Electronic Supplementary Information (ESI)

Sustainable bio-succinic acid production: superstructure optimization, techno-economic and lifecycle assessment

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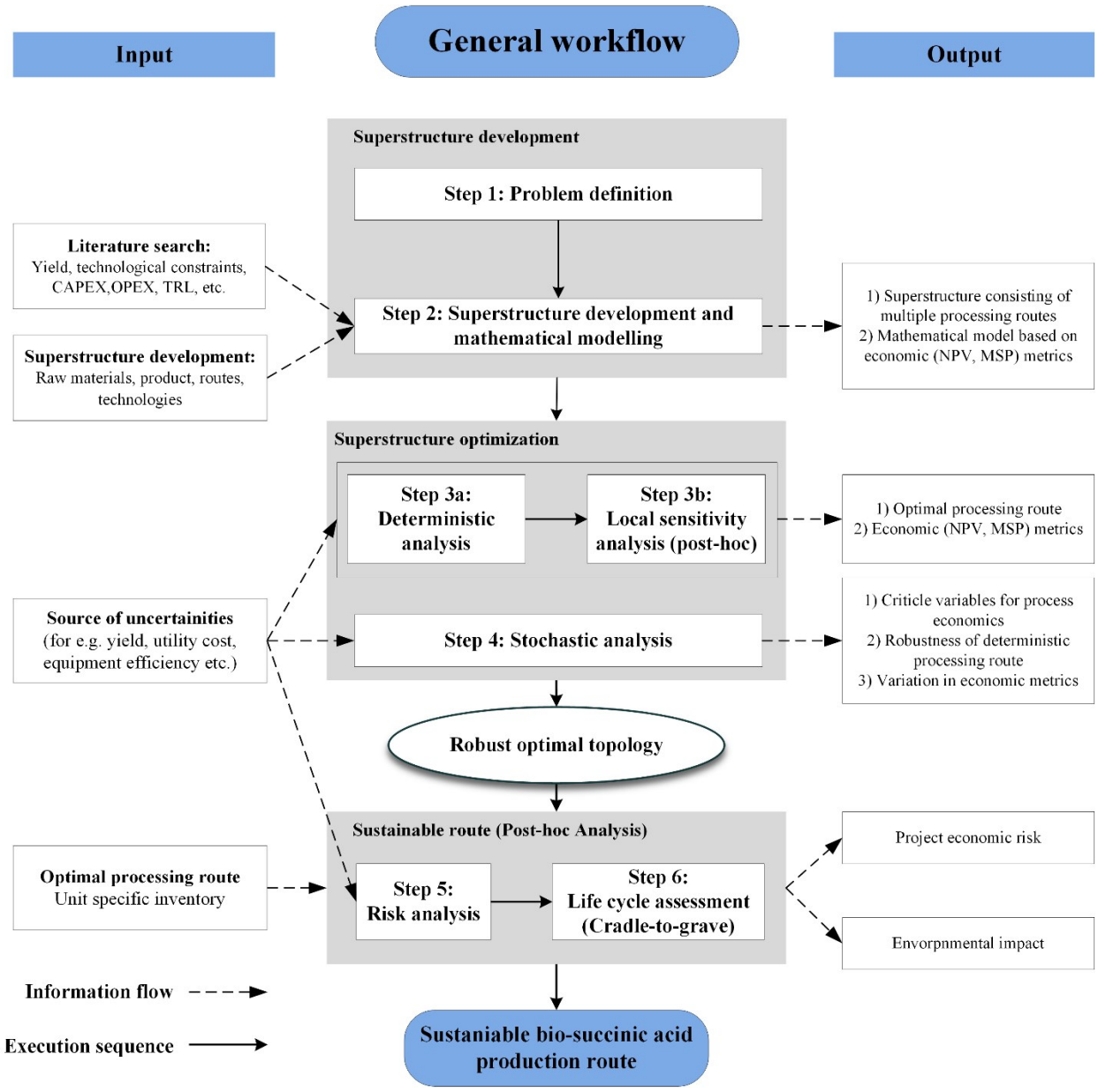


Figure S1 Framework to determine optimal processing pathway to produce bio-SA acid.

