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# Economic and environmental analysis of PHAs production process

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**Abstract** Following the growing demand to improve both economic and environmental performance of PHAs production, the research focused on the evaluation of the economic and environmental performance of PHAs production process via aqueous two-phase extraction (ATPE). Thus, the process analysis of two processes with different PHAs purification and recovery strategies (which are with and without thermoseparating ATPE as primary purification step) was performed. Using the basis of 9000 tons PHAs production per year and 7920 operating hours, the process with thermoseparating ATPE as primary purification step standout in terms of both economic and environmental performance. PHA production cost of 5.77 US$/kg with a payback period of fewer than 4 years and ROI of 25.2% was achieved. The results showed that most of operating cost is contributed by facility-dependent cost and raw material cost, while the main contributor to raw material cost is carbon source. The insight from sensitivity analysis has demonstrated that the economic performance is sensitive to the fluctuation in surfactant cost. This proved that utilizing thermoseparating ATPE as primary recovery step not only helps to reduce chemical consumption, it also minimizes downstream equipment cost and wastewater treatment cost.

**Keywords** Economic evaluation · Environmental impacts · Life cycle assessment (LCA) · Polyhydroxyalkanoates (PHAs) · Waste Reduction (WAR) algorithm

## Abbreviations

| | |
|---|---|
| AP | Acidification or acid-rain potential |
| ATP | Aquatic toxicity potential |
| ATPE | Aqueous two-phase extraction |
| DCW | Dry cell weight |
| EOPO | Ethylene oxide–propylene oxide copolymer |
| GWP | Global warming potential |
| HTPE | Human toxicity potential by either inhalation or dermal exposure |
| HTPI | Human toxicity potential by ingestion |
| LCA | Life cycle assessment |
| NPCM | Non-PHAs cell mass |
| ODP | Ozone depletion potential |
| PEI | Potential environmental impact |
| PHA | Polyhydroxyalkanoates |
| POP | Photochemical oxidation or smog formation potential |
| SPI | Sustainable Process Index |
| TMS | Trace mineral solution |
| TTP | Terrestrial toxicity potential |
| WCO | Waste cooking oil |

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## Introduction

Aqueous two-phase extraction (ATPE) serves as a primary integrated recovery step of isolation, purification and concentration for polyhydroxyalkanoates (PHAs), which partially removes the impurities from the target product and reduces the subsequent downstream processing volume (Kepka et al. 2003). Other than providing a mild environment for bioseparation, ATPE also utilizes safe and relatively environmental-friendly phase-forming components, consumes little time and energy, and the scale-up can be predicted reliably from experimental data. Thermoseparating ATPE is an advanced technique of ATPE that utilizes thermoseparating polymer such as ethylene oxide–propylene oxide copolymer (EOPO) which its solution thermoseparates into two phases as temperature increases. However, there are doubts that does utilization of thermoseparating ATPE as primary recovery step really provide an economic competitive edge as well as arguments that it is not environmental-friendly enough (Ng et al. 2012). Henceforth, it is essential to design the PHAs production process using thermoseparating ATPE as primary recovery strategy and evaluate its performance as well as compare with other isolation and purification strategies as there has not been any study reported on this to date. A detailed economic analysis has to be performed to evaluate the economic performance of PHAs purification and recovery via thermoseparating ATPE.

On the other hand, for environmental performance evaluation, life cycle analysis (LCA) has been widely used for environmental performance assessment of products and services. However, LCA can be very tedious or not suitable to be applied at some process design stages due to insufficient data and the needs of rigorous calculation. Therefore, some researchers focused on certain part of the life cycle by applying the concept of LCA and sets of indicators and indexes have been developed based on the sustainability concept, such as Sustainable Process Index (SPI) (Krotscheck and Nardolawsky 1996), Waste Reduction (WAR) algorithm (Young and Cabezas 1999), Environmental Fate and Risk Assessment Tool (Shonnard and Hiew 2000), Eco-indicator 99 (Goedkoop and Spriensma 2001), GreenPro (Khan et al. 2001), IChemE sustainability metrics (IChemE 2002) and others. The researchers have also applied LCA on PHAs biosynthesis, compared with conventional plastics and other bioplastic, and from many different points of view (Hyde 1998; Grengross and Slater 2000; Akiyama et al. 2003; Harding et al. 2007; Kim and Dale 2005, 2008; Pietrini et al. 2007; Gurieff and Lant 2007; Heimersson et al. 2014; Khoo and Tan 2010; Zhong et al. 2009). Nevertheless, agreed with literature (Heimersson et al. 2014), researchers have always only reported

environmental impacts in terms of carbon dioxide equivalent emissions (global warming potential) and/or energy consumptions, especially the early LCA studies of PHAs (Gurieff and Lant 2007; Miller et al. 2013; Fernandez-Dacosta et al. 2015a, b). These are not comprehensive enough and hardly cover every aspect, including smog formation, ozone depletion, acid-rain formation, water pollution, and other pollutions.

In this paper, both economic and environmental performance of PHAs production process with thermoseparating ATPE as primary recovery step was evaluated and compared with other purification and recovery strategy. Economic and environmental analysis of the production and recovery of PHAs by thermoseparating ATPE will allow for the industrial-scale PHA production price estimation and for the evaluation of the economic and environmental feasibility of PHA production. Computer-aided process simulation has been utilized as a planning, design, optimization and evaluation tool of chemical process. Full-scale process flowsheets of PHA production processes were drawn, and full-scale plants have been designed with simulation basis and explanation on the process flow provided. Detailed cost analysis was utilized to calculate equipment, raw material and utility cost, PHAs selling price, and profitability analysis as well as identify economy "hot-spots". WAR algorithm which was selected as the indicator to evaluate environmental performance not only requires minimum amount of data, but it also covers a wide range of environmental aspects.

