurizer E-202. In the pasteurizer the pasteurization of the waste solids was retained at 70°C for 30 minutes to reduce the pathogen levels. The stream 27 was then cooled in exchanger E-203 before entering the anaerobic digester R-201 where the waste was transformed into biofertilizer and biogas. Material balances and process parameters of the astaxanthin production from *Haematococcus pluvialis* were summarized in table 1.
Figure 1. Process flow diagram of the astaxanthin production from *Haematococcus pluvialis*.
| Stream N° | Temperature (°C) | Pressure (bar) | Algae suspension (kg/h) | Air (kg/h) | CO₂ (kg/h) | Ethanol (kg/h) | Astaxanthin (kg/h) | Byproducts (kg/h) | Biogas (m³/h) | Fertilizer (kg/h) |
|----------|------------------|----------------|------------------------|------------|------------|----------------|-------------------|------------------|---------------|------------------|
| 1        | 20               | 1              | 11,000                 | -          | -          | -              | -                 | -                | -              | -                |
| 2        | 20               | 1              | 750                    | -          | -          | -              | -                 | -                | -              | -                |
| 3        | 20               | 1              | 10,250                 | -          | -          | -              | -                 | -                | -              | -                |
| 4        | 20               | 1              | 703                    | -          | -          | -              | -                 | -                | -              | -                |
| 5        | 20               | 1              | 46.8                   | -          | -          | -              | -                 | -                | -              | -                |
| 6        | 20               | 1              | 46.8                   | -          | -          | -              | -                 | -                | -              | -                |
| 7        | 35               | 1              | 12,991                 | -          | -          | -              | -                 | -                | -              | -                |
| 8        | 20               | 1              | 11.8                   | -          | -          | -              | -                 | -                | -              | -                |
| 9        | 20               | 45             | 28                     | -          | -          | -              | -                 | -                | -              | -                |
| 10       | 20               | 45             | 561                    | -          | -          | -              | -                 | -                | -              | -                |
| 11       | 144              | 173.5          | 561                    | -          | -          | -              | -                 | -                | -              | -                |
| 12       | 76               | 172.5          | 561                    | -          | -          | -              | -                 | -                | -              | -                |
| 13       | 110              | 301            | 561                    | -          | -          | -              | -                 | -                | -              | -                |
| 14       | 60               | 300            | 561                    | -          | -          | -              | -                 | -                | -              | -                |
| 15       | 20               | 1              | 31.56                  | -          | -          | -              | -                 | -                | -              | -                |
| 16       | 20               | 1              | 31.56                  | -          | -          | -              | -                 | -                | -              | -                |
| 17       | 60               | 300            | 561                    | -          | -          | -              | -                 | -                | -              | -                |
| 18       | 20               | 1              | 31.56                  | -          | -          | -              | -                 | -                | -              | -                |
| 19       | 20               | 1              | 11.5                   | -          | -          | -              | -                 | -                | -              | -                |
| 20       | 60               | 300            | 533                    | -          | -          | -              | -                 | -                | -              | -                |
| 21       | 20               | 1              | -                      | -          | -          | -              | -                 | -                | -              | -                |
| 22       | 40               | 1              | -                      | -          | -          | -              | -                 | -                | -              | -                |
| 23       | 20               | 1              | 533                    | -          | -          | -              | -                 | -                | -              | -                |
| 24       | 20               | 1              | 0.3                    | -          | -          | -              | -                 | -                | -              | -                |
| 25       | 20               | 45             | 533                    | -          | -          | -              | -                 | -                | -              | -                |
| 26       | 65               | 1              | -                      | -          | -          | -              | -                 | -                | -              | -                |
| 27       | 70               | 1              | -                      | -          | -          | -              | -                 | 11.5             | -              | -                |
| 28       | 30               | 1              | -                      | -          | -          | -              | -                 | 11.5             | -              | -                |
| 29       | 37               | 1              | -                      | -          | -          | -              | -                 | 6                | -              | -                |
| 30       | 37               | 1              | -                      | -          | -          | -              | -                 | -                | -              | -                |