Table S1 Nomenclature of superstructure producing bio-succinic acid from corn stover, *S. japonica*, glycerol, and glucose.

Notation	Description
	Feed
(1,1)	Corn stover (CS)
(1,2)	*S. japonica* (SJ)
(1,3)	Glycerol (GLY)
(1,4)	Glucose (GLU)
	Pretreatment
(2,1)	Acid thermal hydrolysis of corn stover
(2,2)	Deacetylation + Acid thermal hydrolysis
(2,3)	Alkaline (NaOH) hydrolysis
(2,4)	Acid thermal hydrolysis of *S. japonica*
(2,5)	Hot water washing
(2,6)	Bypass
(2,7)	Bypass
	Fermentation
(3a,1)	Magnesium carbonate
(3a,2)	Magnesium hydroxide
(3a,3)	Sodium carbonate
(3a,4)	Sodium hydroxide
(3a,5)	Ammonia
	Fermentation
(3b,1)	SHF of sugars from CS using *A. succinogen*
(3b,2)	Batch fermentation of sugars from CS using *A. succinogen*
(3b,3)	SSF of sugars from CS using *A. succinogen*
(3b,4)	Dual-Phase fermentation of sugars from SJ using *E. coli*
(3b,5)	SHF of sugars from SJ using *A. succinogen*
(3b,6)	Dual-Phase fermentation of GLU using *E. coli*
(3b,7)	Batch fermentation of sugars from GLU using *A. succinogen*
(3b,8)	Fed-batch fermentation of sugars from GLY using *E. coli*
(3b,9)	Fed-batch fermentation of sugars from GLY using *A. succinogen*
	Cell mass removal
(4,1)	Microfiltration
(4,2)	Centrifuge
	Concentration pre-isolation
(5,1)	Evaporator
(5,2)	Vacuum distillation
(5,3)	Bypass
	Isolation
(6,1)	Reactive extraction
(6,2)	Electrodialysis
(6,3)	Direct crystallization

(6,4)	Ion-exchange column
(6,5)	Reactive crystallization
(6,6)	Membrane technology
	Concentration post-isoloation
(7,1)	Evaporator
(7,2)	Vacuum distillation
(7,3)	Bypass
	Impurities removal
(8,1)	Nanofiltration
(8,2)	Activated carbon
(8,3)	Bypass
	Purification
(9,1)	Solvent
(9,2)	Crystallization
	Drying
(10,1)	Drying
	Product
(11,1)	Bio-succinic acid

1. Superstructure conceptual design

This section describes all alternatives embedded in different sections of the superstructure. Red blocks in the process flow diagrams (Figures S1–S9) are used to select option *k* from stage *j* (conditional task). These are the primary decision variables for the selection of optimal topology.

1.1 Feedstocks (Biomass handling and storage)

Four different feedstocks, including corn stover, *S. japonica*, glycerol, and glucose are considered in the superstructure. The corn stover and *S. japonica* are processed and stored on-site according to NREL guidelines.[1] The glycerol and glucose are obtained from bio-diesel plants and sugar beet crops, respectively. The bio-glycerol produced at bio-diesel plant is crude and requires a pretreatment to used it for bio-succinic acid (bio-SA) production. For simplicity, in the optimization model, it is assumed that bio-glycerol pretreatment occurs at bio-diesel plants, and the cost of refined bio-glycerol is included in the feedstock price.