## Materials and methods

### Process analysis tool

To develop the processing flowsheet, the conceptual process design followed a hierarchical design strategy based on heuristics and experience. Process analysis and economic evaluation for PHAs biosynthesis and recovery were carried out using SuperPro Designer v6.0 from Intelligen Inc. (Scotch Palins, NJ, USA). Microbial biomass was modeled as dextrose, while PHA was modeled as repeating unit of PHA ($C_4H_6O_2$) and the analysis did not account for polymerization energy. Two different process flowsheets for the biosynthesis and recovery of PHAs are shown in Figs. 1 and 2. The first one is extractive bioconversion of PHAs via thermoseparating ATPS coupled with recovery using surfactant and sodium hypochlorite digestion, while the latter is a conventional PHAs production with recovery using surfactant and sodium hypochlorite digestion. Some equipment and other secondary pieces such as valves and piping are omitted for simplicity in the flowsheet but are taken into consideration in the economic analysis. Pumps
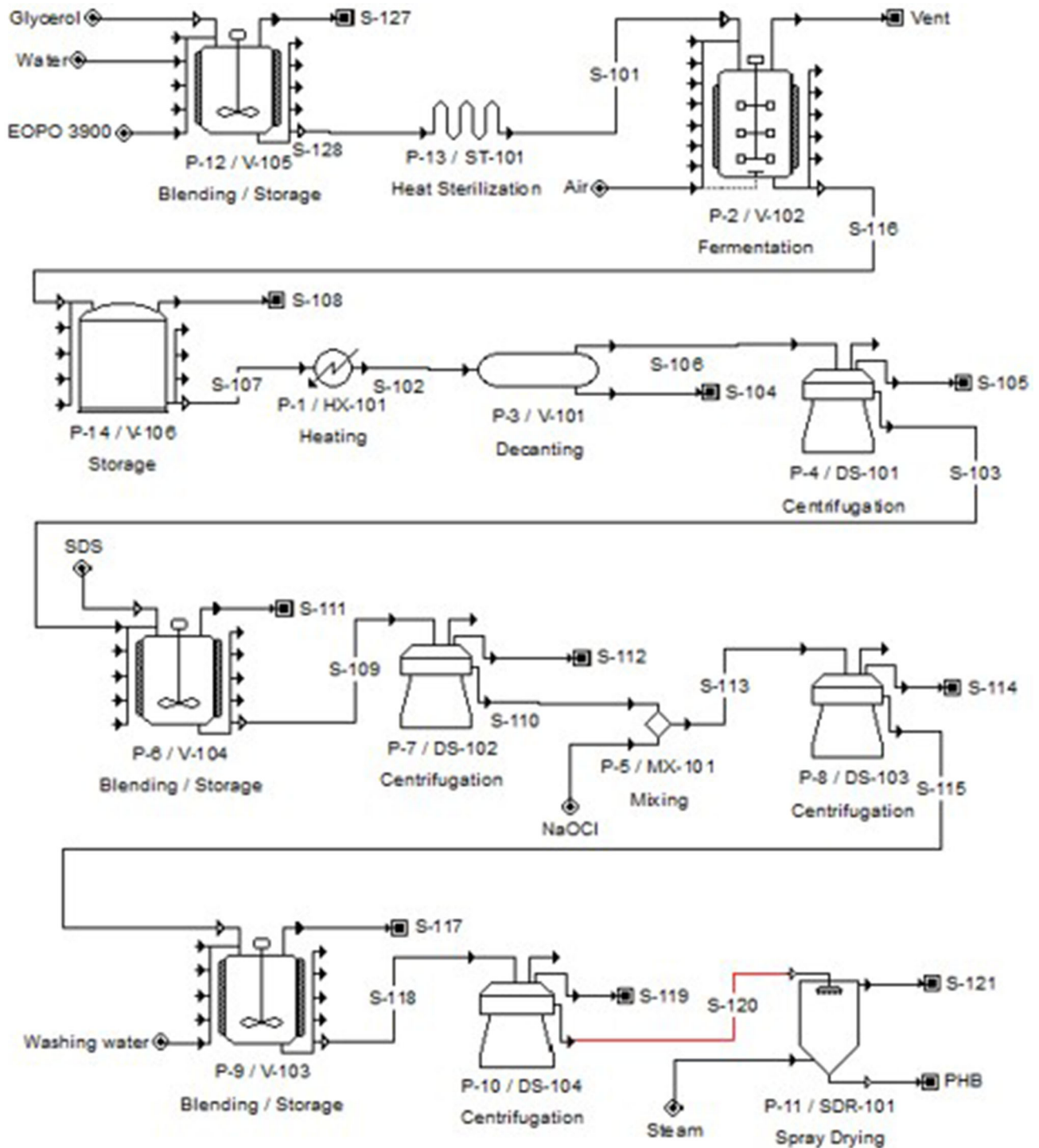
Fig. 1 Process flowsheet of PHAs production using recovery by thermoseparation coupled with surfactant–hypochlorite digestion (process 1)

are provided wherever necessary. For all assessment, the system boundary only covered the "gate-to-gate" analysis from upstream PHAs biosynthesis to the final product of PHAs powder formation. The production of raw materials as well as utilization and disposal of products is assumed to be same for all alternatives.

