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Title: Techno-economic assessment of producing bioplastics from sewage

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Chapter 1. Introduction

Municipal sewage water contains considerable amounts of organic carbon (around 121 gCOD/PE·d) which can be useful carbon-based materials like cellulose or lipids or converted into fatty acids. Polyhydroxyalkanoates (PHA) are produced in special processes, but neither sewage nor sewage sludge contains this substance in larger amounts without enrichment. Utilizing these valuable materials could reduce the use of natural resources and subsequent carbon dioxide (CO_2) emissions helping in realizing a circular economy. The potential recovery of these materials has not been exploited much in North-West Europe (NWE). Currently, very few plants are in operation such as a full-scale plant for cellulose recovery at the STP Aarle-Rixtel in the Netherlands (Cirtec BV 2020) and a pilot-scale plant for recovery of PHA at STP Manresa, Spain (SMART-Plant 2020). In conventional larger sewage treatment plants, part of organic carbon is degraded to CO_2 and H_2O in the aeration tank and the rest of the organic carbon is used in anaerobic digestion to produce biogas. Anaerobic digestion is a mature technology used for the valorization of sewage, however, studies have shown the significance of exploring other technologies to produce higher-value end products such as PHA (Valentino et al. 2019; Fatone 2020; Crutchik et al. 2020). PHAs are polyesters produced naturally by numerous bacteria and are completely biodegradable without any toxic waste. These bacteria accumulate PHA as intracellular carbon and energy reservoirs when limited nutrients are available (Morgan-Sagastume et al. 2016). These bioplastics are also completely recyclable as organic wastes.

Due to the increased awareness of the detrimental effects of synthetic plastics on the environment, research on the production and use of bio-based plastics is receiving increased attention. The physical and chemical properties of bioplastics are comparable to synthetic petroleum-based plastics (Crutchik et al. 2020). Bioplastics have other advantages such as being biodegradable and can be produced from renewable and sustainable resources (Alloul et al. 2018). Due to these advantages, bioplastics have great potential of replacing a major part of the synthetic plastics market as found in the market potential study conduction in the WOW project (Wupperverband 2020). This will aid in mitigating the environmental problems associated with it such as plastic pollution and exploitation of fossil resources (World Economic Forum, Ellen MacArthur Foundation, and McKinsey & Company 2006).

The PHA-based bioplastics have higher market prices due to the use of traditional processes for producing PHA that are not only expensive but also unsustainable. The PHA production price reported in the literature ranges from 1.7-5.3 €/kg (Akiyama, Tsuge, and Doi 2003; Mudliar et al. 2008; Naranjo et al. 2013; Posada et al. 2011; Mumtaz et al. 2010; Leong et al. 2017). These studies were done by using various carbon sources such as palm oil, soybean oil, glucose, glycerol, etc. The market price of PHA produced by using starch as a carbon source is between 7-10 €/kg, which is reported by the company Biomer GmbH (obtained from a survey carried out in the WOW project and presented in the market potential report). It is estimated that the contribution of the carbon source as a substrate to the overall PHA price is between 30-50% (Crutchik et al. 2020). The aforementioned prices are costlier than the market price of synthetic plastics which are less than 1 €/kg (Gholami et al. 2016). Therefore, for reducing the PHA production costs, cheaper carbon sources as substrates are necessary. Many recent studies have identified the waste streams from STPs, agriculture, or the food industry (Fernández-Dacosta et al. 2015; Morgan-Sagastume et al. 2016) as the cheaper source of carbon which can reduce the PHA price.

The use of primary/waste-activated sludge (PS/WAS) from an STP as a carbon source for PHA production has been demonstrated in many studies (Dacosta, Posada, and Ramirez 2015; Bengtsson et al. 2017; Crutchik et al. 2020). The sludge can be converted to volatile fatty acids in an acidogenic fermentation process by inhibiting the methanogenic step in the anaerobic digestion process. The WAS can also be used as a PHA-accumulating bacteria source as it contains several bacteria that produce PHA naturally. The WAS consists of mixed bacteria culture and is proved to be cheaper than other traditional PHA production methods since working with a sterile monoculture is not necessary. Thus, an STP has a cheaper carbon substrate for PHA production and also offers a mixed bacterial culture that has the potential to reduce the PHA production costs significantly. Further research into PHA production integrated with an STP could make it competitive with synthetic plastics.

The valorization of sewage sludge by PHA production has been investigated by several researchers. However, the economic potential of the whole value chain is missing in the open literature. In the WOW project, the value chain consists of several steps including acidogenic fermentation, biomass selection/enrichment, PHA accumulation, PHA separation, drying, extraction, and producing the end product. The partners involved in developing and demonstrating the PHA pilot plant and the value chain are Technische Universität Kaiserslautern (TUK) (Germany), Avans Hogeschool (Netherlands), and NaturePlast (France). The PHA pilot plant until the drying step is developed by TUK and the dried PHA-rich biomass is sent to Avans for PHA extraction. The recovered virgin PHA material is mixed with other materials and used in an injection moulding process to produce end products by NaturePlast. In this report, the results of the techno-economic evaluation are discussed and strategies to optimize the process from an economic point of view are presented. In the next section the overall methodology, the process adopted in the WOW project to produce, extract and convert the PHA into useful end products are described, including the process flow diagram and the mass and the energy balances. The mass and energy data had been provided by the respective partners (TUK, Avans, and NaturePlast) unless mentioned otherwise. Next, the techno-economic assessment methodology adopted to estimate the production cost or the minimum selling price (MSP) of the products is presented. Finally, the results obtained and the key parameters, and their effect on the key performance indicators are discussed. Lastly, the conclusions and recommendations for future research are presented.

Chapter 2. Methodology

In this chapter, the methodology used for the techno-economic assessment, the process flow diagram of the full plant, and the data used to calculate the mass and energy balance, as well as the economic results, are described.

2.1. Techno-Economic Assessment

When developing innovative technologies, such as the production of PHA from primary sludge and its conversion to an end product, it is important to have a clear idea of the economic performance of the process. A techno-economic analysis (TEA) helps to optimize the development of a process and to determine the most important parameters. Consistently applying the methodology will enhance chances of success when introducing (innovative) processes on the market. A TEA considers the entire value chain and can be applied during every technology readiness level (TRL). The methodology can be divided into four different phases. First, a market study is performed. Second, a preliminary process design is defined and translated into a simplified process flow diagram (PFD) and mass and energy balance. Third, this information is directly integrated into a dynamic cost-benefit analysis (CBA) (i.e. economic evaluation). From this analysis, profitability is identified. Fourth, an uncertainty analysis is performed to identify the potential barriers. As information gathering is expensive, a TEA is performed iteratively with a go/no-go decision after every iteration. A graphical representation of the methodology is provided in Figure 1. A detailed description of the methodology can be found in (Van Dael et al. 2015).

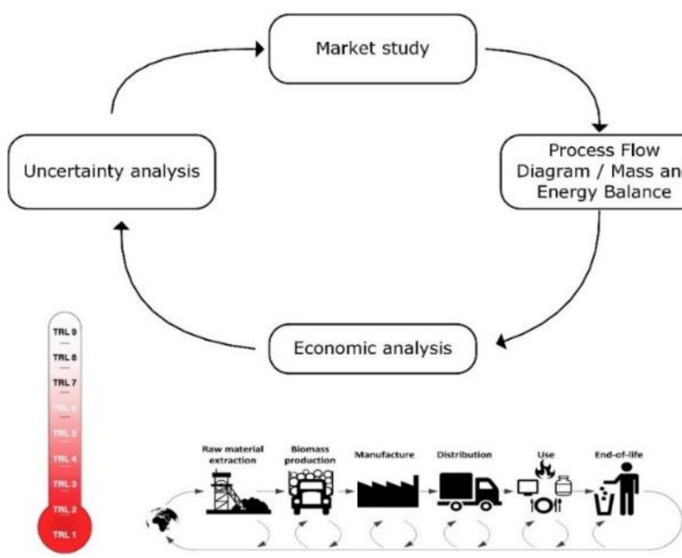

Figure 1. Techno-economic assessment

2.2. Market study

The market study allows the researcher to identify the competitors and customers. It also provides information concerning the size of the market, the needs of the market, and the alternatives on the market. Furthermore, it will also provide information concerning the costs and revenues. Moreover, a market study contains a study of the legislation that is in place. Finally, market research provides insight into market trends. However, the latter is more difficult to estimate when working with innovative technologies. Within the WOW project, a market potential study and factsheet for each envisioned product is prepared by Wupperverband GmbH[1], as lead partner. The other partners involved were Regional Water Authority Vallei en Veluwe[2], University of Kaiserlautern[3], University of Luxembourg[4], Natureplast[5], REMONDIS Aqua Industry[6], Pulsed Heat[7] , and CirTec[8]. The state-of-the-art of legal framework is separately available as a report from Avans Hogeschool[9] as lead partner. The documents can be found on the project website[10].

2.3. Process description and process flow diagram

2.3.1. Pilot plant description

For proving the concept and gaining experiences and design parameters a pilot plant for PHA accumulation was operated on the STP Buchenhofen. The pilot plant layout is shown in Figure 2 and consists of a reactor where the primary sludge was fermented to produce volatile fatty acids (VFA). The VFA-rich solution was then separated from solid biomass using a chamber filter press. The VFA-rich

[1] https://www.wupperverband.de/startseite
[2] https://www.vallei-veluwe.nl/
[3] https://www.uni-kl.de/en/home
[4] https://wwwen.uni.lu/
[5] https://natureplast.eu/
[6] https://www.remondis-aqua.com/en/aq/home/
[7] https://www.pulsedheat.com/
[8] https://www.cirtec.nl/en/
[9] https://www.avans.nl/
[10] https://www.nweurope.eu/projects/project-search/wow-wider-business-opportunities-for-raw-materials-from-wastewater/

solution was split into two streams, one was connected to the biomass selection reactor and the second stream to the accumulation reactor. The biomass stream from the selection reactor was also fed into the accumulation reactor. The PHA-rich biomass from the accumulation reactor was dewatered using a chamber filter press followed by drying in a conventional oven dryer. The dried PHA-rich biomass was sent for extraction.

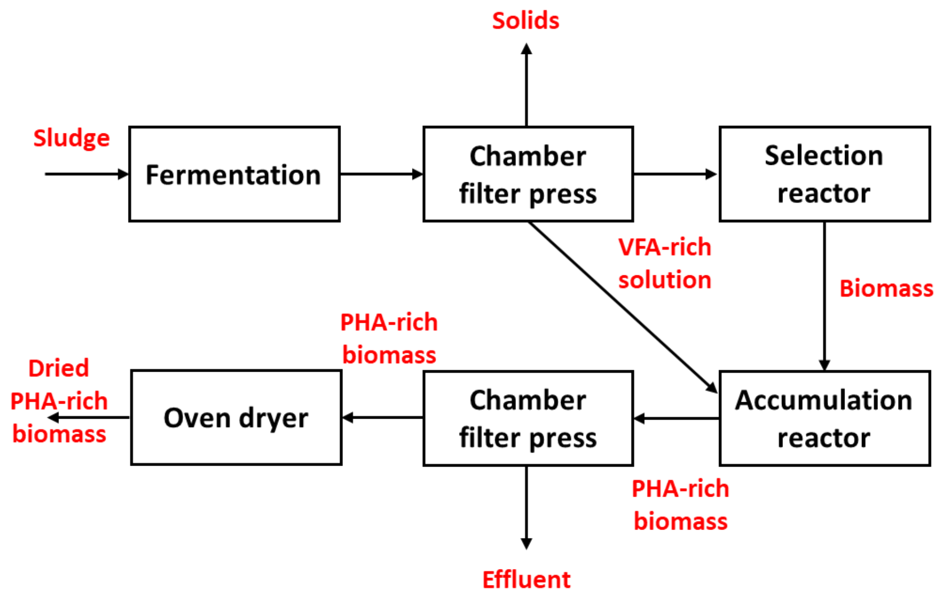

Figure 2. Pilot plant process flow diagram

The mass flow rates, estimated energy consumption, and other equipment specifications are listed in Table 1. This data was provided by the partners, the University of Kaiserslautern (TUK), and for the extraction part by Avans Hogeschool. The pilot plant was designed at a scale of 1.3 m^3 of primary sludge per week.