1.2 Pretreatment

The process flow diagrams and reaction conditions for all five pretreatment technologies are detailed in Figures S1–S2 and Table S2, respectively. The acid thermal hydrolysis follows NREL[1] design in which corn stover is first milled (Milling-101) and then mixed with hot water to reach 30 wt% solid loading in the outlet of the reactor (ATH-101). The slurry is sent to the prestreamer (M-102), where steam is injected directly to increase the temperature to 100 ^{o}C. The slurry is further heated (H-101) to 158 ^{o}C and sent to the pretreatment reactor (ATH-101), where hydrolysis occurs for 5 minutes using sulfuric acid as a catalyst (22.1 mg/g dry biomass), converting 90% of xylan to xylose. The reactor hydrolysate is discharged to a flash tank (Fl-101) that operates at atmospheric pressure. Slurry from Fl-101 is sent to a conditioning reactor (COND-101) for 30 minutes, where water is added to dilute the slurry to reach a solid loading of 20 wt%. In addition, ammonia gas is mixed into dilution water to raise the hydrolysate pH to 5. The neutralized slurry is cooled (C-101) to 48 ^{o}C and transferred to the saccharification reactor (SCR-101), where cellulase enzyme converts cellulose to glucose. The cellulase loading of 20 mg enzyme/g cellulose is used to achieve a 90% conversion to glucose. For *S. japonica*, the pretreatment via acid thermal hydrolysis (Figure S2) follows the same step as detailed for corn stover; the only difference is reaction conditions reported in Table S2.

The process design of deacetylation followed by acid thermal hydrolysis follows the method developed by Salvachúa *et al.,*[2] where corn stover is first milled (Milling-101) and mixed with water to reach 8 wt% solid loading in the hydrolysate of pretreatment reactor (DEACE-101). The slurry is heated (H-102) to 80 ^{o}C and sent to the reactor (DEACE-101) for 2 hours, where sodium hydroxide at 0.4% w/v concentration is used to remove 80%–90% acetate and 11%–22% lignin.[3] The deacetylated slurry is then washed (W-101) with water to remove alkali and sent for acid thermal hydrolysis (ATH-102) and enzymatic hydrolysis (SCR-102), which operates the same way as described previously.

The design of alkaline hydrolysis follows the method developed by NREL.[4] In this method, water is added to milled (Milling-101) corn stover to reach 10 wt% solid loading in the outlet of the pretreatment reactor (ALTH-101). The slurry is then heated (H-104) to 120 ^{o}C and sent to the reactor (ALTH-101), where hydrolysis occurs for 10 minutes using NaOH as a catalyst (55 mg/g dry biomass). The slurry from the reactor is dewatered to 25%–35% dry matter using a belt filter press (BFP-103) and then fed to the saccharification reactor (SCR-103) that operates the same way as described previously.

S. japonica can be pretreated by a hot water washing method proposed by Alvarado *et al.*[5] Here, milled (Milling-201) *S. japonica* is mixed with water to achieve 25 wt% solids in the outlet of the reactor (HWW-201). The reactor operates at 121 ^{o}C for 20

minutes. The hydrolysate from the pretreatment reactor is cooled (C-204) and sent to a saccharification reactor (SCR-202) that operates at 25 wt% solid loading and a temperature of 50 °C with a cellulase loading of 20 mg/g cellulose.

Glucose and glycerol are pure at biorefinery gates; therefore, they will bypass this processing interval as shown in Figure 2 of the manuscript.