**Raw materials**

The media of PHAs biosynthesis consists of glycerol (as carbon source) and various inorganic salts. The cost of inorganic salts was excluded due to the insignificant share in the final production cost of PHAs (Choi and Lee 1997).
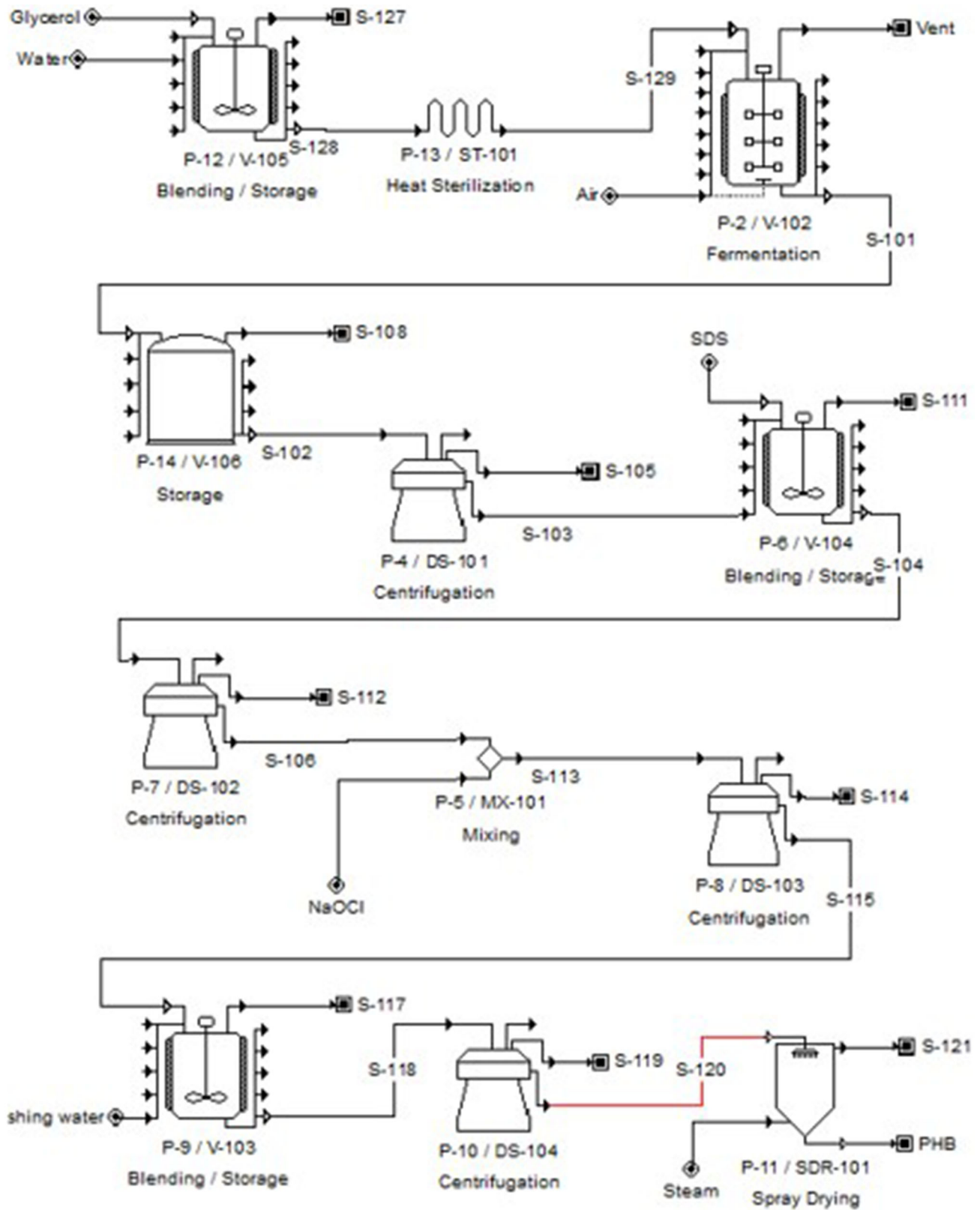
**Fig. 2** Process flowsheet of PHAs production using recovery by surfactant–hypochlorite digestion (process 2)

*Cupriavidus necator* H16 was the bacterial strain selected for this analysis. Thermoseparating polymer used in this paper is poly(ethylene glycol-ran-propylene glycol) monobutyl ether Mr—3900 ($EO_{50}PO_{50}$), also known as EOPO 3900.

## Fermentation

Fermentation medium was prepared in a blending tank (P-12) which was sterilized in a continuous heat sterilizer (P-13). Then, the medium and air were supplied to fermenter (P-2) for 42 h of fermentation period at 30 °C. After the cultivation, fermentation broth was moved to the holding tank (P-14). The PHAs yield for the both process is same given the identical upstream fermentation condition.

## Recovery of PHAs by thermoseparating ATPE coupled with surfactant–hypochlorite digestion (Fig. 1)

Culture broth collected in the holding tank was transferred to the thermoseparation tank (P-3 and P-4) for thermoseparating ATPE to take place at 60 °C for 15 min residence time. The PHAs after thermoseparation achieved a purification factor of 1.42 and recovery yield of 94.8% based on results of the previous work. 95% of thermoseparating polymer was assumed to thermoseparate to the bottom phase which makes up approximately 60 wt/wt% of bottom phase with the remainder being mostly water. Afterward, the bottom phase was sent for wastewater treatment, while the top phase consisted of mostly water, and the PHAs product was transferred to a centrifuge (P-4) to separate PHAs from aqueous solution. Centrifugation also removes soluble cellular materials (especially nucleic acids) to improve the efficiency of subsequent downstream processing and reduce the viscosity of process fluid (Van Wegen et al. 1998).

Following that, surfactant solution (10% w/v) was added to the outflow of aqueous solution from centrifuge with surfactant-to-biomass ratio 1:3 and mixed at 55 °C for 15 min of mean residence time to obtain recovery yield of 86.6% and purity of 98% (Dong and Sun 2000). PHAs were then separated from the aqueous solution containing dissolved NPCM by centrifugation (P-7). The PHAs were then cleaned with 30% wt/v sodium hypochlorite solution to achieve purity of 99% with a yield of 95% (Fernandez-Dacosta et al. 2015a; Ramsay et al. 1990). After centrifugation (P-8), the PHAs granules were then washed with an equal volume of water (P-9 and P-10) (Yu and Chen 2006) and were finally spray-dried (P-11) using continuous spray drying system to obtain PHAs products of 99.9 wt/wt % purity. This process will be referred as process 1 throughout the paper.

## Recovery of PHAs by surfactant–hypochlorite digestion (Fig. 2)

The process is similar as above except the culture broth was directly transferred to continuous centrifugation (P-4) after the holding tank skips the thermoseparating ATPE step. The PHAs recovery strategy of surfactant–hypochlorite was employed in this study as it was demonstrated to be more economical and environmental-friendly than the others (Choi and Lee 1997; Jacquel et al. 2008; Posada et al. 2011; Fernandez-Dacosta et al. 2015b). This process will be referred as process 2 throughout the paper.

## Economic analysis

All costs in this paper are given in 2016 US$. This analysis was estimated for a 10-year period at an annual interest rate of 6%, considering the straight line depreciation method and a 40% income tax. The price of carbon source and other major raw materials was as follows in US$/kg: glycerol, 0.5; EOPO 3900, 1.0; surfactant (SDS), 0.8; hypochlorite solution, 0.2; water, 0.0012. Also, the price of electricity, water and low-pressure vapor used was 0.03044 US$/kWh, 0.0004 US$/kg and 0.0042 US$/kg, respectively. The labor cost of operatives was assumed to be 25US$/man-hour. SuperPro Designer v6.0 has been used to estimate the capital costs of process units and the operating cost as well as perform the profitability analysis.

## Environmental assessment

Based on the potential environmental impact (PEI) balance concept introduced by Hilaly and Sikdar (Hilaly and Sikdar 1995), Young and Cabezas (Young and Cabezas 1999) have introduced WAR algorithm that focuses on waste minimization across the process boundary. With an extensive chemicals database, WAR algorithm covers four local toxicological impact categories which are human toxicity potential by ingestion (HTPI), terrestrial toxicity potential (TTP), human toxicity potential by either inhalation or dermal exposure (HTPE), and aquatic toxicity potential (ATP), and four global atmospheric impacts which are global warming potential (GWP), ozone depletion potential (ODP), photochemical oxidation or smog formation potential (POP), and acidification or acid-rain potential (AP). Using WAR algorithm, the environmental performance of both processes was evaluated with the most potential hazard equipment identified.