Table 1. Pilot plant data

Acidogenic fermentation		
Primary sludge	m^3/week	1.3
Density	kg/m^3	1000
Electricity use	kWh/m^3 sludge	96.9
Heat use	kWh/m^3 sludge	23.4
Temperature	°C	35
Retention time	d	7

Reactor size	m^3	1.5
Chamber filter press 1		
Fermented sludge	m^3/week	1.30
Volatile fatty acid solution	m^3/week	1.00
Solids output	m^3/week	0.3
Selection reactor		
Excess sludge	l/batch	350
VFA	l/batch	700
Dilution water	l/batch	700
Electricity use	kWh/m^3	2.51
Hydraulic retention time	d	2.25
Sludge retention time	d	6
Reactor size	m^3	0.45
Accumulation reactor		
Organic acid VFA	m^3 VFA/m^3 volume	0.64
Solid content (TSS)	%	0.1-0.5
Electricity use	kWh/m^3	2.4
Retention time	h	24
Chamber filter press 2		
PHA accumulated biomass	m^3/batch	0.74
PHA-rich output DM	%	20-30%
Lab drying cabinet		
Capacity	kg/week	7.5 - 30
Electricity	kW	2
Temperature	°C	80
Retention time	h	48
Extraction		
Dried PHA	% of input	5%
Dimethyl carbonate solvent	% of input	95%
Solvent recovery	wt.%	99.5%
Extraction time	h	1

Temperature	°C	90

2.3.2. Large-scale plant description

Within the WOW project, a techno-economic assessment of the whole value chain consisting of PHA production from primary sludge and converting it to end products was performed. A large-scale plant was designed based on the pilot plant and literature data and used for the current assessment. Both the chamber filter presses used in the pilot plant were replaced by the centrifuges because better dewatering results were expected. The oven dryer was also replaced by a flue gas dryer to accommodate large inputs. The total suspended solids (TSS or solid content) in the accumulation reactor was assumed to be 0.9% as opposed to 0.1-0.5% listed in Table 1. A higher value was chosen because it can be maintained higher with the optimized operation. For example, Campanari et al. (Campanari et al. 2017) reported a volatile suspended solid (VSS) value of 5 g/L in their batch tests. Assuming the VSS/TSS ratio in the range 0.6-0.9, the TSS would be around 0.5-0.9% making the current assumption valid.

A theoretical large-scale plant was designed for the current assessment with the PHA production capacity of 5000 t/y. The process flow diagram of the entire value chain is shown in Figure 3. The connections to the classical STP are depicted by the red-bordered boxes in the figure. The primary sludge was pumped into an acidogenic reactor for fermentation up to a sludge retention time of 7 days. This is a batch process and thus 7 reactors were assumed to facilitate continuous operations of the process. A standard temperature of 35 °C was maintained in the reactors with regular mixing of the reactor contents. The electricity and heat consumption by this process were 96.9 and 23.4 kWh/m^3 sludge, respectively, similar to the pilot plant. The fermentation process produced volatile fatty acids (VFA), which served as the feed for the PHA-producing bacteria in later stages. The contents from the fermentation step were pumped to separation equipment. A centrifuge with a separation efficiency of 99% and electricity consumption of 1.88 kWh/m^3 (Crutchik et al. 2020) was considered as opposed to a chamber filter press in the pilot plant. At the outlet of the centrifuge, about 77 wt.% of VFA-rich solution were recovered while the rest is removed as solids. These solids were sent to the anaerobic digestion of the STP for biogas production.

The VFA-rich stream was split into two streams where the first stream containing about 36 wt.% of VFA-rich solution was sent to the biomass selection reactor while the rest was sent to the accumulation reactor. For a large-scale plant assessment, a sequencing batch reactor (SBR) was considered for biomass selection and accumulation. In this reactor, the bacteria which can produce PHA naturally were enriched whereas the other bacteria were displaced off by following the feast-famine strategy. The electricity consumption for the biomass selection operation including aeration and mixing was 2.51 kWh/m^3. The output from the selection reactor was supplied to the PHA accumulation reactor. The second split stream containing about 64 wt.% of VFA-rich solution was also fed to the accumulation reactor. The retention time considered was about 24 hours and the whole operation required 2.4 kWh/m^3 of electricity (rough estimation). The outlet of this reactor consisting of accumulated PHA was sent to a centrifuge. It is assumed that about 3 wt.% of the input feed was recovered as PHA-rich biomass and consisted of initial TSS of 0.9 wt.%. The effluent of the centrifuge was returned to the biological treatment step of the STP.

The PHA-rich biomass from the centrifuge was dried using a flue gas dryer as it would be a suitable option at a large scale. The PHA-rich biomass was assumed to be dried at a temperature of 100 °C similar to that used for drying regular biomass. The electricity and heat consumption for the drying step were 0.16 and 1.45 kWh/kg of evaporated water, respectively. The equivalent natural gas cost was estimated based on the heat requirement. The dried PHA-rich biomass was supplied to a reactor where the PHA was extracted using a solvent extraction method. Dimethyl carbonate (DMC) was used for PHA extraction and was heated to a temperature of 90 °C for 1 h and continuously mixed. The required heat was assumed to be supplied by using superheated steam. The steam requirement was 1.1 t/t dried PHA-rich biomass. The electricity required for mixing was 0.01 kW/m^3. The solvent requirement was 19 t/t dried PHA-rich biomass. A cell disruption efficiency of 100% was assumed for this solvent. The waste biomass was filtered out and the filtrate with dissolved PHA was sent to the evaporator. In this step, the solvent was recovered up to 99.5 wt.% of the initial amount. The solvent was cooled and sent to a storage tank for use in the next cycle after making up for the lost solvent. The steam and the cooling duty requirement in the evaporation step were 0.06 and 0.39 t/t filtrate and calculated based on the heating or cooling temperatures. About 50 wt.% of PHA of the initial dried PHA-rich biomass (0.5 g PHA/g PHA-rich biomass) was assumed to be recovered. This value is based on the literature reporting higher PHA yields around and above 50% (Valentino et al. 2020; Mudliar et al. 2008). This virgin PHA was compounded with other raw materials in a compounder and used in an injection moulding process to produce the end products. The details of the compounder and the injection moulding equipment are presented in Table 2.

Table 2. Compounder and injection moulding equipment

Parameter	Compounder	Injection moulding
Capacity (t/h)	0.5	0.045
Operating hours (h/y)	5100	5100

Electricity (kWh/t)	441.7	1503.4
PHA (%)	70%	-
Raw material (%)	30%	-
Waste (%)	-	18

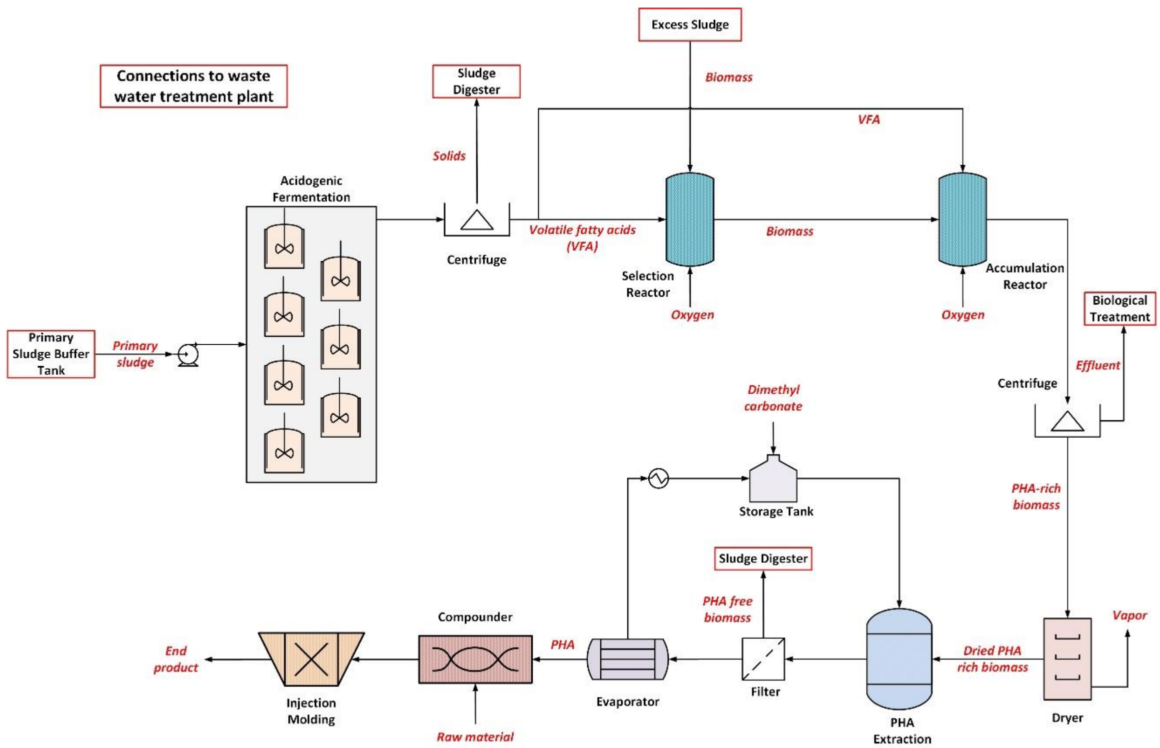

Figure 3. Process flow diagram of the producing bioplastics from sewage

Due to the large amount of primary sludge required to produce 5000 tons of PHA yearly, a decentralized PHA-rich biomass generation seems practical. Therefore, the PHA-rich biomass generation including the steps of acidogenic fermentation, separation, selection, and accumulation was decentralized. The extraction facility including drying, extraction, compounding and injection moulding can be centralized as shown in Figure 4. The sludge is converted to PHA-rich biomass in several STPs and transported to the extraction facility. The number of STPs selected in the current assessment is 10 and is assumed that each STPs generates equal amounts of PHA-rich biomass for simplicity. The effect of increasing the STPs and the average distance between the STP and the extraction facility is investigated in section 3.1.2.

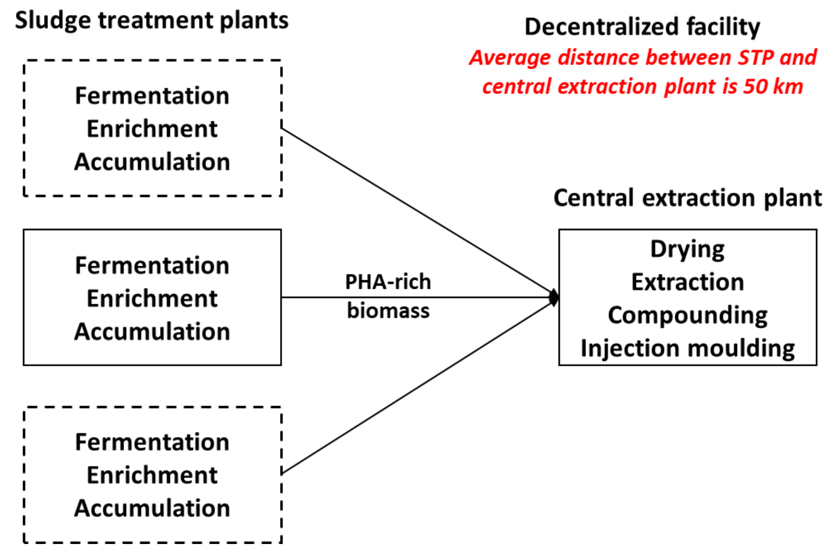

Figure 4. Decentralized PHA production plant configuration

2.4. Economic analysis

To check whether the process is economically feasible and thus worthwhile of investigating from an investor's point of view the mass and energy balance calculations are directly coupled with the economic analysis. The economic analysis should give a clear idea of the capital expenditures (CAPEX) and operational expenditures (OPEX) of the technology. The combination of both provides the total production cost and can be translated into the minimum selling price (MSP). In addition, the revenues are calculated by using the assumed market prices in this study. Using this information, the net present value (NPV), internal rate of return (IRR), and discounted payback period (DPBP) were calculated.