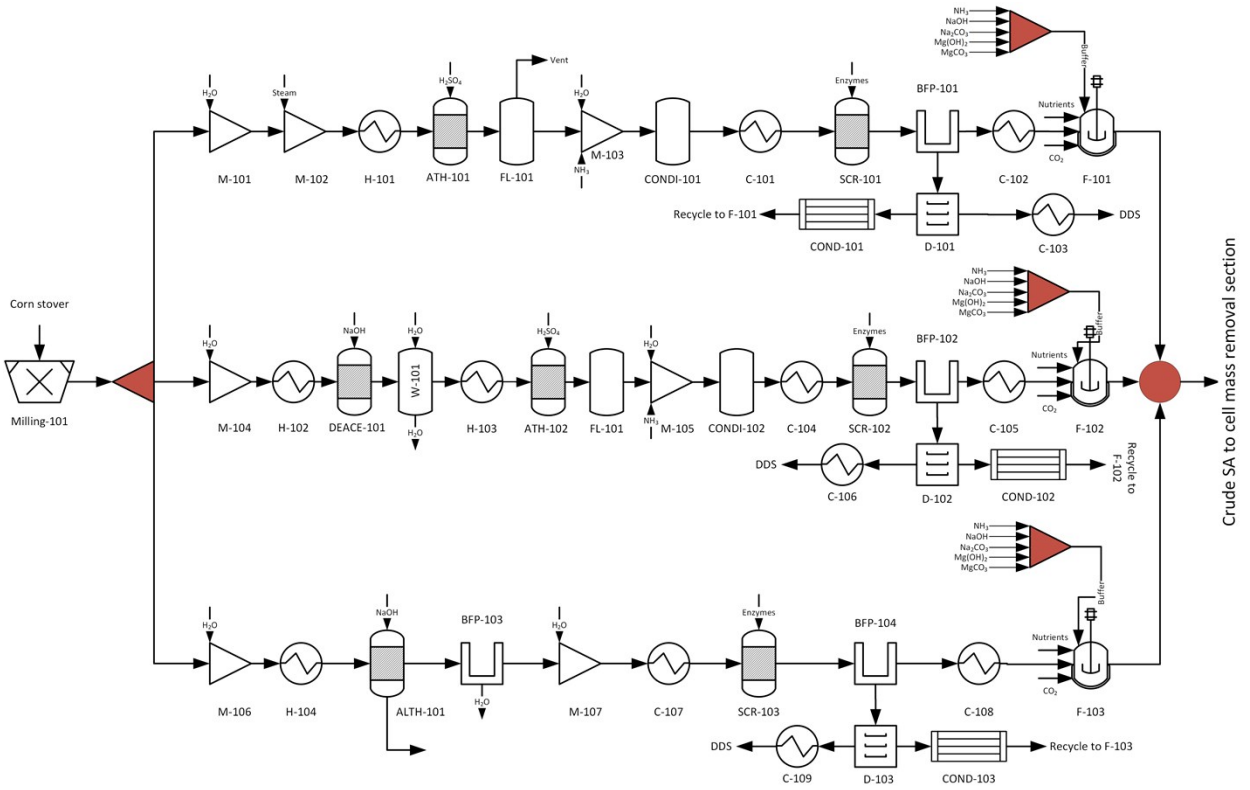
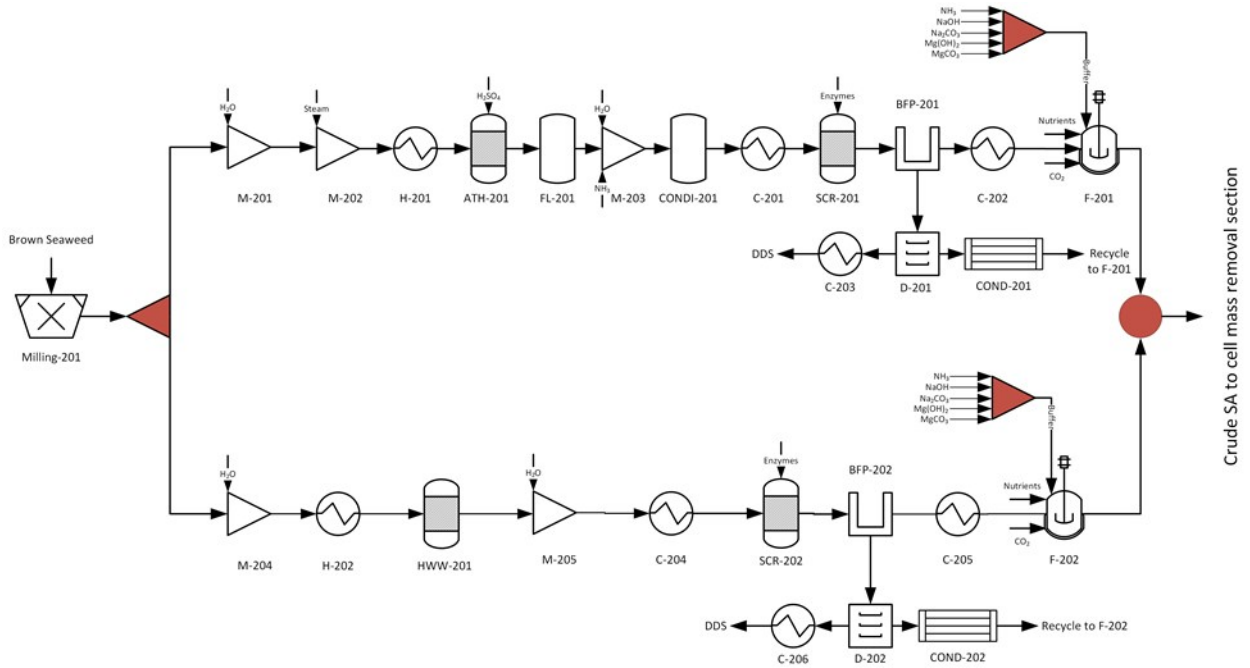