## Sensitivity analysis

Sensitivity analysis was conducted to investigate the influence of different operating costs on the PHAs price

and to identify the opportunity for further improvement. For these purposes, four different cases were considered where the most critical and influencing factors were manipulated using spreadsheet software, Excel 2013 (Microsoft, Redmond, WA, USA).

## Results and discussion

### Material and energy flows

The targeted amount of PHAs production in the simulations was 9000 tons per year. The total operating hour was assumed to be 7920 h per year. Considering 42 h of fermentation period added with the 12 h of turnaround time required for cultivation tank cleaning and recharging, 146 fermentation runs can be carried out per year. These values were used to determine the material flow as well as estimate the sizes of the processing equipment. Table 1 presents the summary of material and energy flows for both processes, while Tables 2 and 3 tabulate the summary of material flow in each stream. Based on the process flowsheet, 190.3 tons of glycerol is required as carbon source per batch to achieve 9000 tons of annual PHAs production. It is clearly shown that process 2 requires almost 10-fold of surfactant and hypochlorite solution compared to process 1. For utility usage, both processes consume slightly over 25,000 MWh of electricity, 129 kilotons of chilled water, and approximately 50 kilotons of steam per year. For material flow, the two processes have similar mass flowrate at the fermentations stage: Process 1 has additional EOPO 3900, while process 2 has a higher mass flowrate at the downstream processing due to a large amount of surfactant and hypochlorite solution utilized.

### Economic analysis

From Figs. 1 and 2, it can be observed that the fermentation and recovery processes were decoupled by integrating

a holding tank preceding the recovery process. Thus, the entire purification process can be carried out continuously during the fermentation plus its turnaround period. Tables 4 and 5 summarize the specification and purchase costs of the major equipment for both processes. To reduce the equipment cost and optimize the operating cost, the sizes of the process unit in the isolation and recovery operations were adjusted so that the entire recovery process could be operated in 54 h. As shown from the tables, much of the capital cost is due to the fermenter and the centrifuges. Based on these results, the fixed capital cost, the annual operating cost, and profitability analysis were estimated and are summarized in Tables 6 and 7, respectively. The capital cost included cost of rotary and static equipment, piping, wastewater treatment system, etc. On the other hands, the operation and maintenance (O&M) cost included cost of chemicals, manpower, utilities, wastewater treatment, maintenance and insurance. For the direct cost, total (f.o.b) equipment costs were calculated to be USD\$ 22.5 million and USD\$ 27.5 million for processes 1 and 2, respectively. The capital cost of processes 1 and 2 is USD\$ 130.6 million and USD\$ 160.1 million, respectively, while the annual operating cost for processes 1 and 2 is USD\$ 51.7 million and USD\$ 57.9 million, respectively.

The profitability analysis was done based on the assumption of selling price of PHAs at 10.0 US\$/kg. The production cost of PHAs for processes 1 and 2 is 5.77 and 6.12 US\$/kg, respectively, at this production scale. The price obtained was comparable to that reported in the literature, with examples of 5.35 US\$/kg at production scale of 4300 tons for PHA production from *E. coli* using dairy whey (Van Wegen et al. 1998), 6.14 US\$/kg at production scale of 2850 tons (Choi and Lee 1997), 8.32 US\$/kg at production scale of 100,000 tons (Lee and Choi 1998), 5.35–10.7 USD\$/kg for PHA production from whey (Koller et al. 2007), and 5.38 USD\$/kg at fermentation capacity of 1000 m³/day (Mudliar et al. 2007). Due to the high amount of surfactant and sodium hypochlorite used in process 2, the wastewater treatment cost also doubled

**Table 1** Material and energy flows per batch for 9000 tons PHAs production per annum using two different recovery strategies

| Input (ton/batch) | Process 1 | Process 2 | Outputs (ton/batch) | Process 1 | Process 2 |
|---|---|---|---|---|---|
| Glycerol | 190.3 | 190.3 | Carbon dioxide | 29.5 | 29.5 |
| Oxygen | 43.3 | 43.3 | Biomass | 70.3 | 70.4 |
| Surfactant (SDS) | 3.1 | 46.3 | Oxygen | 1.5 | 1.5 |
| Hypochlorite solution | 0.9 | 9.7 | PHAs | 68.4 | 68.5 |
| Water | 576.4 | 1031.70 | Water | 639.9 | 1095.30 |
| EOPO 3900 | 38.2 | 0.0 | | | |
| Energy (per year) | | | | | |
| Chilled water (ton) | 129,302 | 129,267 | | | |
| Steam (ton) | 52,630 | 47,914 | | | |
| Electricity (MWh) | 25,602 | 27,290 | | | |

**Table 2** Material flows for each stream for a process 1 (in tons/batch)

| Chemicals | Glycerol | Water | EOPO 3900 | S-128 | S-101 | Air | Vent | S-116 | S-107 |
|---|---|---|---|---|---|---|---|---|---|
| PHAs | | | | | | | | 68.4 | 68.4 |
| Biomass | | | | | | | | 70.3 | 70.3 |
| Water | | 535.6 | | 535.6 | 535.6 | | 30 | 569.2 | 569.2 |
| Glycerol | 190.3 | | | 190.3 | 190.3 | | | | |
| Oxygen | | | | | | 43.3 | 1.5 | | |
| CO$_2$ | | | | | | | 29.5 | | |
| EOPO 3900 | | | 38.2 | 38.2 | 38.2 | | | 38.2 | 38.2 |

| Chemicals | S-102 | S-106 | S-104 | S-105 | S-103 | SDS | S-109 | S-112 | S-110 |
|---|---|---|---|---|---|---|---|---|---|
| PHAs | 68.4 | 64.8 | 3.6 | | 64.8 | | 64.8 | 8.7 | 56.2 |
| Biomass | 70.3 | 27.7 | 42.6 | | 27.7 | | 27.7 | 26.6 | 1.2 |
| Water | 569.2 | 514.7 | 54.5 | 514.7 | 0.05 | 27.8 | 27.8 | 27.9 | |
| SDS | | | | | | 3.1 | 3.1 | 3.1 | |
| EOPO 3900 | 38.2 | 1.9 | 36.3 | 1.9 | | | | | |

| Chemicals | NaOCl | S-113 | S-114 | S-115 | Washing water | S-118 | S-119 | S-120 | PHB |
|---|---|---|---|---|---|---|---|---|---|
| PHAs | | 56.2 | 2.8 | 53.4 | | 53.4 | | 53.4 | 53.4 |
| Biomass | | 1.2 | 0.6 | 0.6 | | 0.6 | | 0.6 | 0.6 |
| Water | 2.1 | 2.1 | 2.1 | | 54.8 | 54.8 | 54.8 | | |
| NaOCl | 0.9 | 0.9 | 0.9 | | | | | | |