Equipment costs were obtained from the partners, the literature, and/or quoted by the vendors for a certain cost basis. This basis may be a land area, capacity in terms of volume or flow rate, operating pressure, etc. When the values used in the analysis differ from these cost bases, those equipment prices need to be scaled to reflect the new data. One methodology to do this is called the 'six-tenth rule'. It is mainly used for an order-of-magnitude estimation. The rule relates the fixed capital investment cost of a new process to the fixed capital investment cost of a similar previously constructed plant with a known capacity by an exponential ratio relying on the nonlinear relationship between plant capacity and plant cost. This is done using equation [1] by applying a scaling exponential specific to each equipment. The investment costs for the pilot plant were obtained from the literature or the vendors directly and are listed for each equipment in Table 4. To scale up the plant to larger flows of dry matter the reference capacities as shown in Table 3 were used. The general scaling exponent used is 0.6, hence the name 'six-tenth rule', however, the exponents are different for different equipment and are listed in Table 4.

$$Cost\ of\ equipment\ A = (cost\ of\ equipment\ B) \times \left(\frac{Capacity\ of\ A}{Capacity\ of\ B}\right)^{exponent} \quad [1]$$

One problem that might arise with the previous method is that the estimates are based on historical data and that these need to be updated to current prices and economic conditions. The prices that are not up-to-date can be adjusted using the Chemical Engineering Plant Cost Indices (CEPCI) according to equation [2] (CEPCI 2011). Something to consider is that this method is accurate for cost estimates based on data not older than 10 years. If data is older, one needs to be careful with using this index. In this report, the plant lifetime assumed is 25 years and all the costs were estimated based on the year 2019 or converted into 2019 euros using the CEPCI. The operating hours for the plant were assumed to be 8000 h/y. It was expected that with the startup, there would certainly be several things that would need adjustment. The process, however, was developed to run continuously but for such a new technology and new design, the actual running hours would be different. Therefore, the assumption in the current assessment seems valid.

$$Present\ cost = Original\ cost \times \left(\frac{CEPCI_{present}}{CEPCI_{original\ year}}\right) \quad [2]$$

The capital costs were annualized using the equation [3]. The formula for the weighted average cost of capital (WACC) is described in equation [4]. The WACC is the average cost of capital, taking into account the different sources of capital that a firm uses. In this report, a WACC of 4.1% was assumed.

$$Annualized\ cost = \frac{Total\ capital\ cost\ over\ plant\ lifetime}{\frac{1-(1+WACC)^{-Plant\ lifetime}}{WACC}} \quad [3]$$

$$WACC = (Equity\ ratio \times Discount\ rate) + (Debt\ ratio \times (1 - Tax\ rate) \times Interest\ rate) \quad [4]$$

The NPV indicates the profitability of the technology using equation [5], where T is the life span of the investment, CF_n is the difference between revenues and costs in year n, I_0 is the initial investment in year 0, and i is the discount rate. A technology is considered interesting when the NPV is positive (Levy and Sarnat 1994). The NPV compares the amount of money invested in a project today to the present value of the future cash receipts from the investment. In other words, the amount invested is compared to the future cash amounts after they are discounted by a specified rate of return (i.e. discount rate). The NPV considers the investment today and the revenues and expenses from each year of the lifetime of a project. The riskier an investment, the higher the estimated discount rate must be. Typical discount rates are (i)

10% for cost improvement of conventional technologies, (ii) 15% for the expansion of conventional technologies, (iii) 20% for product development, and (iv) 30% for speculative venture (Mercken 2004). However, in most articles, a discount rate of 10-15% was opted in combination with a life span of 10-15-20 years.

$$NPV = \sum_{n=1}^{T} \frac{CF_n}{(1+i)^n} - I_0 \qquad [5]$$

Other popular measures for evaluating whether an investment is financially worthwhile are the DPBP and the IRR. The payback period is defined as the point in time when the initial investment is paid back by the net incoming cash flows, but it has the disadvantage of not taking into account the time value of money. Therefore, one can use the DPBP that does take into account the time value of money. The DPBP can be calculated using the equation [6]. In the equation *CF* is the difference between revenues and costs, *i* is the discount rate and I_0 is the initial investment cost. The shorter the DPBP the more attractive the investment is. The IRR is the discount rate at which the NPV is zero. It thus equates the present value of the future cash flows of an investment with the initial investment and provides the effective interest rate being earned on a project after taking into consideration the time periods when the various cash amounts are flowing in or out. For an IRR to be attractive for an investor it must be higher than the return rate that can be generated in lower-risk markets or investments than the project, e.g. saving the investment money in a bank or investing in safe, low-risk bonds. Because the IRR is a percentage, it can only be used as a decision rule for selecting projects when there is only one alternative to a status quo and should certainly not be used to select one project from a group of mutually exclusive projects that differ in size (Boardman et al. 2006). Therefore, when one has to choose between more than one technology or process (i.e. alternatives), the NPV ranking is mostly preferred over the IRR ranking (Lorie and Savage 1955).

$$DPBP = \frac{ln\left(\frac{CF}{CF - iI_0}\right)}{\ln(1+i)} \qquad [6]$$

The performance indicator for this economic assessment is chosen as the minimum selling price (MSP) of the PHA and the end product. The MSP is the total production cost, including annualized CAPEX and OPEX, per amount of product. The formula for the calculation of the MSP is provided in the equation [7].

$$MSP = \frac{Annualized\ CAPEX + OPEX - Revenues\ coproducts\ (€/yr)}{Product\ (kg/yr)} \qquad [7]$$

Table 3. PHA plant equipment and reference capacity for cost data.

Plant equipment	Type	Reference capacity	Units
Fermentation reactor	Heated anaerobic batch	125	m^3
Centrifuge 1	Continuous	4	m^3/h
Selection reactor	Sequencing batch	30	m^3/d
Accumulation reactor	Sequencing batch	30	m^3/d
Centrifuge 2	Continuous	4	m^3/h
Dryer	Falling curtain	51.6	kg-vapor/h
Extraction reactor	Heated batch reactor	125	m^3
Filter	-	100	m^3/da
Evaporator	Rotary	0.02	m^3/h
Compounder	Twin-screw	500	kg/h
Injection moulding	Electric	45	kg/h

Table 4. Capital cost and operating labor assumptions

Plant equipment	Capital cost (€)	Scale factor	Personnel (per shift)
Fermentation reactors	185,000	0.75	0.02
Centrifuge 1	235,000	0.6	0.35
Selection reactor	263,947	0.78	0.5
Accumulation reactor	263,947	0.78	0.5
Centrifuge 2	235,000	0.6	0.35
Dryer	118,000	0.6	0.5
Extraction reactor	50,255	0.78	0.2
Filter	341	0.6	0.15
Rotary evaporator	2,550	0.6	0.25
Compounder	485,930	0.6	0.2
Injection moulding	122,220	0.6	0.2

Table 5 lists the assumptions used for estimating the capital and operating and maintenance (O&M) costs. The economic life of the plant is assumed as 25 years and all the values are presented in 2019 euros. The equipment costs obtained from the literature and the vendors are increased by 96% of the purchase cost to account for the installation and other costs (piping, instrumentation, electrical, engineering costs, civil works, and start-up). For labor estimation, the number of personnel required per equipment per shift were assigned using the methodology by Peter and Timmerhaus (Peters, Timmerhaus, and West 2003). The personnel per shift were taken from the same reference (Peters, Timmerhaus, and West 2003) and are listed in Table 4. An average labor cost of €31.2 /h was assumed for plant operators and maintenance workers in Europe ("Eurostat - Data Explorer" 2019). The maintenance cost was assumed to be two-third (67%) of the operating labor costs similar to the factor used in the PHARIO project (Bengtsson et al. 2017), and insurance (0.5% of CAPEX) (Crutchik et al. 2020). Furthermore, the variable O&M costs for the electricity, fuel, water, steam, water, solvent and raw material were estimated based on the unit prices given in Table 5. The average PHA-rich biomass transport distance between the STPs and the extraction facility was assumed 50 km. The cost of biomass transport and loading/unloading was assumed €0.08 and €1 /t, respectively.

Table 5. General capital (CAPEX) and operational cost (OPEX) assumptions

Item	Unit	Value
Plant lifetime	y	25
Base year	-	2019
Piping	% CAPEX	15%
Instrumentation/Electrical	% CAPEX	25%
Engineering costs	% CAPEX	10%
Civil works	% CAPEX	34%
Start-up	% CAPEX	12%
Operating hours	h/y	8000
Insurance	%Investment	0.5%
Electricity	€/MWh	93
Natural gas	€/MWh	34
Steam	€/t	24.6
Cooling water	€/m^3	0.5
Process water	€/m^3	1
Dimethyl carbonate	€/kg	1
Raw materials	€/kg	3
Biomass transport distance	km	50

Biomass transport cost	€/t	0.08
Loading/unloading cost	€/t	1 (each)
Post-processing (Inj moulding)	% labor	10%

2.5. Sensitivity analysis

As the values used for the calculations were uncertain, a sensitivity analysis was performed. The prediction of the values was often based on literature and checked with expert opinion. The values are therefore deterministic rather than stochastic. A Monte Carlo simulation (5000 trials) was performed to identify the parameters that had the highest influence on economic feasibility. Within this analysis, the variables (technical as well as economic) were varied following a triangular distribution over specified ranges depending on the variable. The goal of this kind of quick scan is to determine the parameters that have the highest impact on the variance of MSP. The analysis searches for the parameters that should be investigated in more detail. For these parameters, a local sensitivity using what-if analysis was performed to see how changes in these parameters influence the economic feasibility.

Chapter 3. Results and discussion

In this chapter the results of each step of the techno-economic assessment for the two cases are described, i.e. (1) base case with PHA production and (2) base case with the end product. Firstly, the results for the base case when PHA is the product are discussed. Next, the results for the base case when the end product is produced are presented. In the third section of this chapter, the two cases are compared using a comprehensive sensitivity analysis.

3.1. Production of PHA (bioplastic)

3.1.1. Mass and Energy Balances

The mass and energy balances of the entire plant including all the steps for producing the PHA are depicted in Figure 5. The pilot plant data until the production of PHA-rich biomass at the outlet of the second centrifuge was provided by the University of Kaiserslautern (TUK) whereas the data on the PHA extraction process was provided by Avans Hogeschool. The unavailable data was acquired or estimated from the literature as discussed in the previous chapter. Using the pilot plant data and partly literature data, mass and energy balances for a large-scale plant were calculated. The large-scale plant is a combination of several decentralized plants designed for a combined PHA production capacity of 5000 t/y which corresponds to the primary sludge flow of 843,313 t/y. The fermentation of the primary sludge produced about 649,351 t/y of VFA-rich solution. At the outlet of the selection reactor, the amount of biomass produced was about 584,416 t/y whereas the stream consisting of PHA-rich biomass after the accumulation reactor was about 1,000,000 t/y. The amount of dewatered PHA-rich biomass obtained after the centrifuge was about 30,000 t/y. After drying out most of the water, the amount of dried PHA-rich biomass obtained was 10,000 t/y. In the extraction process, 5,000 t/y of virgin PHA was obtained and required a make-up solvent in the amount of 950 t/y. The total electricity and heat requirement of the plant was about 90,764 and 48,734 MWh/y, respectively. The steam and cooling water requirement in the extraction process was 21,337 and 75,382 t/y, respectively. The assumptions and detailed mass balance calculations are presented in the Appendix.