Figure S2 Process flow diagram for corn stover pretreatment, enzymatic hydrolysis, and fermentation.

Figure S3 Process flow diagram for *S. japonica* pretreatment, enzymatic hydrolysis, and fermentation.

Table S2 Summary of operating conditions and yields for various pretreatment technologies.

Processing conditions	**P1**[1,6]	**P2**[2,7]	**P3**[4]	**P4**[8,9]	**P5**[5]



	ATH	EH	DEA	ATH	EH	ALTH	EH	ATH	EH	HWW	EH
	Corn Stover							*S. Japonica*			
Solid loading (wt%)	30	20	8	12	12	10	20	30	20	25	25
Residence time (mins)	5	84	120	10	84	10	120 (5 days)	15	84	20	48
Temperature ©	158	50	80	160	50	120	50	121	50	121	50
Catalysis loading	22.1[a]	20[b]	0.4[c]	8[a]	20[b]	55[a]	20[b]	18[a]	20[b]	0	20[b]
Glucose yield (%)	-	91	6.2	-	91	-	90	-	91	-	78.2
Xylose yield (%)	90	-	-	81	-	70	-	-	-	-	-
Mannitol yield (%)	-	-	-	-	-	-	-	95	-	78.2	-

P1-P5 = pretreatment; ATH = Acid thermal hydrolysis; EH = Enzymatic hydrolysis; DEA = Deacetylation; ALTH = Alkaline thermal hydrolysis
P1 and P4: acid thermal hydrolysis followed by enzymatic hydrolysis for corn stover and *S.japonica*, respectively
P2: Deacetylation followed by acid thermal- and enzymatic - hydrolysis
P3: Alkaline hydrolysis followed by enzymatic hydrolysis
P5: Hot water washing followed by enzymatic hydrolysis
[a]mg/g dry biomass; [b]mg/g cellulose; [c]wt%

1.3 Fermentation

The hydrolysate from any saccharification reactor (SCR-101–103, SCR-201, or SCR-202) is filtered using a belt filter press (BFP-101–103, BFP-201, or BFP-202) to remove unreacted feedstock that is rich in proteins and can be sold as animal feed at a local market. The separated solids from the belt filter press are wet and sent to a dryer (D-101–103, D-201, or D-202) that removes the moisture, which is recycled to the process (F-101–103, F-201, or F-202). The dried solids are cooled (C-101–103, C-201, or C-202) and stored. The filtrate from the belt filter press or feed from the bypass of the pretreatment section (in case of glucose and glycerol) is pumped to the fermenter to convert sugars into bio-SA. Here, nine fermentation technologies are included: three for corn stover,[2,6,10] and two for *S. japonica*,[5,8] glucose,[5,11] and glycerol[12,13]. The process flow diagrams pertaining to fermentations are detailed in Figures S1–S4. The inlet feed of the fermenter is first cooled (C-102, C-105, C-108, C-202, or C-205) or heated (H-301, H-302, H-401, or H-402) to fermentation temperature and then fed to the fermenter (F-101–103, F-201, F-202, F-301, F-302, F-401, or F-402) where host microorganisms convert sugar monomers to bio-SA in the presence of carbon dioxide and nutrients. A summary of operating conditions, yields, host microorganisms, productivity, broth concentration, and fermenter type are detailed in Table S3. As indicated before, a buffering agent's choice is important regarding process economics; therefore, five buffering agents, including magnesium hydroxide, magnesium carbonate, sodium hydroxide, sodium carbonate, and ammonia are included in the superstructure.