**Table 3** Material flows for each stream for process 2 (in tons/batch)

| Chemicals | Glycerol | Water | S-128 | S-129 | Air | S-101 | Vent | S-102 |
|---|---|---|---|---|---|---|---|---|
| PHAs | | | | | | 68.5 | | 68.5 |
| Biomass | | | | | | 70.4 | | 70.4 |
| H$_2$O | | 535.6 | 535.6 | 535.6 | | 569.3 | 30 | 569.3 |
| Glycerol | 190.3 | | 190.3 | 190.3 | | | | |
| O$_2$ | | | | | 43.3 | | 1.5 | |
| CO$_2$ | | | | | | | 29.5 | |
| NaOCl | | | | | | | | |

| Chemicals | S-105 | S-103 | SDS | S-104 | S-112 | S-106 | NaOCl | S-113 |
|---|---|---|---|---|---|---|---|---|
| PHAs | | 68.5 | | 68.5 | 9.2 | 59.3 | | 59.3 |
| Biomass | | 70.4 | | 70.4 | 69.2 | 1.2 | | 1.2 |
| H$_2$O | 559.2 | | 416.6 | 416.6 | 416.6 | 0.04 | 22.6 | 22.6 |
| SDS | | | 46.3 | 46.3 | 46.3 | | | |
| NaOCl | | | | | | | 9.7 | 9.7 |

| Chemicals | S-114 | S-115 | Washing water | S-118 | S-119 | S-120 | PHB |
|---|---|---|---|---|---|---|---|
| PHAs | 3 | 56.3 | | 56.3 | | 56.3 | 56.3 |
| Biomass | 0.6 | 0.6 | | 0.6 | | 0.6 | 0.6 |
| H$_2$O | 22.6 | | 56.9 | 56.9 | 56.9 | | |
| NaOCl | 9.7 | | | | | | |

compared to process 1. Therefore, it proved that utilizing thermoseparation-based ATPE as primary recovery step not only helps to reduce chemical cost, it also minimizes downstream equipment cost and wastewater treatment cost.

At the selling price assumed, it was shown that PHAs production using thermoseparation recovery strategy has a payback period of fewer than 4 years and ROI of 25.2%, while the surfactant–hypochlorite recovery strategy

**Table 4** Major equipment specification and purchase cost (based on 2016 year) for process 1

| Equipment (quantity) | Description | Cost ($) |
|---|---|---|
| Fermenter P-2 (1) | Volume = 321.3 m$^3$ | 9,774,000 |
| | Power = 2189 kW | |
| Blending tank P-12 (1) | Volume = 80.9 m$^3$ | 506,000 |
| Heat sterilizer P-13 (1) | Diameter = 1.2 m | 758,000 |
| | Length = 12.8 m | |
| Disk-Stack Centrifuge P-4 (6) | Sigma factor = 188,335 m$^2$ | 3,498,000 |
| | Power = 648 kW | |
| Flat-bottom tank P-14 (2) | Volume = 117.6 m$^3$ | 166,000 |
| Spray dryer P-11 (1) | Dryer diameter = 0.127 m | 115,000 |
| | Dryer height = 0.382 m | |
| Thermoseparation tank P-1 and P-3 (1) | Volume = 46.2 m$^3$ | 278,000 |
| Blending tank P-6 (2) | Volume = 67.7 m$^3$ | 948,000 |
| | Power = 12 kW | |
| Disk-stack centrifuge P-7 (3) | Sigma factor = 154,435 m$^2$ | 1,569,000 |
| | Power = 121 kW | |
| Disk-stack centrifuge P-8 (2) | Sigma factor = 196,405 m$^2$ | 1,192,000 |
| | Power = 76 kW | |
| Blending tank P-9 (1) | Volume = 36.8 m$^3$ | 390,000 |
| | Power = 6 kW | |
| Disk-stack centrifuge P-10 (2) | Sigma factor = 145,513 m$^2$ | 1,014,000 |
| | Power = 65 kW | |

**Table 5** Major equipment specification and purchase cost (based on 2016 year) for process 2

| Equipment (quantity) | Description | Cost ($) |
|---|---|---|
| Fermenter P-2 (3) | Volume = 305.3 m$^3$ | 9,519,000 |
| | Power = 2073 kW | |
| Blending tank P-12 (1) | Volume = 76.6 m$^3$ | 496,000 |
| Heat sterilizer P-13 (1) | Diameter = 0.8 m | 747,000 |
| | Length = 15.6 m | |
| Disk-stack centrifuge P-4 (6) | Sigma factor = 214,650 m$^2$ | 3,750,000 |
| | Power = 365 kW | |
| Flat-bottom tank P-14 (1) | Volume = 111.5 m$^3$ | 159,000 |
| Spray dryer P-11 (1) | Dryer diameter = 0.3 m | 115,000 |
| | Dryer height = 0.9 m | |
| Blending rank P-6 (9) | Volume = 73.5 m$^3$ | 4,392,000 |
| | Power = 60 kW | |
| Disk-stack centrifuge P-7 (5) | Sigma factor = 218,230 m$^2$ | 3,155,000 |
| | Power = 285 kW | |
| Disk-stack centrifuge P-8 (2) | Sigma factor = 132,394 m$^2$ | 964,000 |
| | Power = 65 kW | |
| Blending tank P-9 (1) | Volume = 63.8 m$^3$ | 465,000 |
| | Power = 6 kW | |
| Disk-stack centrifuge P-10 (2) | Sigma factor = 144,942 m$^2$ | 1,012,000 |
| | Power = 67 kW | |

requires more than 4½ years with ROI of 21.8%. This is due to the higher facility-dependent cost which includes depreciation, equipment maintenance, insurance, local taxes, and factory expenses.