3. Construction of the process

3.1. Harvesting phase

To date, there is no optimal way of harvesting microalgae. The choice of the best process depends on microalga cell densities, algae species, the growth rates of microalgae, the acceptable level of moisture in the product and the production cost [7–11]. Taking into consideration the average size of a red cell of *Haematococcus pluvialis* (i.e. 20 μm) and the concentration of suspended microalgae (i.e. 0.1% dry biomass) a two-step approach was used for the harvesting process where the disk-stack centrifugation is combined with the gravity sedimentation technique. Centrifugation is considered an efficient and reliable technique for microalgal recovery especially for concentrating algal slurries from 1% to 5% solids to >15% solids. However, this technique is too cost and energy intensive to be applied alone, therefore, it is recommended as a second stage dewatering technique after a gravity sedimentation step [8, 12, 13, 14]. In this study two sedimentation tanks with a cylindrical shape were...
considered. Each day the slurry from the algal culture facility would be pumped to these tanks. The sedimentation tanks were sized using Stokes law to allow the maximum possible collection efficiency for a residence time of 1 day. To further increase the biomass recovery (>95%) a disk-stack centrifuge with a relative centrifugal force of 13000 g is used. For mass balance calculations the biomass recovery efficiency was considered to be 95%. The disk-stack centrifuge was sized using the Σ concept; Σ being the equivalent size of the tank in m².

3.2. Cell disruption phase
Bead milling is a mechanical method already used in industries for cell disruption and have been successfully applied for microalgae [15]. By colliding with each other, the beads with a spherical form and a diameter ranging from 1 to 5 mm will trap and stress the biomass cell which result in breaking the algal cell and releasing the compounds. The bead milling process is a complex operation with many parameters that interact with each other. This interaction is difficult to predict and vary in function of microorganisms, medium viscosity and flow rate. The main operating parameters are the bead filling, bead density, bead diameter, the speed of the agitation and the feed rate [16]. Taking into consideration the high viscosity of the slurry entering the bead miller, ceramic beads specifically alumina (ρ = 3.8 kg/m³) were used in this study [17]. The optimal bead diameter and the volume fraction of bead loading were obtained from the literature and they were respectively: 0.5 mm and 60% [6]. Additional factors contributing to the milling performance such as the stress number (SN) and the stress intensity (SI) were calculated using the stress model developed by Kwade and Schwedes [18]. This model which is based on stress energies and bead collision frequencies in the field of microorganisms describes the cell disruption process using two numbers: SN, the stress number which is the number of media contact and SI, the stress intensity of these events. The SN and the SI can be calculated as:

\[ SN = \frac{\varphi_{GM}(1-C)}{(1-\varphi_{GM}(1-C))C_v} \times \frac{n \times t}{d_{GM}} \]  

\[ SI = SI_{GM} = d_{GM}^3 \times \rho_{GM} \times v_t^2 \]  

Where ϕ_GM is the grinding media filling ratio, ε is the bulk porosity, d_GM is the bead diameter (m), C_v is the cell concentration (kg/m³), ρ_GM is the grinding media density: alumina (kg/m³), v_t is the grinding media velocity (m/s), n is the agitator speed (revolutions) (m/s) and t is the comminution time: time required to disrupt the cells. For mass balance calculations the biomass recovery efficiency was considered to be 100%.

3.3. Drying phase
In order to have high efficiency, the biomass must be dried first, since the moisture can decrease the yield of the supercritical CO₂ extraction by playing the role of a barrier facing the diffusion of the CO₂ into the biomass. The mass flow rate of the air entering the dryer was determined from the material balance performed on the dryer and its value is equal to 12,991 kg air/h. The evaporative rate which is the rate of water evaporated and removed corresponds to 26,125 kg/h and the moisture content in the biomass corresponds to 5%. The residence time in the evaporator was specified to be 2 minutes according to heuristics [19]. The dryer and the cyclone were designed to be a cylindrical chamber with a diameter similar to their height and a 60° conical bottom.