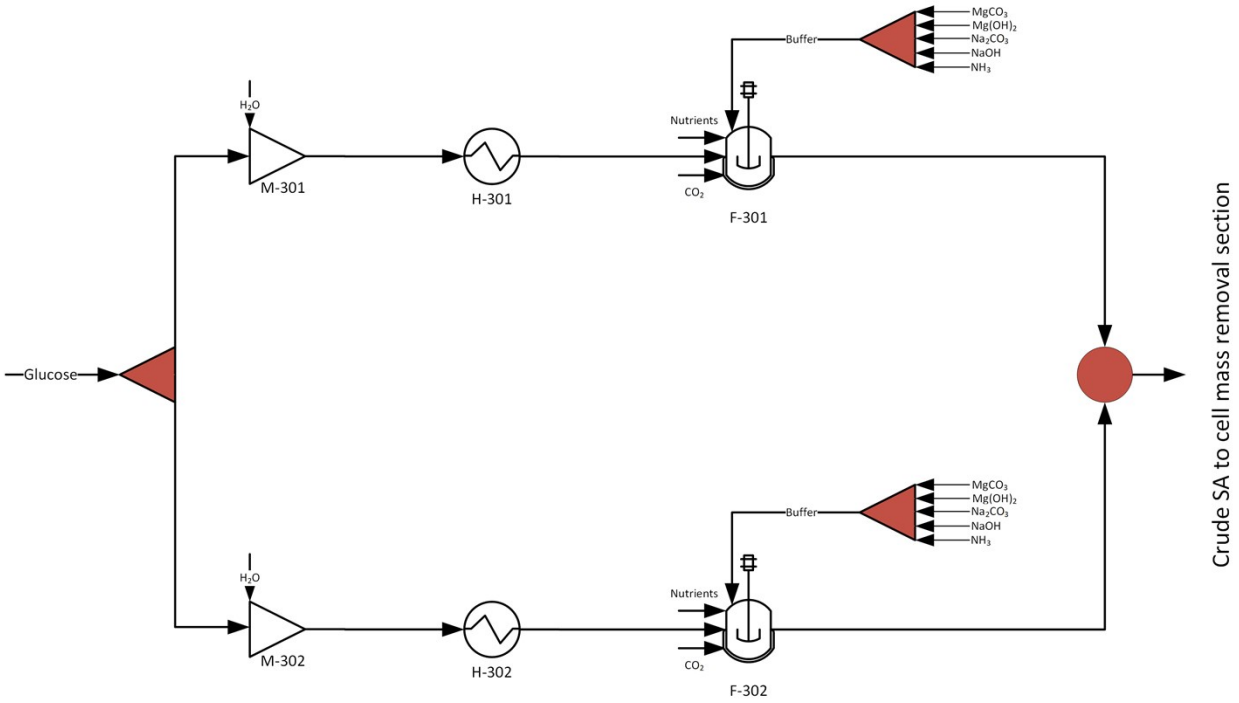


Figure S4 Process flow diagram for glucose fermentation.