Figure 3a, b depicts the breakdown of operating cost for PHAs production using the two different recovery strategies. For process 1, contribution of both raw material cost and facility-dependent cost is almost the same (45 and

**Table 6** Economic analysis of process 1

| | Cost ($) |
|---|---:|
| *Fixed capital estimate summary (2016 prices)* | |
| (A) Total plant direct cost (TPDC) (physical cost) | |
| (a) Equipment purchase cost (PC) | 22,454,000 |
| (b) Installation (summed over all units) | 6,287,000 |
| (c) Process piping (0.35 × PC) | 7,859,000 |
| (d) Instrumentation (0.40 × PC) | 8,982,000 |
| (e) Insulation (0.03 × PC) | 674,000 |
| (f) Electrical (0.10 × PC) | 2,245,000 |
| (g) Buildings (0.45 × PC) | 10,104,000 |
| (h) Yard improvement (0.15 × PC) | 3,368,000 |
| (i) Auxiliary facilities (0.40 × PC) | 8,982,000 |
| | 70,955,000 |
| (B) Total plant indirect cost (TPIC) | |
| (a) Engineering (0.25 × TPDC) | 17,739,000 |
| (b) Construction (0.35 × TPDC) | 24,834,000 |
| | 42,573,000 |
| (C) Other costs (OTC) | |
| (a) Contractor's fee [0.05 × (TPDC + TPIC)] | 5,676,000 |
| (b) Contingency [0.10 × (TPDC + TPIC)] | 11,353,000 |
| | 17,029,000 |
| (D) Direct fixed capital (DFC) | |
| TPDC + TPIC + OTC | 130,557,000 |
| *Annual operating cost (2016 price)* | |
| (A) Raw materials | |
| (a) Glycerol | 15,982,000 |
| (b) Water | 116,000 |
| (c) Surfactant (SDS) | 415,000 |
| (d) Hypochlorite solution | 31,000 |
| (e) EOPO 3900 | 6,418,000 |
| | 22,962,000 |
| (B) Labor-dependent items | |
| (a) Lumped operating labor (57,370 h × 57.5 $/h) | 3,299,000 |
| (b) Laboratory (0.15 × a) | 494,850 |
| | 3,793,850 |
| (C) Facility-dependent items | 23,858,000 |
| (D) Other consumables | 0 |
| (E) Utilities | 1,052,081 |
| (F) Waste treatment/disposal | 61,000 |
| Total annual operating cost | 51,726,931 |
| *Profitability analysis (2016 prices)* | |
| (A) Direct fixed capital | 130,557,000 |
| (B) Working capital | 2,490,000 |
| (C) Startup cost | 6,528,000 |
| Total investment | 139,575,000 |

| | Cost ($/year) |
|---|---:|
| (E) Revenues | |
| PHAs | 89,619,000 |
| (F) Annual operating cost | 51,727,000 |

**Table 6** continued

| | Cost ($/year) |
|---|---:|
| (G) Gross profit (E–F) | 37,892,000 |
| (H) Taxes (40%) | 15,156,800 |
| (I) Net profit (G − H + depreciation) | 35,138,000 |
| (J) Return on investment (ROI) | 25.18% |
| (K) Payback period | 3.97 years |
| (L) IRR (after taxes) | 21.48% |
| (M) NPV (at 7.0% interest) | 130,369,000 $ |

46%, respectively). The high raw material cost is due to the high cost of EOPO 3900 which can be further reduced if it is recycled as EOPO 3900 can be recycled up to twice as reported in the previous study. On the other hand, facility-dependent cost dominated operating cost for process 2 with 51% share, while raw material cost contributed 39%. The cost of the carbon source also contributes significantly as glycerol alone was as high as 30.9 and 27.6% of total operating cost for processes 1 and 2, respectively. This was comparable to those as reported in the literature, such as 30.7% for PHAs production by recombinant *E. coli* utilizing glucose as carbon source (Choi and Lee 1997) and up to 45% as mentioned by Posada and coworkers (Posada et al. 2011). Henceforth, utilizing a cheaper carbon source such as cheese whey, waste cooking oil, cane and beet molasses, sludge waste, POME, and others has the potential to reduce the final PHAs production price significantly. Nevertheless, it must be kept in mind that PHAs content, productivity, yield, and downstream recovery process should be considered as well when using cheaper carbon source.
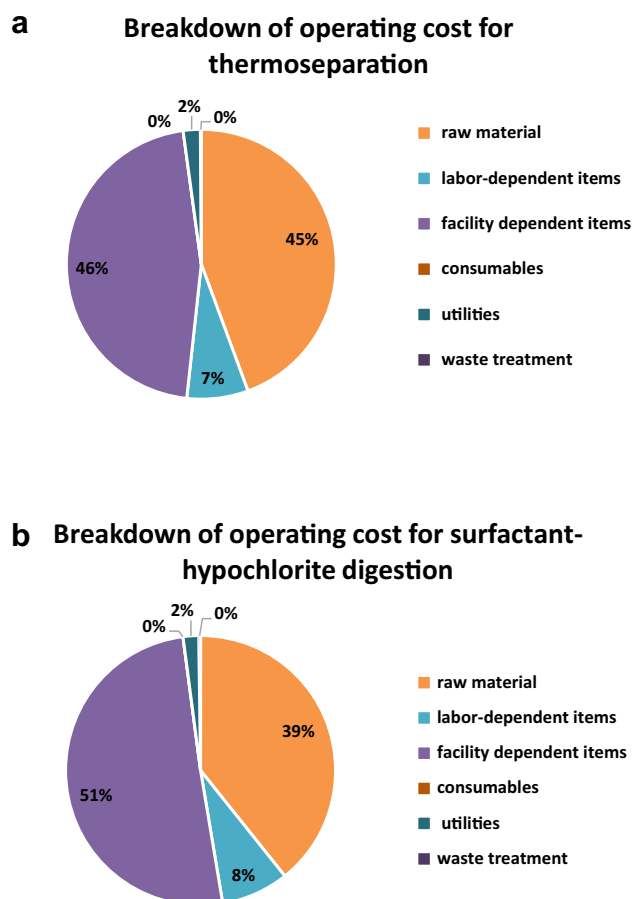
## Environmental analysis