3.4. Extraction phase
As soon as the biomass is fully dried the recovery of the astaxanthin is possible. Supercritical CO₂ extraction is widely used to extract high value products from microalgae such as astaxanthin [20, 21]. The fundamental principle behind this method is the use of a supercritical fluid whose properties combine the solvation property of liquids with the high diffusivity of gases [22]. Carbon dioxide is an ideal component for this type of process. Besides being nontoxic, cheap and inert it possesses low critical temperature and pressure (31.1°C and 7.4 MPa). Also, under normal temperature and pressure
CO₂ is a gas which allows the elimination of the solvent after the extraction [20, 23, 24]. In furtherance of enhancing the solvating characteristic of CO₂ the extraction is usually performed using CO₂ as fluid above its critical points at a temperature of 60°C and a pressure of 30 MPa while using ethanol as co-solvent [25]. The rate of CO₂ used in this study is 0.2 kg/min/kg of dry microalgae and the percentage volume of ethanol/CO₂ is 4%. The supercritical CO₂ extraction performed in our study is composed of several units: a CO₂ gas holder tank, a two-stage compressor, a single stage compressor, two heat exchangers, a pump, an extractor column and a cyclone. The two-stage compressor is used to draw the CO₂ off a storage tank and boost the pressure to 30 MPa. Two-stages were used in this study since the compression ratio is equal to 6.66. The compressor utilized is a reciprocating compressor with an efficiency of 80% [26]. The discharge temperature from the first stage being high (T = 145.5 °C), an intercooler was introduced between the first and the second stage. Also, another intercooler was considered between the compressor and the extractor since the temperature of the CO₂ in the extractor should not exceed 60°C. The two-stage compressor was designed by considering that the total work for the system is the sum of the work requirement for the two compressors. To validate our calculations Aspen Hysys was used (See figure 2) assuming an isentropic compression with an efficiency of 80% and employing the SRK (Soave Redlich Kwong) method since the working fluid CO₂ is non-polar [27]. The extraction chamber where the microalgae is placed for subjection to supercritical fluid is cylindrical in shape and closed with stainless steel porous discs. The extraction was carried out at 60°C and 300 bars during 4h. For mass balance calculation we considered that the efficiency of the extraction is 97% and that according to literature [6] 2.5% of astaxanthin are present in the biomass. ProRox SL 930 (alumina and glass) were used for the insulation with a thermal conductivity of 0.038 W/m. The thickness of the insulation was calculated as being equal to 0.0078 m. The cyclone used to separate the astaxanthin from the CO₂ was designed by considering an inlet flow velocity of 0.4 m/s.

![Aspen HYSYS Simulation Flow Sheet and stream conditions of the two-stage compressor.](image)

**Figure 2.** Aspen HYSYS Simulation Flow Sheet and stream conditions of the two-stage compressor.

### 3.5 Residual biomass treatment phase

The residual biomass left in the extraction chamber is rich in carbohydrates and could undergo further processing in order to obtain more valuable products. Anaerobic digestion is one of the most cost effective and environmentally friendly approaches used to treat the residual microalgae. The integration of anaerobic digestion with astaxanthin production improves sustainability in the production process. This process is a natural biological process that stabilizes in the absence of air the waste as organic waste and transforms it into biofertilizer and biogas. To reduce pathogen levels in the residual solids obtained from anaerobic digestion, a pasteurization step was performed prior to the anaerobic process. During pasteurization the waste solids were retained at 70°C for 30 minutes and steam injection was used to heat the viscous influent slurry. The digester was designed for a 21 day retention time and as having a cylindrical shape with a height equal to its diameter. ProRox SL 930 (alumina and glass) were used as insulation with a thermal conductivity of 0.038 W/m.K [28]. The
thickness of the insulation was calculated as being equal to 0.003 m. Most of the biogas was produced in the digester at atmospheric pressure and 35 to 37°C [29]. The gas holder volume was assumed to be half of the size of the digester volume. Since the methane is the only component in the biogas with an energy value, the biogas energy value is equal to that of methane. Knowing that 1m³ of methane contained in the biogas produce about 6 kWh of electricity, the quantity of methane produced was used to power some equipment of the plant. The biomass produced was used as a soil fertilizing amendment.