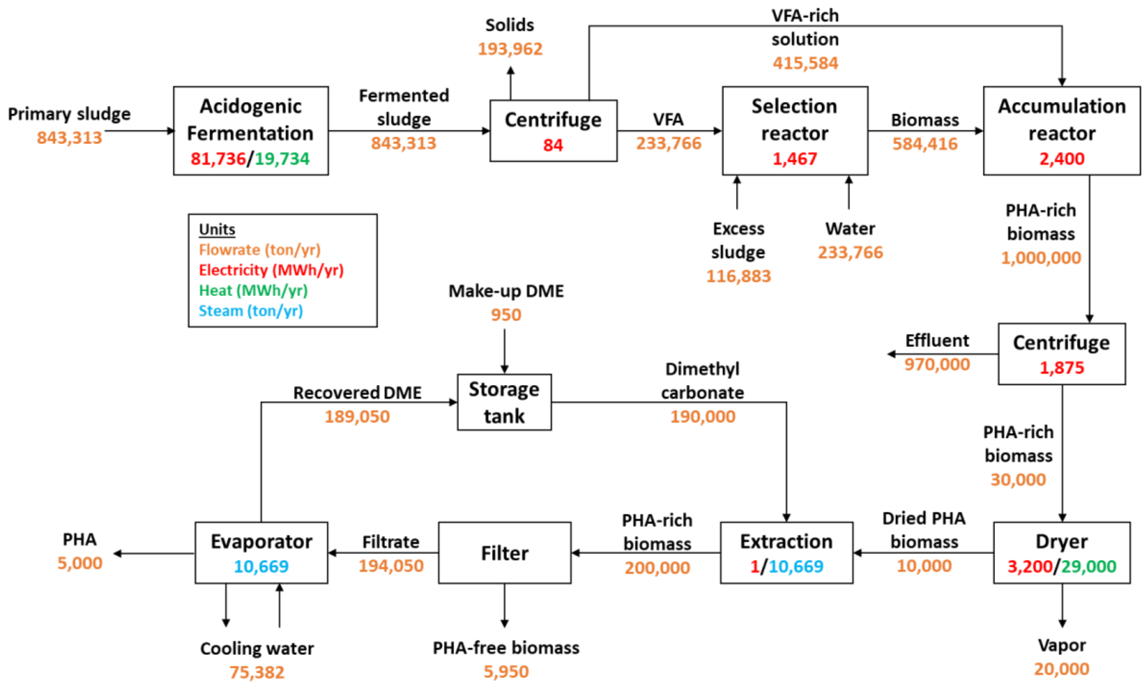

Figure 5. Mass and energy balance of PHA production

3.1.2. Economic analysis of PHA production plant (decentralized)

The economic assessment results for the base case are discussed in this section. Figure 6 shows the cost breakdown of capital costs (CAPEX) and O&M costs (OPEX). For the base case, annualized CAPEX and OPEX were €3,871,377 and €13,834,282, respectively. The major contribution to CAPEX comes from the PHA accumulation reactor (30%) since this reactor was fed with large inflows of VFAs and the enriched biomass stream and thus required a larger volume. The biomass selection reactor contributes about 21% of the overall CAPEX. This reactor receives inflows of smaller VFA stream and biomass bacteria mix which have lower mass flow rate than that to accumulation reactor. It is to be noted that these two SBRs constitute about 51% of the overall CAPEX. Therefore, a cheaper alternative reactor configuration that performs the same operations as SBR could lower the PHA production costs. The dryer and the fermentation reactors contribute 2% and 15% to the overall CAPEX, respectively. As mentioned earlier, the number of fermentation reactors is equivalent to the sludge retention time in days. This was done to have a continuous operation of the plant. However, if there is any other arrangement similar to an SBR reactor where all the operations can be done in a single reactor then the production cost can be brought down. The two centrifuges contribute almost equally with a total of 32% to the overall CAPEX. The extraction process consisting of a reactor, filter, evaporator, and storage tank contributes only about 1% to the overall CAPEX since these equipment are relatively simple and commonly available.

The OPEX breakdown in Figure 6(b) shows that the largest contribution was from electricity (61%) followed by heat (12%) required in the fermentation and drying processes. The steam requirement during the extraction step contributes about 4% to the overall OPEX. Thus, much of the OPEX is due to the energy requirement of the plant, about 77%. The fixed OPEX including labor, maintenance, and insurance contributes 6%, 4%, and 3%, respectively. The solvent contribution is about 7% despite 99.5% recovery. This is due to the large requirement of the solvent per ton of dried PHA. An alternative solvent with lower cost, lower requirement, and higher recovery would reduce the PHA production cost considerably. The cooling and process water contributes about 2% to the overall CAPEX of which the cooling water can be recovered completely. The least OPEX was due to the transport of sludge from decentralized STPs to a centralized extraction facility, about 1% of total OPEX.

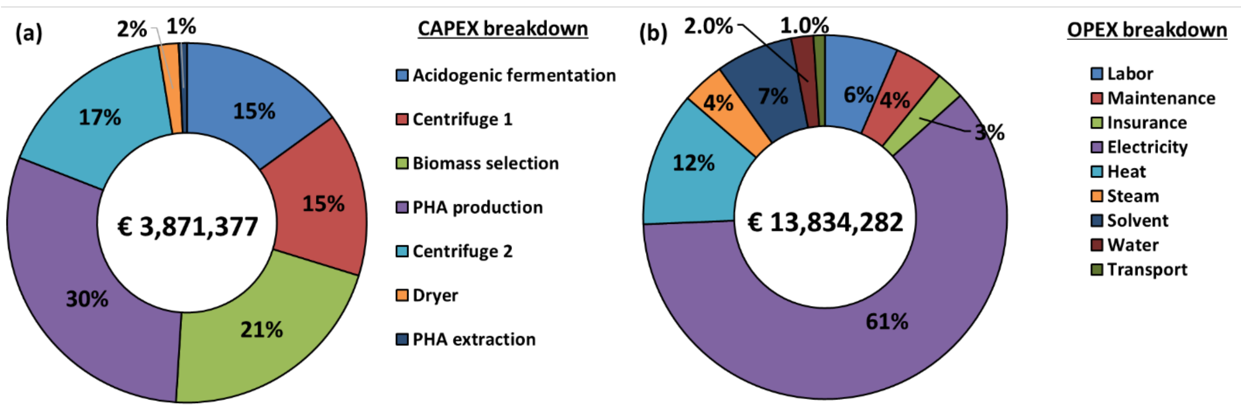

Figure 6. PHA production (a) Capital cost and (b) Operating cost breakdown

The overall breakdown of the minimum selling price (MSP) or the production cost of PHA is shown in Figure 7. The MSP of the PHA estimated in this assessment is €3.54 /kg PHA which is in parity with that reported in the literature and listed in Table 6. Compared with the literature using other feed sources, the current MSP is much lower. However, compared with the literature where wastewater was the feed source, the current MSP is considerably higher. These studies did not provide details on the economic assessment methodology used and thus the results cannot be compared. Moreover, these studies were based on the regions outside Europe and it is expected that the cost estimations will differ considerably. The results from the PHARIO project (Bengtsson et al. 2017) estimated the PHA production price to be

€3.6 /kg which is higher than the current estimation. The commercial company Biomer[11] GmbH in Germany produces PHA from starch and estimated the market price to be in the range of €7-10 /kg. Furthermore, a market potential study conducted in the WOW project estimated the PHA market price to be in the range of €3.5–4.5 /kg (Wupperverband 2020), which is higher than that in the current assessment. As mentioned earlier, the cost of utilities (heat, electricity, and steam) has the largest contribution to the MSP of PHA. They contribute about 60% of the overall PHA price, mainly by the electricity and heat requirement in the dryer. Therefore, adopting strategies that would reduce the utility consumption would bring down the PHA price significantly. The contribution from the investment was about 22% and thus lowering the CAPEX by using the large STPs for PHA-rich biomass generation would affect the PHA production price considerably. The fixed OPEX including the labor, maintenance, and insurance contributed just 10% to the PHA production price. Another important contribution was from the solvent (5%) which can be reduced by using a cheaper solvent with less requirement per ton of dried PHA-rich biomass during the extraction process. This will reduce the steam requirement during the extraction and solvent recovery steps and will also reduce the equipment size.

[11] https://www.biomer.de/

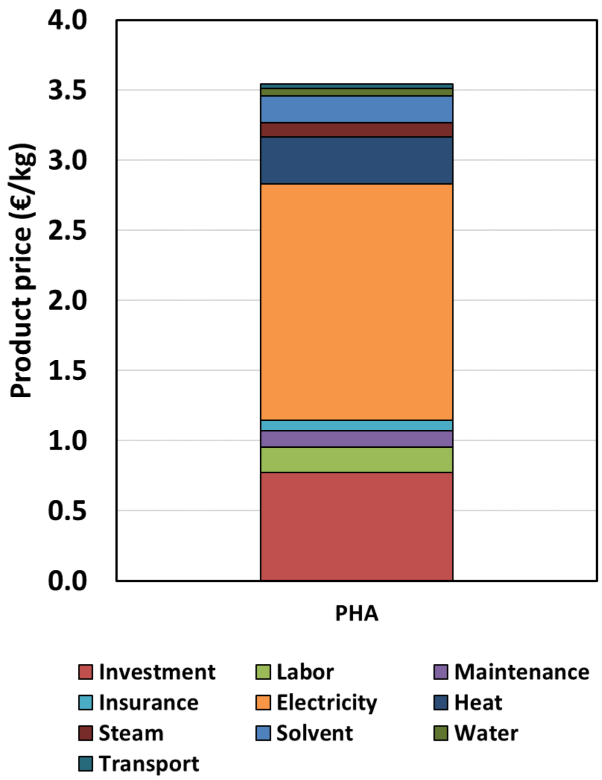

Figure 7. MSP breakdown of PHA

Table 6. PHA production price

Source	Price, €/kg (2019)	Reference
Soybean oil	4.5	(Akiyama, Tsuge, and Doi 2003)
Glucose	4.9	(Akiyama, Tsuge, and Doi 2003)
Activated sludge	4.8	(Mudliar et al. 2008)
Palm oil	1.9	(Mumtaz et al. 2010)
Glycerol	1.7	(Posada et al. 2011)
Glycerol	2.4	(Naranjo et al. 2013)
Wastewater	1.5	(Fernández-Dacosta et al. 2015)
Paper wastewater	1.7	(Dacosta, Posada, and Ramirez 2015)
Glycerol	5.3	(Leong et al. 2017)
Secondary sludge	3.4	(Bengtsson et al. 2017)
Primary sludge	1.0	(Crutchik et al. 2020)
Starch	7-10	Survey report – Biomer GmbH

Table 7 compares the current results with that of the PHARIO project. The CAPEX estimated in the current assessment is ~1.9 times lower than that estimated in the PHARIO project. For the selection and accumulation reactor, the cost data were acquired from the literature as opposed to the PHARIO project where a company was hired which specialized in cost estimation. It can be concluded that the cost estimation in the PHARIO project was in more detail. However, the estimation was done for a plant capacity of 2000 tPHA/y and then multiplied by a factor of 2.5 to get the cost estimates of a plant capacity of 5000 tPHA/y. This makes the cost estimation inaccurate since it is well known that the cost increases with the scale by a scaling exponent as shown in Eq [1]. Nevertheless, the current capital cost estimation can be considered underestimated due to the lack of details available on the construction of a real large-scale plant. The OPEX in the current assessment was estimated to be ~24.8% higher than that in the PHARIO project. Many details especially the heat requirement in the dryer was missing in the data provided by the PHARIO project. Nonetheless, the estimated PHA price is almost the same, about 1.6% lower than that reported in the PHARIO project. Considering the PHA market price of €4 /kg as found in the market potential study conducted in the WOW project (Wupperverband 2020), the economic indicators such as NPV, IRR, and DPBP are very promising as shown in Table 7. This could be due to the underestimation of the CAPEX in this assessment. Therefore, the effect of increasing the CAPEX on the MSP of PHA is investigated in the sensitivity analysis in Section 3.3.

Table 7. Comparison with PHARIO project results

	Current study	(Bengtsson et al. 2017)
Feedstock	Primary sludge	Primary sludge
Normalized CAPEX (€/t PHA)	774	1455
Normalized OPEX (€/t PHA)	2767	2080
PHA output (t/y)	5000	5000
PHA MSP (€/t)	3541	3600
Net present value (€)	26,933,785	-
Internal rate of return (%)	15%	-
Discounted payback period (y)	8.4	-

3.2. Production of the end product

3.2.1. Mass and energy balance

The mass and energy balance for producing an end product from PHA is shown in Figure 8. In this project, an injection moulding technology is considered. The PHA is mixed with another raw material (to enhance the properties) in the ratio of 70%:30% by mass in a compounder. The compounding process produces homogenized granules which were then fed to injection moulding equipment. From the previous section, the amount of PHA obtained was 5,000 t/y, and when mixed with 2,143 t/y of raw material and recycled waste from the injection moulding process produced about 8,711 t/y of granules. Wastage of 18% is assumed for the injection moulding process which is relatively higher than other plastic processing technologies such as extrusion and thermoforming. The waste material is considered to be recyclable and is re-used in the compounding process. Having more waste material after post-processing increases the energy consumption since this material is unused and can only be reused after re-compounding with the fresh granules. An optimized mould design would be required to reduce wastage. After accounting for wastage, the amount of end product obtained was 7,143 t/y. The electricity consumption for the process is 16,251 MWh/y as shown in Figure 8. The total electricity consumption for the entire plant including the PHA-rich biomass generation would be 107,015 MWh/y. However, it is unlikely that the injection moulding is done at the same site. Therefore, the PHA could be considered as a separate product and can be sold to the companies specializing in producing end products. But for the sake of completeness, the results and the analysis are presented for the whole value chain.