Following the rising concern of community and government on the environment, minimizing the environmental impact of PHA production process is a necessity other than achieving economic competitiveness. Among range of indicators stated in "Introduction" section, SPI (Koller et al. 2013), Tools for the Reduction and Assessment of Chemical and Other Environmental Impacts model (Kim and Dale 2005), CML 2 Baseline 2000 v2.03 assessment method (Harding et al. 2007), and Eco-indicator 99 (Zhong et al. 2009) have been used to evaluate the environmental performance of PHAs production aside from those that utilized complete LCA study. In this research, environmental analysis was done to evaluate and compare the environmental performance of both process using WAR

**Table 7** Economic analysis of process 2

| | Cost ($) |
|---|---|
| *Fixed capital estimate summary (2016 prices)* | |
| (A) Total plant direct cost (TPDC) (physical cost) | |
|   (a) Equipment Purchase Cost (PC) | 27,535,000 |
|   (b) Installation (summed over all units) | 7,710,000 |
|   (c) Process piping ($0.35 \times$ PC) | 9,637,000 |
|   (d) Instrumentation ($0.40 \times$ PC) | 11,014,000 |
|   (e) Insulation ($0.03 \times$ PC) | 826,000 |
|   (f) Electrical ($0.10 \times$ PC) | 2,753,000 |
|   (g) Buildings ($0.45 \times$ PC) | 12,391,000 |
|   (h) Yard improvement ($0.15 \times$ PC) | 4,130,000 |
|   (i) Auxiliary facilities ($0.40 \times$ PC) | 11,014,000 |
| | 87,010,000 |
| (B) Total plant indirect cost (TPIC) | |
|   (a) Engineering ($0.25 \times$ TPDC) | 21,752,000 |
|   (b) Construction ($0.35 \times$ TPDC) | 30,453,000 |
| | 52,205,000 |
| (C) Other costs (OTC) | |
|   (a) Contractor's fee [$0.05 \times$ (TPDC + TPIC)] | 6,961,000 |
|   (b) Contingency [$0.10 \times$ (TPDC + TPIC)] | 13,922,000 |
| | 20,883,000 |
| (D) Direct fixed capital (DFC) | |
|   TPDC + TPIC + OTC | 160,098,000 |
| *Annual operating cost (2016 prices)* | |
| (A) Raw materials | |
|   (a) Glycerol | 15,982,000 |
|   (b) Water | 208,000 |
|   (c) Surfactant (SDS) | 6,221,000 |
|   (d) Hypochlorite solution | 325,000 |
| | 22,736,000 |
| (B) Labor-dependent items | |
|   (a) Lumped operating labor (70,861 h $\times$ 57.5 $/h) | 4,075,000 |
|   (b) Laboratory ($0.15 \times$ a) | 611,250 |
| | 4,686,250 |
| (C) Facility-dependent items | 29,256,000 |
| (D) Other consumables | 0 |
| (E) Utilities | 1,083,663 |
| (F) Waste treatment/disposal | 122,000 |
| Total annual operating cost | 57,883,913 |
| *Profitability analysis (2016 price)* | |
| (A) Direct fixed capital | 160,098,000 |
| (B) Working capital | 2,555,000 |
| (C) Startup cost | 8,005,000 |
| Total investment | 170,658,000 |

| | Cost ($/year) |
|---|---|
| (E) Revenues | |
|   PHAs | 94,656,000 |
| (F) Annual operating cost | 57,883,000 |
| (G) Gross profit (E − F) | 36,772,000 |

**Table 7** continued

| | Cost ($/year) |
|---|---|
| (H) Taxes (40%) | 14,708,800 |
| (I) Net profit (G − H + depreciation) | 37,273,000 |
| (J) Return on investment (ROI) | 21.84% |
| (K) Payback period | 4.58 years |
| (L) IRR (after taxes) | 18.05% |
| (M) NPV (at 7.0% interest) | 117,213,000 $ |

**a Breakdown of operating cost for thermoseparation**

Legend: raw material, labor-dependent items, facility dependent items, consumables, utilities, waste treatment

Values: 45%, 7%, 46%, 0%, 2%, 0%

**b Breakdown of operating cost for surfactant-hypochlorite digestion**

Legend: raw material, labor-dependent items, facility dependent items, consumables, utilities, waste treatment

Values: 39%, 8%, 51%, 0%, 2%, 0%

**Fig. 3** Breakdown of operating cost for PHAs production (9000 tons per annum) using the recovery strategy: **a** thermoseparation coupled with surfactant–hypochlorite digestion and **b** surfactant–hypochlorite digestion

algorithm due to its wide-covering environmental impacts and comprehensive chemical database as compared to other indicators such as CML 2 Baseline 2000 v2.03 assessment method. There are seven main chemicals which involved in the processes, including glycerol, water, oxygen, carbon dioxide, SDS, sodium hypochlorite solution, and EOPO 3900. However, it should be noted that the
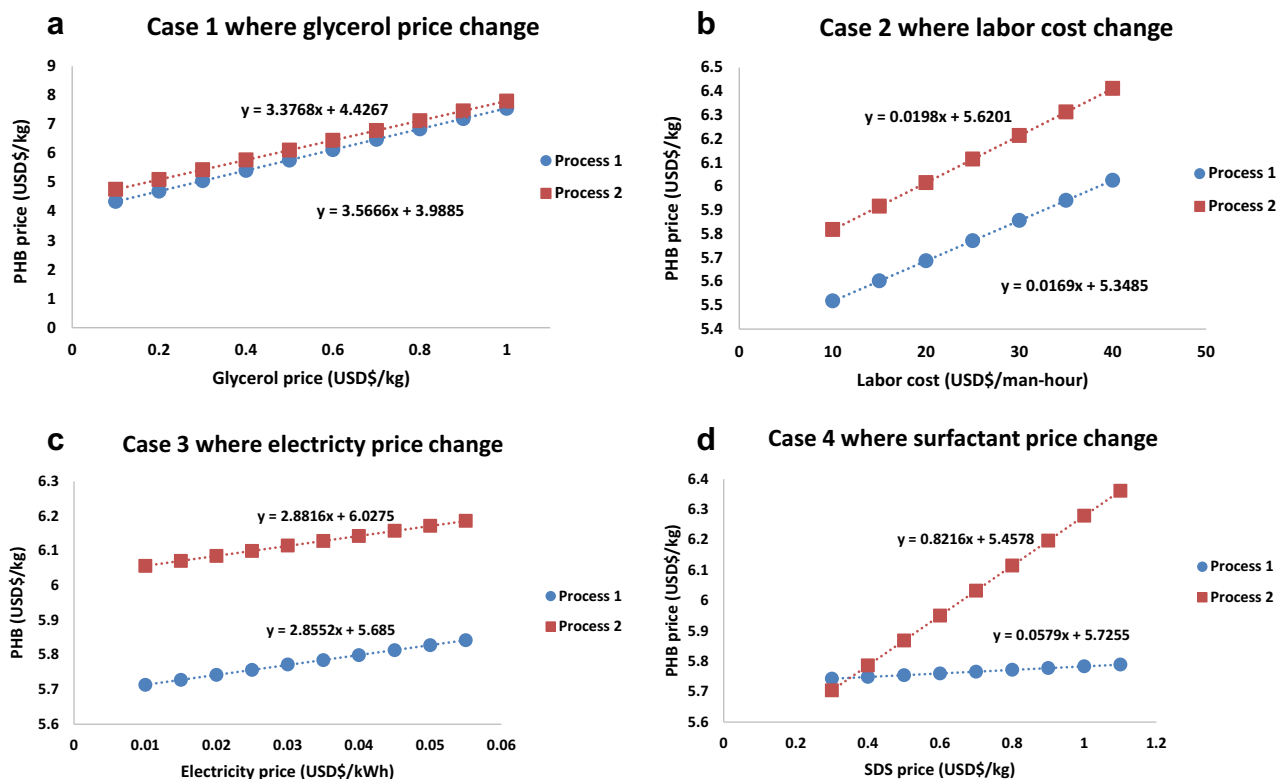
**Fig. 4** Sensitivity analysis of four different cases (*circle* process 1, *square* process 2)