4. Process design
Using the principles defined in the literature [30– 38], the equipment were designed accordingly. The specifications and performance characteristics of each unit are shown in table 2.

Table 2. Design specification for each equipment of the process.

| Code     | Name                        | Capacity/size                                      | Quantity |
|----------|-----------------------------|---------------------------------------------------|----------|
| V-101/V-102 | Sedimentation tank        | Volume of 180 m³                                    | 2        |
| V-103   | Disk-stack centrifuge      | Length of 0.27 m                                   | 1        |
|         |                             | Area of 3.24 m²                                    |          |
|         |                             | Bowl diameter of 10 inch                           |          |
|         |                             | Speed of 9500 RPM                                  |          |
|         |                             | Biomass recovery efficiency 95%                    |          |
| Stream 5 | Conveyor belt              | Width of 0.6 m                                    | 2        |
|         |                             | Length of 6 m                                    |          |
| V-104   | Bead miller                | Stress number $\Delta N$ of 2.9× 10⁷              | 1        |
|         |                             | Stress Intensity $\phi$ of 1,043                   |          |
|         |                             | Biomass recovery efficiency 100%                   |          |
| V-105/V-106 | Spray Dryer            | Volume of 442 m³                                  | 1        |
|         |                             | Biomass recovery efficiency 95%                    |          |
| M-101   | Gas mixer                  | Retention time of 10 s                            | 1        |
|         |                             | Volume of 0.016 m³                                |          |
| C-101   | CO₂ compressor             | Efficiency of 80%                                 | 1        |
|         |                             | Actual Work of 11.9 kW                            |          |
| E-101   | Intercooler                | Efficiency of 80%                                 | 1        |
|         |                             | Area of 3.5 m²                                    |          |
|         |                             | Work of 21.1 kW                                   |          |
| C-102   | CO₂ compressor             | Efficiency of 82%                                 | 1        |
| C-103   | CO₂ Recycled compressor   | Efficiency of 80%                                 | 1        |
|         |                             | Actual work of 4.48 kW                            |          |
| E-102   | Cooler                     | Area of 2.55 m²                                   | 1        |
|         |                             | Work of 15.96 kW                                  |          |
| P-101   | Ethanol pump               | Efficiency of 77%                                 | 1        |
|         |                             | Actual work of 0.44 W                             |          |
| V-1     | Valve                      | Valve flow coefficient of 1.06                    | 1        |
| T-101   | Extractor column           | Volume of 5.6 m³                                  | 1        |
|         |                             | Biomass recovery efficiency 97%                    |          |
| V-107   | Cyclone                    | Cyclone dimension of 0.075 m                      | 1        |
|         |                             | Biomass recovery efficiency 95%                    |          |
| TK-103  | Astaxanthin storage tank   | Volume of 2 m³                                    | 1        |
| E-201   | Heater                     | Area of 16.95 m²                                  | 1        |
| E-202   | Pasteurizer                | Area of 10.96 m²                                  | 1        |
| E-203   | Cooler                     | Area of 21.28 m²                                  | 1        |
| R-201   | Digester                   | Retention time of 21 days                         | 1        |
|         |                             | Volume of 11.6 m³                                 |          |
|         |                             | Diameter = Height = 2.4 m                         |          |
| TK201   | Gas holder volume          | Volume of 5.8 m³                                  | 1        |
| TK-202  | Digestate storage tank     | Volume of 1.5 m³                                  | 1        |
5. Economic Assessment
To evaluate the project profitability, we performed an economic estimation based on the process flow diagram and a rough sizing of the major equipment. The operation days correspond to 360 day in an annual basis and the cost associated with the services and raw materials are listed in Table 3. The total capital investment and the total manufacturing cost are demonstrated in Table 4 and Table 5.