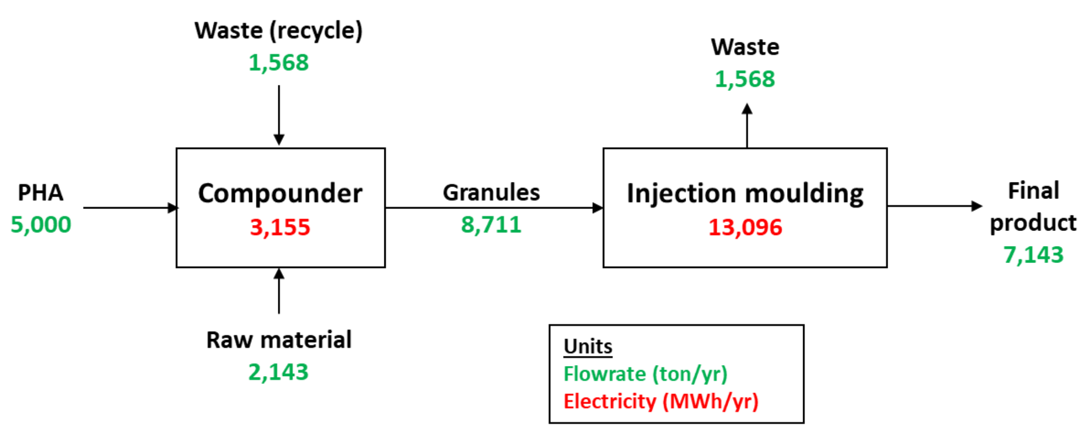

Figure 8. Mass and energy balance of production of the end product

3.2.2. Economic analysis

The CAPEX and OPEX breakdown for the entire plant is shown in Figure 9(a) and (b), respectively. The annualized CAPEX and OPEX were €4,156,810 and €21,956,342, respectively. Nearly half of the CAPEX still comes from the selection (20%) and accumulation (28%) reactors. The addition of the injection moulding process contributes only about 7% to the overall CAPEX. In terms of OPEX, the major contribution comes from the raw material (29%) which is priced at €3 /kg. Therefore, cheaper raw materials (synthetic or bio-based) would reduce the production cost of the end product significantly. Overall, including the injection moulding process increased the OPEX by €8,122,060 as compared to the case of PHA production. The breakdown of the end product MSP is shown in Figure 10. The MSP estimated was €3.7 /kg, of which the raw material contributed €0.9 /kg. This shows the great dependence of the MSP on the raw material price and the blend ratio. These parameters are investigated further in the sensitivity analysis in the subsequent section.

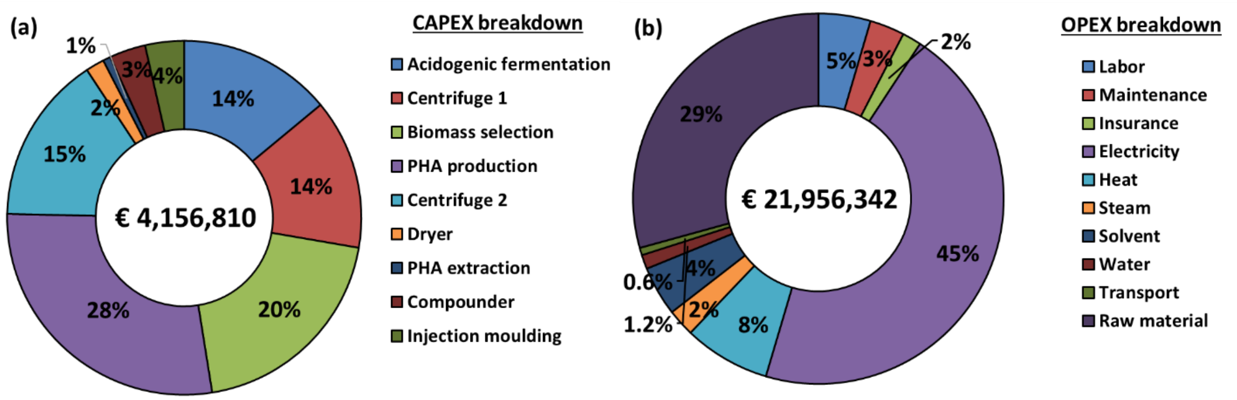

Figure 9. End product (a) Capital cost and (b) Operating cost breakdown

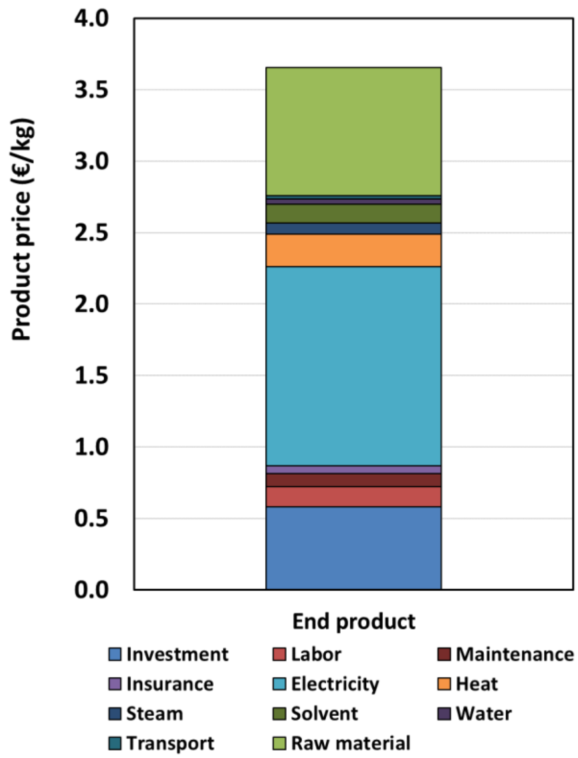

Figure 10. MSP breakdown of the end product

3.3. Sensitivity analysis

In this section, firstly the impact of changes in four technical parameters (plant scale, PHA yield, drying heat, and cell disruption efficiency) and three economic parameters (CAPEX, labor requirement, and fuel price) on the economic feasibility of PHA production are discussed. Next, each of these parameters is investigated separately in terms of PHA and end product MSPs using a local sensitivity analysis. Finally, the impact of the blend ratio and the raw material price on the MSP of the end product are also discussed.

3.3.1. Monte-Carlo analysis

The Monte-Carlo simulation was done only for the base case where PHA is produced. Since the process under study is innovative, not much information was available in the literature. Therefore, the variables selected for sensitivity were based on their contribution to the products' MSP and the opinion of the pilot plant owners. In total, 8 variables were considered and varied over specified ranges namely plant scale (represented by primary sludge flowrate in t/d), PHA yield (%), drying heat requirement (%), cell disruption efficiency (%), solid content (%), CAPEX (% increase or decrease), labor requirement (% increase or decrease), and fuel price (€/MWh).

The lower boundary for the plant scale was set at -40% (1518 t/d) from the reference value (2530 t/d) while the maximum scale was set at +60% (4048 t/d). This was done to take into account the large variations in the scale of STPs. In a decentralized system, the lower value indicates less number of PHA-rich biomass generation plants whereas the higher value indicates more number of PHA-rich biomass generation plants. The sources can be industries, small municipalities, and bigger municipal sources such as towns or cities. The PHA yield was varied from 30% to 70% of dried PHA-rich biomass with 50% as the reference value. The yield obtained from the pilot plant data was low but can be increased by optimizing the pilot plant operations. The highest yield reported in the literature is 64% (Mudliar et al. 2008) and can be expected to reach up to 70%. For the drying heat requirement, the maximum was kept the same as the reference value and a lower boundary was set at -40% of the reference value. The goal is to reduce the energy required for drying. The cell disruption efficiency considered for this assessment is 100% but in reality, it might be lower due to practical challenges. Thus, it was varied from 80% to 100% to cover the uncertainty in the disruption efficiency. The PHA-rich biomass output at the second centrifuge was used to represent the solid content in the accumulation reactor. The PHA-rich biomass was assumed to be about 3 wt.% which along with a dry matter value of 30 wt.% resulted in 0.9 wt.% solid content in the accumulation reactor. The PHA-rich biomass value was varied from 2 to 10 wt.% at the centrifuge outlet and was assumed to be controllable. The CAPEX is highly uncertain because of differences in available data. Therefore, the CAPEX is varied -50% to +100% of the reference value to accommodate a large range of CAPEX. The labor requirement was lowered to -60% of the reference value. A lower wage rate indicates less manual labor and high automation whereas a higher wage rate indicates more manual labor and less automation. This mainly depends on the plant design and thus the variation cannot be verified from the literature. The heat required in this plant is assumed to be generated by burning natural gas. As the heat requirement is considerable, the fuel price is expected to have a considerable effect on the PHA production cost. Therefore, the fuel price is varied from -60% to +60% of the reference value (14-54 €/MWh).

Figure 11 shows the contribution of each selected variable towards the variance in MSP of PHA over selected ranges. A positive value in the figure indicates that an increase in the variable results in an increase in the MSP whereas the negative values indicate that MSP decreases with an increase in the variable. The variables having negative sensitivity are PHA-rich output (-68.4%), PHA yield (-23.6%), plant scale (-2.3%) and cell disruption efficiency (-1.6%) whereas the variables having positive sensitivity are

CAPEX (+3.2%), fuel price (+0.6%), drying heat (+0.2%), and labor (+0.0%). The PHA-rich output and yield have a significant impact on PHA production cost than all the other variables combined. The impact of these variables on the MSP is further investigated in the following subsections.

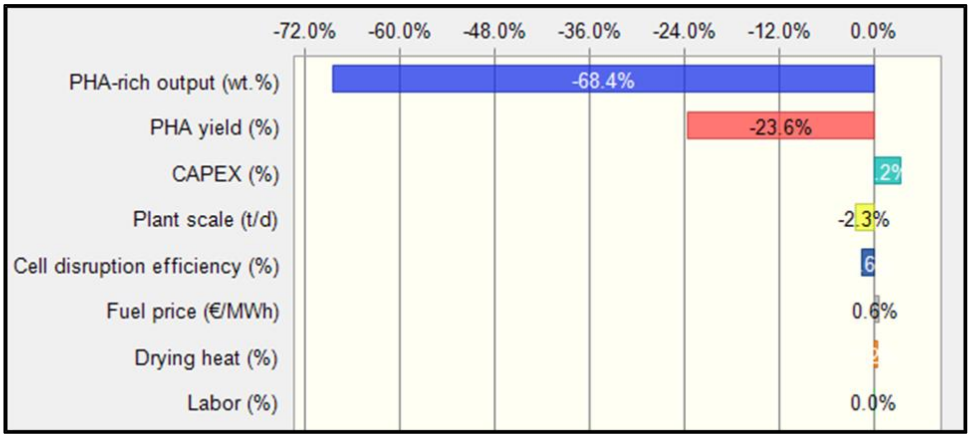

Figure 11. Relative contribution of the variables' range to the variance in MSP of PHA

3.3.2. Impact of plant scale

The primary sludge is varied from 1518 t/d to 4048 t/d with a base value of 2530 t/d. This base value is calculated to have the PHA production capacity of 5000 t/y. Considering the average sludge generation of 35 g solid/PE·d at 3% solids and sludge density of 1000 kg/m^3, the STP size would be around 2,168,518 PE. However, plants of this size are very rare in NWE. So, 10 decentralized PHA-rich biomass production plants were considered and the biomass was transported to a centralized PHA extraction facility. The effect of the plant scale on the MSPs of the PHA and the end product is shown in Figure 12. It can be observed that the MSPs of the two products decrease as the scale goes from small to large. With an increase in the plant scale, the equipment size was also increased by a specific exponential relationship particular to each equipment. This resulted in lower operating expenses per unit plant capacity accounting for the economies of the scale factor. The reduction in MSP of the PHA and the end product was 16% and 12%, respectively, as the scale is varied from small to large. Moreover, when compared to the market price of PHA (€4 /kg), even at the lowest scale, the PHA production cost is about ~2% lower indicating that the PHA production is economically feasible at all plant scales considered.