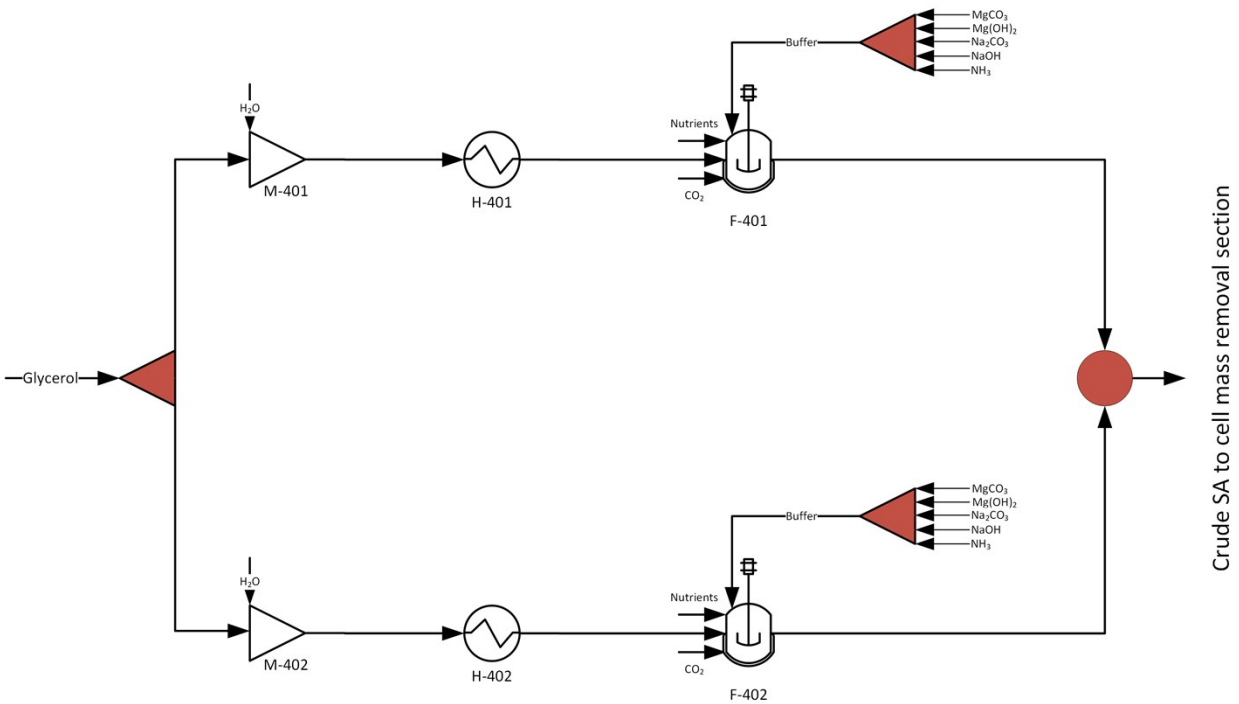


Figure S5 Process flow diagram for glycerol fermentation.

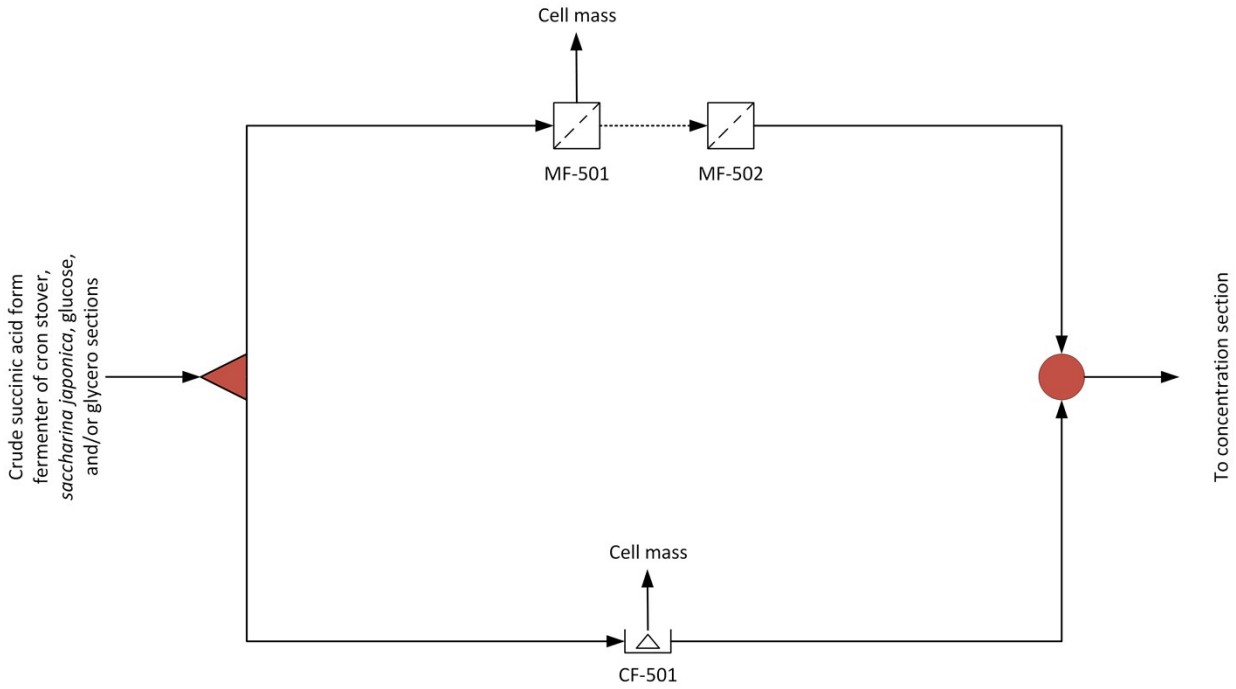
Table S3 Summary of operating conditions and yields for various fermentation technologies for corn stover (F1-F3), *S. japonica* (F4-F5), glucose (F6-F7), and glycerol (F8-F9).