Table 3. Costs of raw materials and services used in the process.

| Item          | Price   |
|---------------|---------|
| **Chemicals** |         |
| CO₂           | 0.4706 $/kg |
| Ethanol       | 784.76 $/m³ |
| **Utilities** |         |
| Biomass       | 32.16 $/kg |
| Cooling water | 0.0294 $/m³ |
| Steam         | 0.0049 $/kg |
| Electricity   | 0.135$/kWh |

Table 4. Equipment costs, fixed capital costs and total capital investment.

| Type           | Cost ($) |
|----------------|----------|
| Vessels        | V-101    | 208,019 |
|                | V-102    | 208,019 |
| Centrifuge     | V-103    | 247,308 |
| Bead miller    | V-104    | 64,800  |
| Spray dryer    | V-105    | 30,000  |
|                | V-106    |         |
| Conveyor belt  | (x2)     | 32,000  |
| Mixer          | M-101    | 5,000   |
| Compressors    | C-101    | 305,056 |
|                | C-102    | 305,056 |
|                | C-103    | 295,811 |
| Heat exchangers| E-101    | 10,650  |
|                | E-102    | 8,275   |
|                | E-201    | 29,612  |
|                | E-202    | 29,955  |
|                | E-203    | 30,118  |
| Pump           | P-101    | 40      |
| Extractor      | T-101    | 36,591  |
| Cyclone        | V-107    | 5,000   |
| Storage tank   | TK-103   | 4,871   |
|                | TK-201   | 4,915   |
|                | TK-202   | 3,777   |
| Reactor        | R-201    | 146,393 |
| Valve          | V-1      | 64,152  |
| Total equipment costs (TEC) | 2,075,418 |
| Equipment erection = 0.5*TEC | 1,037,709 |
| Piping = 0.6*TEC | 1,245,251 |
| Instrumentation and Control = 0.3*TEC | 622,625 |
| Electrical = 0.2*TEC | 415,084  |
| Civil = 0.3*TEC | 622,625  |
| Structures and Buildings = 0.2*TEC | 415,084  |
| Lagging and Paint = 0.1*TEC | 207,542  |
| Offsite = 0.4*TEC | 830,167  |
| Design and Engineering = 0.25*TEC | 518,855  |
| Contingency = 0.1*TEC | 207,542  |
| Fixed capital costs | 6,122,483 |
| Fixed capital investment (FCI) | 8,197,901 |
| Working capital investment (WCI) = 0.15*FCI | 1,229,685 |
| Total capital investment = FCI + WCI | 9,427,586 |
Table 5. Total manufacturing cost.

| Item                                      | Cost ($) |
|-------------------------------------------|----------|
| Direct manufacturing cost (DMC)           |          |
| Raw materials                             |          |
| Biomass                                   | 2,778,624|
| CO₂                                       | 1,056    |
| Ethanol                                   | 4520     |
| Utilities                                 |          |
| Cooling water                             | 200      |
| Steam                                     | 1,587    |
| Electricity                               | 82,331   |
| Labor                                     | 237,089  |
| Supervision = 0.2 * labor                 | 47,418   |
| Plant overhead = 0.5 * labor              | 118,545  |
| Interest = 0.08 * FCI                     | 655,832  |
| Insurance = 0.01 * FCI                    | 81,979   |
| Rent = 0.01 * FCI                         | 81,979   |
| Royalties = 0.01 * FCI                    | 81,979   |
| Maintenance = 0.075 * FCI                 | 614,843  |
| Miscellaneous = 0.1 * maintenance         | 61,484   |
| Shipping and packaging = 0.02 * raw materials | 55,684 |
| Subtotal (DMC)                            | 4,905,150|
| Indirect manufacturing cost (IDMC) = 0.25 * DMC | 1,226,287|
| Total manufacturing cost (TMC) = DMC + IDMC | 6,131,438|

5.1. Total capital investment (TCI)
The TCI is the sum of the fixed-capital investment (FCI) and the working capital (WC). The FCI represents the capital needed to supply the necessary manufacturing and plant facilities and it includes the equipment cost, installation, piping, instrumentation and control, electrical, civil, structures and buildings, lagging and paint, offsite, design and engineering and contingency. The WC refers to the capital for sustaining operations before any product is sold and any income is available.