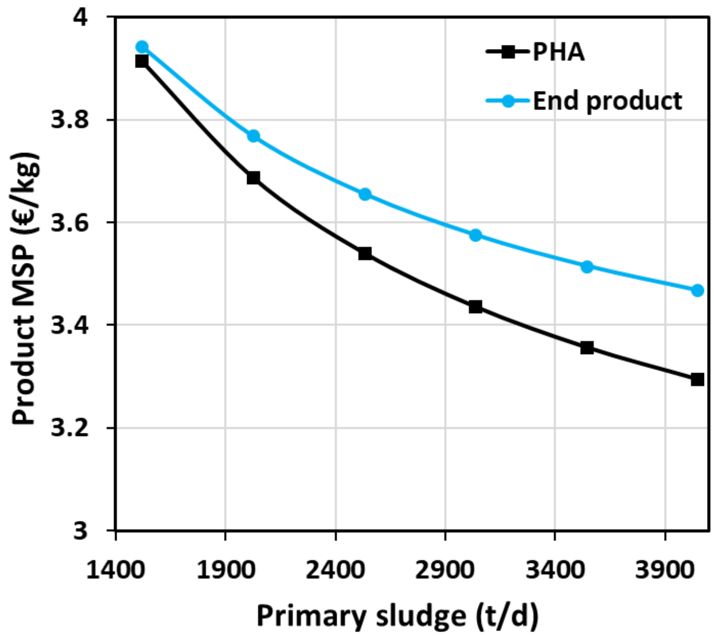

Figure 12. Impact of plant scale on the MSP of PHA and end product

In a decentralized system, the primary sludge from several smaller STPs is enriched to produce PHA-rich biomass and transport it to the extraction facility. Figure 13 shows the effect of the number of STPs and it can be observed that as the number of STPs increase the MSP of PHA and the end product also increase. This is due to the increase in CAPEX in smaller STPs taking into account the economies of scale. If there are only 2 STPs considered, the MSP of PHA estimated is €3.2 /kg whereas if there are 20 STPs involved, the MSP increases to €3.8 /kg. Nonetheless, it is lower than the average PHA market price of €4 /kg. The effect of average transport distance from the STPs to the extraction facility is shown in Figure 14 and it is varied from 25 to 100 km with 50 km as the base distance. The effect of the transport distance on the price of the products is very small.

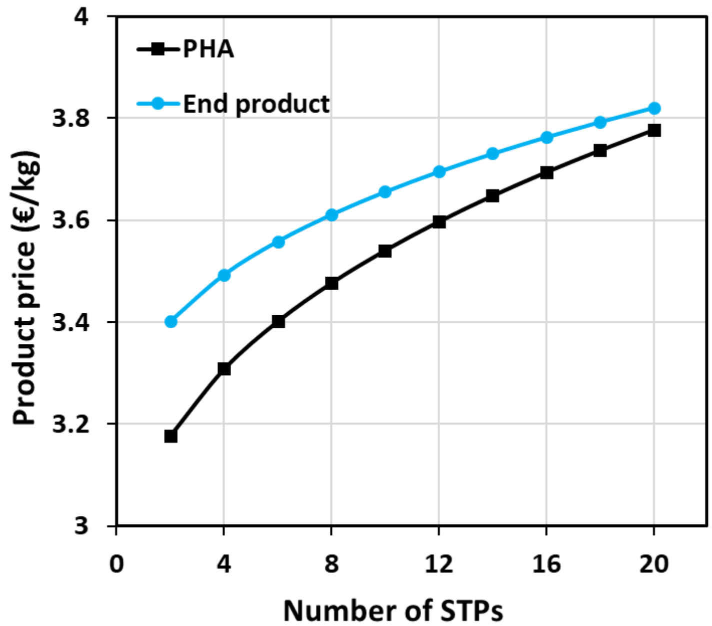

Figure 13. Impact of number of STPs on the MSP of PHA and end product

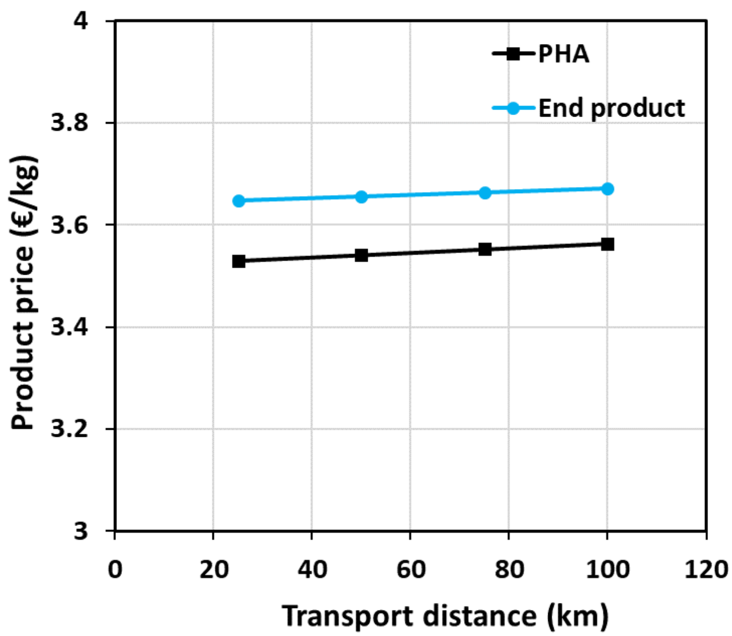

Figure 14. Impact of transport distance on the MSP of PHA and end product

3.3.3. Impact of PHA yield

As discussed in section 3.3.1, the Monte-Carlo analysis showed that the PHA yield is the second most significant parameter. The PHA yield is varied from 30% to 70% of dried PHA-rich biomass based on the information provided in the literature. The effect of varying the yield on the MSPs of PHA and the end product is shown in Figure 15. The corresponding reduction in MSPs observed is 57% and 45%, respectively. The PHA production capacity increased from 3000 t/y to 7000 t/y, which is an increase of 133%. However, there is no increase in either the CAPEX or OPEX since the same amount of biomass is processed but with higher PHA content. In the case of the end product, an increase of 3% and 34% was observed in the CAPEX and the OPEX, respectively. This is due to an increase in the size of injection moulding equipment and the amount of raw material required to maintain the 70:30 blend ratio. The PHA yield is an intrinsic process parameter, which is independent of plant capacity. Though the reference value obtained from the pilot plant data was 35%, it was reported that the PHA yield can be altered by manipulating the carbon to nitrogen ratio, using inhibitors and novel fermentation strategies (Mudliar et al. 2008). It is interesting to note that at around a PHA yield of 45%, the MSP of virgin PHA and the end product are the same. The reason is the blend ratio of 70% raw material with a price of €3 /kg, 30% of virgin PHA with a price of €3.9 /kg, and the additional contribution from CAPEX and OPEX of the injection moulding process. The sum of these cost elements becomes equal to the MSP of virgin PHA.

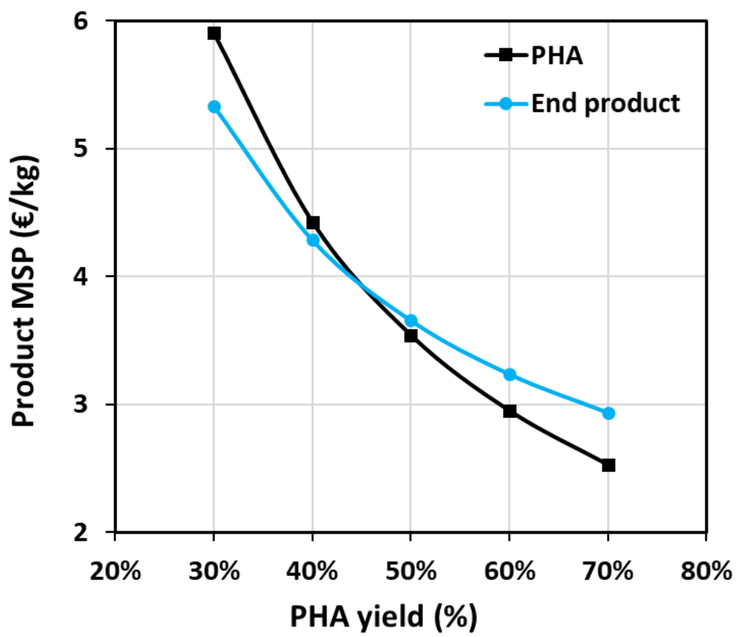

Figure 15. Impact of PHA yield on the MSP of PHA and end product

3.3.4. Impact of heat requirement in drying

Drying is an important step in the PHA production process where a lot of energy is consumed. The dewatered PHA-rich biomass from the centrifuge was dried up to 90% DM in this step. It required about

2.15 MWh of heat per ton of PHA-rich biomass. The required heat was assumed to be generated by burning the natural gas in a burner. The drying heat was reduced to 40% of the reference value and the effect on the MSP is shown in Figure 16. It can be observed that if the drying heat requirement were reduced by 50%, the reduction in the MSP of the PHA and the end product was 5% and 3%, respectively.

If an alternative drying process is developed which consumes less thermal energy, the MSP can be reduced further. The vacuum evaporation method could be used to remove the moisture from the PHA-rich biomass at low energy consumption. Due to vacuum pressures, the boiling point of water is lowered which requires a less intensive source of heating. The operating costs especially for energy in the current process can be significantly reduced by using the vacuum evaporation method. The energy requirement per kilogram of water evaporated was 2.98 MWh/t in the current study. This can be reduced to as low as 0.2 MWh/t by using the vacuum evaporation method (Aquadest 2021). The only drawback is that these types of evaporators can accommodate only semi-solid state material. Due to this reason, a hybrid drying process would be beneficial that has an energy consumption somewhere between the two extremes.

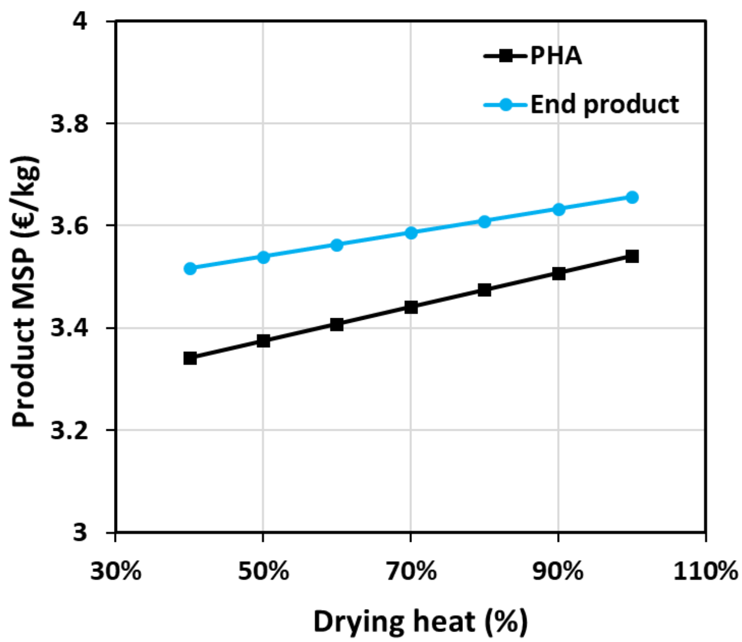

Figure 16. Impact of drying heat on the MSP of PHA and end product

3.3.5. Impact of cell disruption efficiency

The most common PHA recovery method is the solvent extraction method. The cell disruption efficiency is dependent on the solvent used. Many studies reported in the literature have explored the use of environmentally friendly chemicals and the strategies to improve cell disruption efficiency. The solvent extraction method leads to high recovery but at the same time is expensive and not environmentally friendly (Gholami et al. 2016). However, due to the extraction being a simple process and a very high recovery of the solvent for re-use, solvent extraction is considered the standard method in the industry (Gholami et al. 2016). Several solvents had also been used as reported in the literature such as chloroform (Mudliar et al. 2008), dichloromethane (Samorì et al. 2015), etc. The solvent used in this project was dimethyl carbonate (DMC) since it is a green solvent, has high disruption efficiency, and has low toxicity (de Souza Reis et al. 2020). The reference efficiency for the base case considered was 100% and it was reduced up to 80% to investigate the effect on the MSP. The results obtained are shown in Figure 17 and it can be observed that the cell disruption efficiency has a significant effect on the MSP. The disruption efficiency of 80% indicates that 20% of the intracell PHA accumulated was not extracted and discarded as waste. This reduced the PHA production capacity and consequently, increased the products' MSP. The increase in the MSP of PHA at 80% disruption efficiency is about 20% when compared to the MSP at 100% disruption efficiency. As seen earlier, the contribution of the solvent towards the overall PHA production cost is 5%. Therefore, a green alternative solvent that is cheaper, has high disruption efficiency and low toxicity would aid in reducing the MSP of the PHA and the end product. At 90% disruption efficiency, the end product price becomes equal to the virgin PHA price. The reason is the blend ratio, raw material price, and additional CAPEX and OPEX of the injection moulding process as explained in the earlier section.