environmental data on SDS and EOPO 3900 are not available. Being toxic to human (through inhalation) as well as terrestrial and aquatic life, sodium hypochlorite solution is the most hazardous chemical present in the PHA production process. Since the consumption for both processes are almost similar for most chemical except for SDS, hypochlorite solution, and EOPO 3900, it is obvious that process 2 has a worse environmental performance as it utilizes almost 10-fold hypochlorite solution compared to process 1. Therefore, it is advised that the recycling of hypochlorite solution should be done as much and as efficient as possible.

On the other hand, the chemical which brought the most negative impacts on the environment is glycerol due to the amount utilized. Glycerol dominates as much as 99.7 and 97.2% of environmental impacts for processes 1 and 2, respectively. Other than causing harm to human via inhalation and exposure as well as terrestrial and aquatic life, glycerol also causes the formation of photochemical smog as 2.4 kg of $C_2H_2$ (ethylene) equivalent, which is released into the atmosphere per kg PHA produced. Careful handling of chemical and excellent ventilation is required to avoid any accident or leakage to the surroundings. Other than that, carbon dioxide produced during fermentation is one type of greenhouse gases and contributes to global warming as 0.55 kg of $CO_2$ is released into the atmosphere

per kg PHA produced. The major process contributing to the environmental impacts in the PHA production is PHA fermentation. Henceforth, the fermenter is the piece of equipment among all the process units that needed most attentions to minimize environmental impacts of the process.

## Sensitivity analysis

Aiming to investigate the consequences of "what if?" scenarios, a sensitivity analysis was conducted by considering the main parameters that influence economic performance of PHA production by different recovery strategies. Four different cases were considered as the parameters considered contribute to significant share of operating cost and might have an influence on the result when the value of the parameters changes. The results of the four cases are displayed in Fig. 4a–d.

In case 1, the price of carbon source, glycerol, was varied from 0.1 to 1.0 USD\$/kg due to its significant contribution to operating cost as mentioned in the previous section. As predicted, the PHAs price of both processes increases linearly with similar increment (0.34–0.36 USD\$/kg PHAs increment per 0.1 USD\$/kg glycerol price increased) as the glycerol price increases. On the other hand, the labor cost was varied from 10 to 40 USD\$/man-

hour in case 2 to investigate the effect of fluctuation of labor cost on PHAs production price. Though lesser than the previous case, manipulating labor cost still has a significant influence on PHAs price for both processes. Similarly, the price of electricity was manipulated from 0.01 to 0.055 USD\$/kWh in case 3 to study the sensitivity of PHAs price toward fluctuation of electricity price. Compared to the previous 2 cases, PHAs price is much less sensitive to change in electricity price in both processes as an insignificant increment in PHAs price is observed over a range of electricity price.

Case 4 is a sensitivity analysis which investigates the impact of change in surfactant price on PHAs price in both cases. The price of SDS is manipulated from 0.3 to 1.1 USD\$/kg. For process 1, the PHAs price is insensitive to change in the SDS price due to the small amount of surfactant consumed. However, this is not the case for process 2 as the PHAs price changes drastically as the surfactant price varied. The PHAs prices drops following the fall in surfactant price, even until a point (SDS price $\sim$ 0.35 USD\$/kg) where process 2 becomes more economically favorable than process 1. It should be noted that original price of SDS is 0.8 USD\$/kg. Therefore, it is possible for process 2 to outperform process 1 in terms of economic performance (though not environmental performance) where a low cost surfactant with similar performance as SDS is present. In conclusion, the cost of carbon source and labor has significant influences on PHAs price for both processes, while the surfactant price is a crucial parameter to evaluate the economic performance of process 2.

## Conclusion

In the paper, the PHAs production process using thermoseparating ATPE as primary purification and recovery strategy was designed and the performance was evaluated. The evaluations on economic and environmental performance of two PHAs production processes using surfactant–hypochlorite digestion recovery strategy with and without thermoseparating ATPE were performed. The results show that most of O&M cost is contributed by facility-dependent cost and raw material cost, while the main contributor of raw material cost is carbon source. An integrated approach including process design, economic and environmental assessments identifies PHAs production process with thermoseparating ATPE as primary recovery step outperformed the other in terms of both economically and environmentally. PHAs production process which utilizes thermoseparating ATPE recovery strategy achieves PHA production cost of 5.77 US\$/kg with payback period of less than 4 years and ROI of 25.2% assuming selling cost of 10 US\$/kg. Insight from the sensitivity analysis has demonstrated that while costs of carbon source and labor have significant influence on PHAs price, the cost of surfactant is still the crucial parameter when comparing performance of both the processes. This proved that utilizing thermoseparating ATPE as primary recovery step not only helps to reduce chemical consumption, it also minimizes downstream equipment cost and wastewater treatment cost. It also demonstrated that thermoseparating ATPE is an economical viable and technically feasible PHAs primary purification strategy.

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