5.2. Total manufacturing cost (TMC)
All costs related to the manufacturing operations or the physical equipment of a plant are comprised in the manufacturing cost. This cost must be known to judge the viability of a project, and to choose the best alternative processing scheme. The operating costs are divided into two groups: the direct manufacturing cost (DMC) and the indirect manufacturing cost (IDMC). The DMC comprise cost of utilities, raw materials, miscellaneous, shipping and packing, labor, plant overhead, supervision, depreciation, interest, insurance, royalties, rent and maintenance. The operating labor cost is calculated by multiplying the total number of worker by the salary or hourly rate of each worker. The monthly individual labor cost is considered to be 500 $/month. The indirect manufacturing cost (IDMC) includes the distribution, sales, research, general overheads and development and it amounts to 25% of the DMC.

5.3. Return on investment (ROI) and breakeven point
The ROI refers to the return on investment and is used to check the feasibility of the astaxanthin production from a business point of view. The annual profit must be calculated in order to determine the ROI. The ROI is calculated using Equation 3 where CAD refers to the cash available to be distributed as shareholder dividends and it is the subtraction between EBITDA and taxes. EBITDA is the net income before interest, taxes, depreciation, and amortization. Interest on capital during construction period and average feedstock and product inventories are not considered in this production.

$$ROI = \frac{CAD}{FCI} \times 100$$

(3)
To determine the point at which the production starts to generate profit to the industry, we calculated the payout time using the equation below:

\[
Payout\ time = \frac{Original\ depreciable\ fixed\ capital\ investment}{EAT}\ (4)
\]

where EAT represents the sum of the earnings after taxes and the depreciation for a one-year period. Table 6 shows the ROI and the payout time evaluation for an astaxanthin production from microalgae calculated for different market prices of pure astaxanthin (1500, 3000, 4500, 6000, 7000 $/Kg). The fertilizer was also exploited with an average market price of 40$/Kg. This table shows that an economic viability of the company could be achieved for an astaxanthin market price higher than 1500$/Kg. The values of ROI were found to increase with astaxanthin market prices to attain 113% for an astaxanthin price of 6000$/Kg. Also, for this market price the cash flow will completely offset the depreciable investment after one year of production.

Table 6. Financial statement of the plant.

| Price ($)/Kg | Astaxanthin Kg | Astaxanthin Kg |
|--------------|----------------|----------------|
| 1500         | 2,592          | 2,592          |
| 3000         | 2,592          | 2,592          |
| 4500         | 2,592          | 2,592          |
| 6000         | 2,592          | 2,592          |
| 7000         | 2,592          | 2,592          |
| Gross revenue | 6,687,360      | 10,575,360     |
| VAT (11%)     | 735,610        | 1,163,290      |
| Total revenues| 7,421,970      | 11,738,650     |
| EBITDA        | -179,688       | 3,280,632      |
| Depreciation (10%) | 819,790 | 819,790 |
| EBIT          | -999,478       | 2,460,842      |
| Interest expense debt | 9,381,482  | 11,688,362    |
| EBT           | -999,478       | 2,460,842      |
| TAX (10%)     | 2,592,162      | 9,381,482      |
| EAT           | -999,478       | 2,214,758      |
| CAD           | -179,688       | 3,034,548      |
| FCI           | 8,197,901      | 8,197,901      |
| ROI (%)       | -2.19          | 37.01          |
| Payout time (years) | 3.15       | 0.97          | 0.78        |

6. Conclusion

This theoretical study investigated the large scale production of astaxanthin from *Haematococcus pluvialis* for a potential company located in Bekaa (Lebanon). The technical performance of the process was evaluated by sizing the major equipment of the process. Taking into consideration an astaxanthin production of 2592 kg per year, a complete process flowsheet was designed. Detailed operating conditions and equipment designs for the process were obtained. Then, a detailed economic assessment of the project was developed by performing a Profit and Loss analysis. The results have proved the economic feasibility of the production for different astaxanthin market prices. The economic estimation in this paper is classified as a “study estimate” since some information like the process instrumentation diagram or piping and instrumentation requirements, were not considered. Finally, this study could be used as a reference or part of another study that might include the process design of the cultivation of microalgae.

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