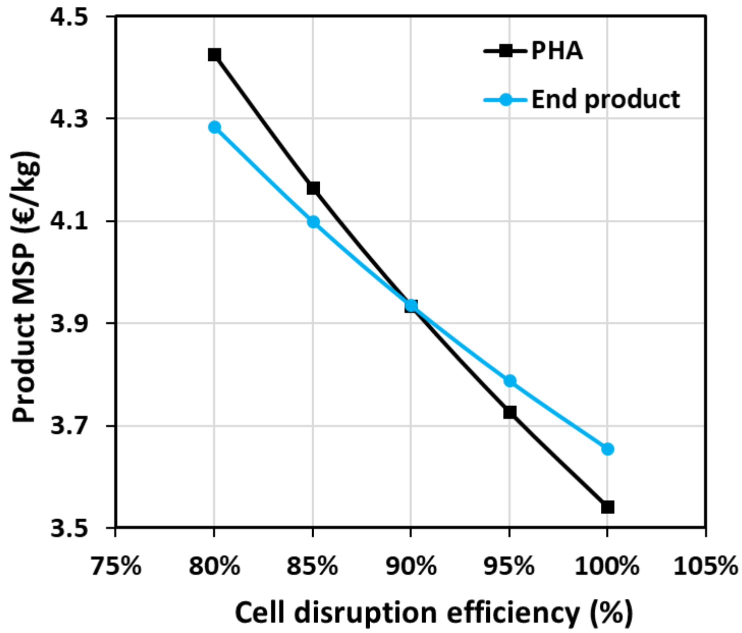

Figure 17. Impact of cell disruption efficiency on the MSP of PHA and end product

3.3.6. Impact of PHA-rich output (solid content/TSS)

The Monte-Carlo assessment showed that the PHA-rich biomass at the centrifuge outlet was the most significant parameter. As mentioned earlier, this parameter directly represents the solid content in the accumulation reactor which is shown in Figure 18. By using the PHA-rich output value of 3 wt.% and dry matter value of 30 wt.%, the base case value of solid content was 0.9 wt.%. Since the PHA yield assumed is 50%, half of the solid content would be PHA. Therefore, this value is highly significant to the MSP. The PHA-rich output value was varied from 2 to 10 wt.% which corresponds to a solids content of 0.6 to 3 wt.%. By optimizing the accumulation process to have 3 wt.% solid content, the PHA can be produced at €1.5/kg, which is significantly cheaper than the market price. However, very low values of solid content in the range of 0.5 to 1 wt.% were reported from the pilot plant and the literature (Campanari et al. 2017; Valentino et al. 2020).

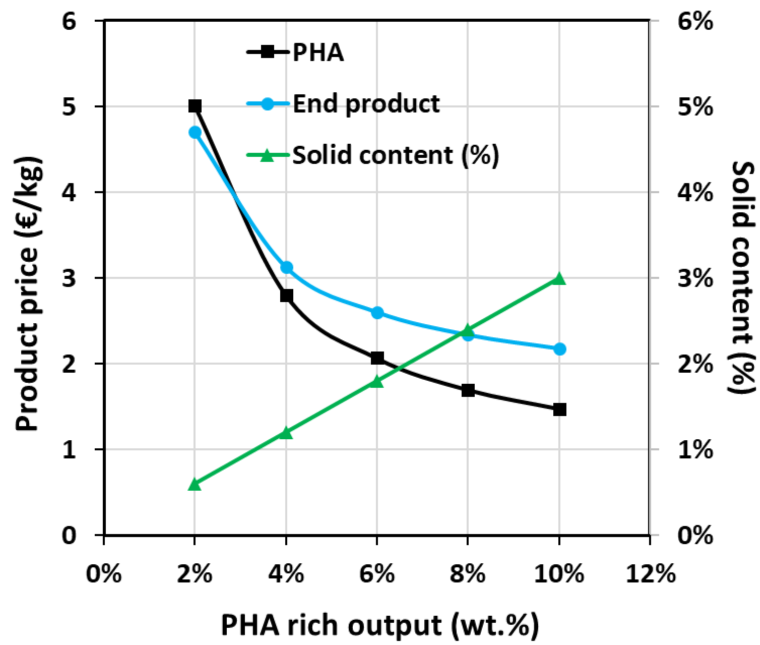

Figure 18. Impact of solid content on the MSP of PHA and end product

3.3.7. Impact of CAPEX

As mentioned earlier in section 3.1.2, the CAPEX estimated in the current assessment is less than one-third of that estimated in the PHARIO project. For the selection and accumulation reactor, the cost data were acquired from the literature as opposed to the PHARIO project where a company was hired which specialized in cost estimation. It can be concluded that the cost estimation in the PHARIO project was in more detail. However, the estimation was done for a plant capacity of 2000 tPHA/y and then multiplied by a factor of 2.5 to get the cost estimates of a plant capacity of 5000 tPHA/y. This makes the cost estimation inaccurate since it is well known that the cost increases with the scale by a scaling exponent as shown in Eq [1]. Nevertheless, the current capital cost estimation can be considered underestimated due to the lack of details available on the construction of a real large-scale plant. Therefore, the CAPEX was varied from -50% to +100% of the reference value and the results obtained are shown in Figure 19. When the CAPEX is doubled (+100% of reference value), the increase in the MSP of the PHA and the end product is 22% and 16%, respectively. Therefore, it is evident that the CAPEX is not the influential factor on the MSP as also shown in the Monte-Carlo analysis in section 3.3.1. The PHA MSP at double the CAPEX is €4.3 /kg, which is about 8% higher than the PHA market price (€4 /kg).

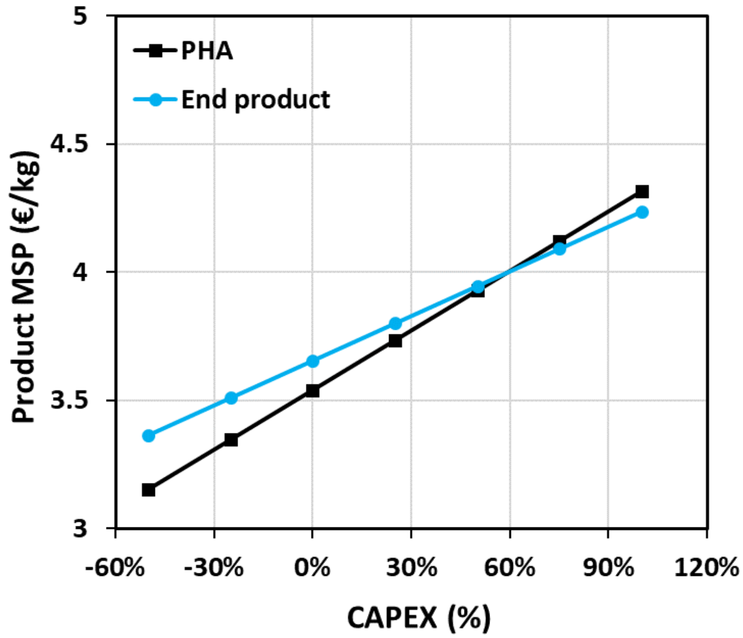

Figure 19. Impact of CAPEX on the MSP of PHA and end product

3.3.8. Impact of labor requirement

The labor requirement is a significant fixed operating cost. Therefore, if the plant is made automated as much as possible then the operating labor required can be reduced. The labor requirement in the current project was estimated based on the labor required for each equipment (Peters, Timmerhaus, and West 2003). However, due to advancements in technology, the plant operation can be made more automated requiring less human intervention. The downside is that it will increase the power consumption which might affect the overall plant costs. For the sake of simplicity, it had not been considered in this analysis. The variation MSP for the PHA and the end product are shown in Figure 20. The labor requirement was reduced up to 60% of the reference value to have a broad overview of its effect. The MSP of the PHA and the end product at -60% labor was €3.4 /kg and €3.6 /kg, respectively. This shows that by reducing the labor requirement by at least 50%, the profit margin could be up to just 14% when compared to the PHA market price (€4 /kg). However, as seen in the Monte-Carlo assessment, the labor requirement is the least influential parameter on the PHA production cost.

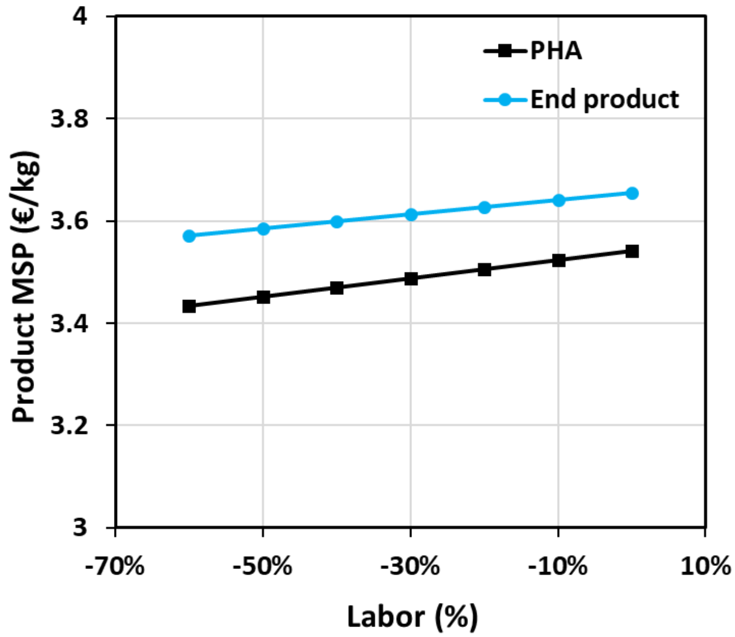

Figure 20. Impact of labor requirement on the MSP of PHA and end product

3.3.9. Impact of fuel price

The fuel is used to provide the heat and this is the second most significant parameter after PHA yield as shown in the Monte-Carlo analysis. The fuel considered in this study is natural gas which is burnt in a burner and the flue gas is used to provide heat, mainly in the dryer and some for the fermentation reactors. The fuel price is varied from -60% to +60% of the base case value (€34 /MWh) and the results obtained are shown in Figure 21. A 60% reduction in the fuel price results in a 12% reduction in the MSP of the PHA and an 8% reduction in the MSP of the end product. Alternatively, if there is the availability of waste heat at or near the STP, then this cost can be reduced considerably. The heat from the steam used during the extraction process and from the cooling of the recovered solvent can be recovered and used either to preheat the air/flue gas or directly used in the dryer. This will reduce the consumption of fuel significantly and consequently, reduce the products' MSP.

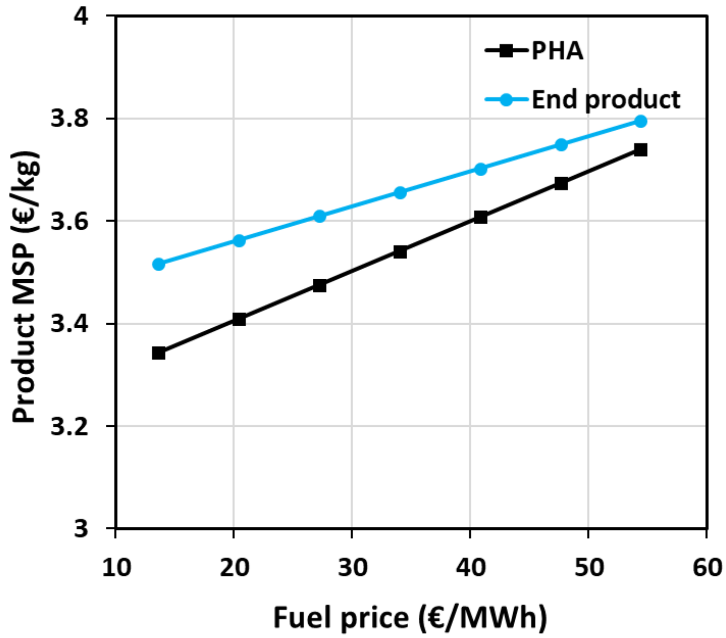

Figure 21. Impact of fuel price on the MSP of PHA and end product

3.3.10. Impact of blend ratio and raw material price

The blend ratio and the price of the raw material that is mixed with PHA to produce the end product are highly influential parameters. The material for the moulding process consists of 70% PHA and 30% other raw material (polymer). The raw material blend ratio is varied from 10% to 90% whereas the price is varied from 1 to 5 €/kg with €3 /kg as the base case price. It is interesting to note from Figure 22 that at the base case raw material price of €3.6 /kg, the MSP of the end product is almost the same for all the blend ratios. The difference between the lowest and the highest blend ratio is just €0.01 /kg. Towards the left of this point, lower blend ratios along with lower raw material price result in lower product MSP. The higher blend ratios in the left region tend to make the MSP closer to the raw material price. Similarly, towards the right, higher blend ratios along with higher raw material prices result in lower MSP. The higher blend ratios dominate and are the deciding factor for the product MSP. However, the choice of the raw material and the blend ratio depends on the desired product quality and the intended application.

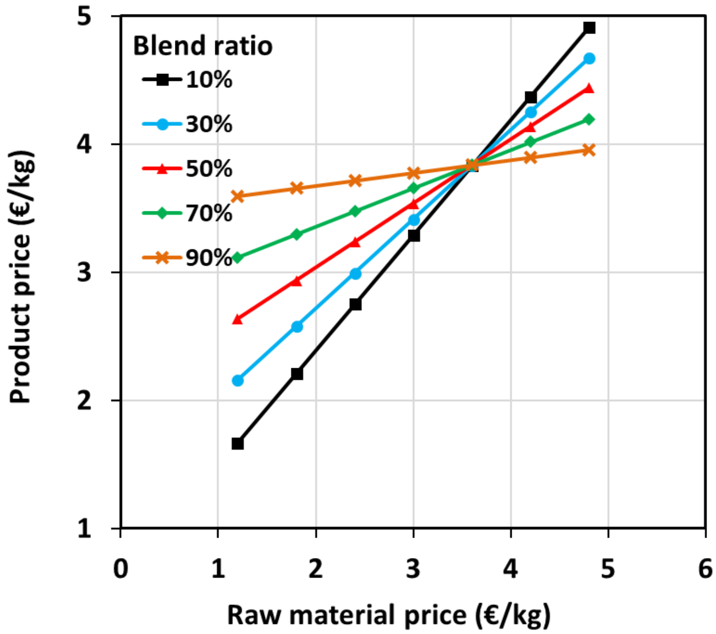

Figure 22. Impact of blend ratio and raw material price on the end product MSP

Chapter 4. Conclusions and future work

Municipal sewage water contains considerable amounts of organic carbon which can be useful carbon-based materials like cellulose or lipids or converted into fatty acids. Polyhydroxyalkanoates (PHA) are produced in special processes, but neither sewage nor sewage sludge contains this substance in larger amounts without enrichment. Utilizing these valuable materials could reduce the use of natural resources and subsequent carbon dioxide (CO_2) emissions and hence, realize a circular economy. In this report, a techno-economic assessment on the PHA production plant from primary sludge as feedstock was performed. In the WOW project, the value chain consists of several steps including acidogenic fermentation, biomass selection/enrichment, PHA accumulation, PHA separation, drying, extraction, and producing the end product. The PHA pilot plant until the drying step was developed by Technische Universität Kaiserslautern (TUK) and the dried PHA sent to Avans Hogeschool (Netherlands) for PHA extraction. The recovered raw PHA material is mixed with other materials and used in an injection moulding process to produce end products by NaturePlast (France). The MSP of the PHA estimated in this assessment is €3.54 /kg PHA which is in parity with that reported in the literature. Compared with the literature using other feed sources, the current MSP is much lower. The results from the PHARIO project estimated the PHA production price to be €3.6 /kg which is similar to the current estimation. A market potential study conducted in the WOW project estimated the PHA market price to be in the range of €3.5-4.5 /kg (Wupperverband 2020), which is considerably higher than the current assessment.

Several technical and economic parameters affect the overall performance of the plant. It can be observed that the MSPs of the PHA products decrease as the scale goes from small to large. With an increase in the plant scale, the equipment size was also increased by a specific exponential relationship particular to each equipment. This resulted in lower operating expenses per unit plant capacity accounting for the economies of the scale factor. The PHA yield is an intrinsic process parameter, which is independent of plant capacity. Though the reference value obtained from the pilot plant data was 35%, it was reported that the PHA yield can be altered by manipulating the carbon to nitrogen ratio, using inhibitors and novel fermentation strategies. From the Monte-Carlo analysis, it was evident that the PHA yield is the most significant factor that decides the PHA production cost. The solvent used in this project was dimethyl carbonate (DMC) since it is a green solvent, has high cell disruption efficiency, and has low toxicity. The reference cell disruption efficiency for the base case considered was 100% and it was reduced up to 80% to investigate the effect on the MSP. The results obtained showed that cell disruption efficiency has a significant effect on the MSP. The economic parameter, CAPEX is the key parameter that affected the MSP significantly. Therefore, if CAPEX is reduced then a considerable reduction in MSP can be expected. The reduction in labor requirement and fuel price does not have much effect on the MSP as observed from the Monte-Carlo analysis. The heat required during drying can reduce the MSP since much of the thermal energy used in this plant is for drying purposes. Thus, if an alternative drying technology such as a vacuum evaporator or a hybrid system is developed which consumes less thermal energy then there will be savings in fuel costs. Alternatively, a belt conveyor could be used to remove the moisture content by the action of gravity before the drying step. Moreover, the flue gases at the outlet are expected to be at a relatively higher temperature that could be used to recover heat and use it to preheat the air/flue gas or in the fermentation reactors. This will considerably reduce energy consumption. The blend ratio and the raw material price are the two main factors that affect the MSP of the end product. The PHA is mixed with another raw material in a ratio of 70%:30% as a base case. These two factors depend on the desired quality and the intended application of the end product. A cheaper raw material might not provide the best quality and vice versa. Therefore, an optimized value of blend ratio and the raw material price ensures the best quality product at a low cost.

The pilot plant developed in the WOW project is innovative and has a new design. Thus, there will be deviations of the estimated results from the actual operation of the plant. There will be a learning curve that will help in utilizing the resources efficiently and optimizing the plant operation. The performance of this value chain compared to other PHA production plants is promising and results in a positive business case under the assumptions made. It would be interesting to investigate further since it has the potential for reducing the MSPs by optimizing plant operation and efficient energy consumption.

Appendix

Mass balance of a large-scale PHA production plant

Acidogenic fermentation

Primary sludge = 105.4 m^3/h

Sludge density = 1000 kg/m^3

Operating hours = 8000 h/y

Heat requirement = 23.4 kWh/m^3

Electricity requirement = 96.9 kWh/m^3

$$Primary\ sludge = \frac{105.4 * 1000 * 8000}{1000} = 843{,}313\,\frac{t}{y}$$

Centrifuge1

VFA rich solution = 77 wt.%

Waste solids = 23 wt.%

Electricity requirement = 1.88 kWh/m^3

$$Fermented\ sludge\ in = 843{,}313\ \frac{t}{y}$$

$$VFA\ rich\ solution\ out\ = 0.77 * 843{,}313 = 649{,}351\ \frac{t}{y}$$

$$Waste\ solids\ out\ = 0.23 * 843{,}313 = 193{,}962\ \frac{t}{y}$$

Biomass selection

VFA solution split fraction to selection reactor = 36 wt.%

Excess sludge to VFA ratio = 0.5

Water to VFA ratio = 1

Electricity requirement = 2.51 kWh/m^3

$$VFA\ solution\ in = 0.36 * 649{,}351 = 233{,}766\ \frac{t}{y}$$

$$Excess\ sludge\ in = 0.5 * 233{,}766 = 116{,}883\ \frac{t}{y}$$

$$Water\ in = 1 * 233{,}766 = 233{,}766\ \frac{t}{y}$$

$$Enriched\ biomass\ out = 233{,}766 + 116{,}883 + 233{,}766 = 584{,}416\ \frac{t}{y}$$

PHA accumulation

VFA solution split fraction to accumulation reactor = 64 wt.%

Electricity requirement = 2.4 kWh/m^3

$$VFA\ rich\ solution\ in = 0.64 * 649{,}351 = 415{,}584\ \frac{t}{y}$$

$$Enriched\ biomass\ in = 584{,}416\ \frac{t}{y}$$

$$PHA\ rich\ biomass\ solution\ out = 415{,}584 + 584{,}416 = 1{,}000{,}000\ \frac{t}{y}$$

Centrifuge2

PHA rich biomass = 3 wt.% of input

Effluent = 97 wt.% of input

Electricity requirement = 1.88 kWh/m^3

$$PHA\ rich\ biomass\ solution\ in = 1{,}000{,}000\ \frac{t}{y}$$

$$PHA\ rich\ biomass\ out = 0.03 * 1{,}000{,}000 = 30{,}000\ \frac{t}{y}$$

$$Effluent\ out = 0.97 * 1{,}000{,}000 = 970{,}000\ \frac{t}{y}$$

Drying

PHA rich biomass DM = 30 wt.%

Moisture content after drying = 10 wt.%

Electricity requirement = 0.16 kWh/kg-evap

Heat requirement = 1.45 kWh/kg-evap

$$PHA\ rich\ biomass\ in = 30{,}000\ \frac{t}{y}$$

$$Dried\ PHA\ rich\ biomass\ out = \frac{0.3 * 30{,}000}{1 - 0.1} = 10{,}000\ \frac{t}{y}$$

$$Water\ vapor\ out = 30{,}000 - 10{,}000 = 20{,}000\ \frac{t}{y}$$

PHA extraction

Solvent to dried biomass ratio = 19

Electricity requirement = 0.006 kW/m^3

Steam requirement = 1.1 t/t dried PHA

$$Dried\ PHA\ rich\ biomass\ in = 10{,}000\ \frac{t}{y}$$

$$Dimethyl\ carbonate\ in = 10{,}000 * 19 = 190{,}000\ \frac{t}{y}$$

$$PHA\ rich\ solution\ out = 10{,}000 + 190{,}000 = 200{,}000\ \frac{t}{y}$$

Filter

PHA output (yield) = 50 wt.%

Cell disruption efficiency = 100%

Solvent recovery = 99.5%

$$PHA\ rich\ solution\ in = 200{,}000\ \frac{t}{y}$$

$$PHA\ rich\ filtrate\ out = (10{,}000 * 0.5 * 1) + (200{,}000 * 0.995) = 194{,}050\ \frac{t}{y}$$

$$PHA\ free\ biomass\ out = 200{,}000 - 194{,}050 = 5{,}950\ \frac{t}{y}$$

Evaporator

Steam requirement = 0.06 t/t filtrate

Cooling water = 0.39 t/t filtrate

$$PHA\ rich\ filtrate\ in = 194{,}050\ \frac{t}{y}$$

$$PHA = 10{,}000 * 0.5 * 1 = 5{,}000\ \frac{t}{y}$$

$$Recovered\ solvent = 194{,}050 - 5{,}000 = 189{,}050\ \frac{t}{y}$$

Compounding

PHA = 70 wt.% of total input

Raw material = 30 wt.% total input

Electricity requirement = 441.7 kWh/t

$$PHA = 5{,}000\ \frac{t}{y}$$

$$Raw\ material = 5{,}000 * \frac{0.3}{0.7} = 2{,}143 \frac{t}{y}$$

$$Waste\ (from\ injection) = 1{,}568 \frac{t}{y}$$

$$Granules\ out = 5{,}000 + 2{,}143 + 1{,}568 = 8{,}711 \frac{t}{y}$$

Injection moulding

Waste = 18 wt.% of input

Electricity requirement = 1503.4 kWh/t

$$Granules\ in = 8{,}711\,\frac{t}{y}$$

$$Product\ out = (1 - 0.18) * 8{,}711 = 7{,}143\,\frac{t}{y}$$

$$Waste\ (to\ compounding) = 0.18 * 8{,}711 = 1{,}568\,\frac{t}{y}$$

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