No.	Strain name	Fermentation Type	Carbon Source	Titer (g/L)	Yield (g/g)	Productivity (g/L/h)
F1[6]	*A. succinogen*	SHF	Glucose and xylose	56.40	0.73	1.08
F2[2]	*A. succinogen*	Batch	Glucose and xylose	42.80	0.74	0.30
F3[10]	*A. succinogen*	SSF	Glucose and xylose	47.40	0.72	0.99
F4[8]	*E. coli*	Dual-Phase	Glucose and Mannitol	17.40	0.73	0.24
F5[5]	*A. succinogen*	SHF	Glucose and Mannitol	33.78	0.63	0.70
F6[14]	*E. coli*	Dual-Phase	Glucose	99.20	1.10	1.31
F7[12]	*A. succinogen*	Batch	Glucose	105.80	0.82	1.36
F8[13]	*E. coli*	Fed-batch	Glycerol	66.78	1.24	0.93
F9[15]	*A. succinogen*	Fed-batch	Glycerol	49.62	0.87	0.64

F1–F9: Fermentation1– Fermentation 9; SHF: Separate hydrolysis and fermentation; SSF: Simultaneous hydrolysis and fermentation

1.4 Cell mass removal

The process flow diagram pertaining to the microfiltration and the centrifuge is given in Figure S5. In both alternatives, it is assumed that 100% of the biomass is removed, and 2% SA is lost from the hydrolysate. In microfiltration, two membranes (MF-501 and MF-502) operate in parallel to ensure a continuous process during the membrane cleaning. The membrane regeneration is required after 84 hours of operation, and it takes 2 hours to achieve proper cleaning.[16] The chemicals used for cleaning included sodium hydroxide (1 wt%) and hydrochloric acid (1 wt%).[17] The amortization of the membrane is 3 years, representing 30% of membranes' total cost.[18] Compared to microfiltration, a centrifuge (CF-501) is a robust option that requires less maintenance. The maximum throughput of the centrifuge is 96 m^3h^{-1} with a power consumption of 93 kW.[19] To prevent maintenance downtime,



one extra centrifuge is assigned in addition to the minimum number of units required to perform the separation.

Figure S6 Process flow diagram for dead cell mass removal via microfiltration or centrifugation.

1.5 Concentration pre-isolation

The process flow diagram of this section is given in Figure S6 in which permeate (SA solution) from the microfiltration or the centrifuge may reach a concentration of 200 g SA/L in the evaporation[17] (E-601) or vacuum distillation[20] (DIST-601) processes. Alternatively, this processing interval can also be bypassed.

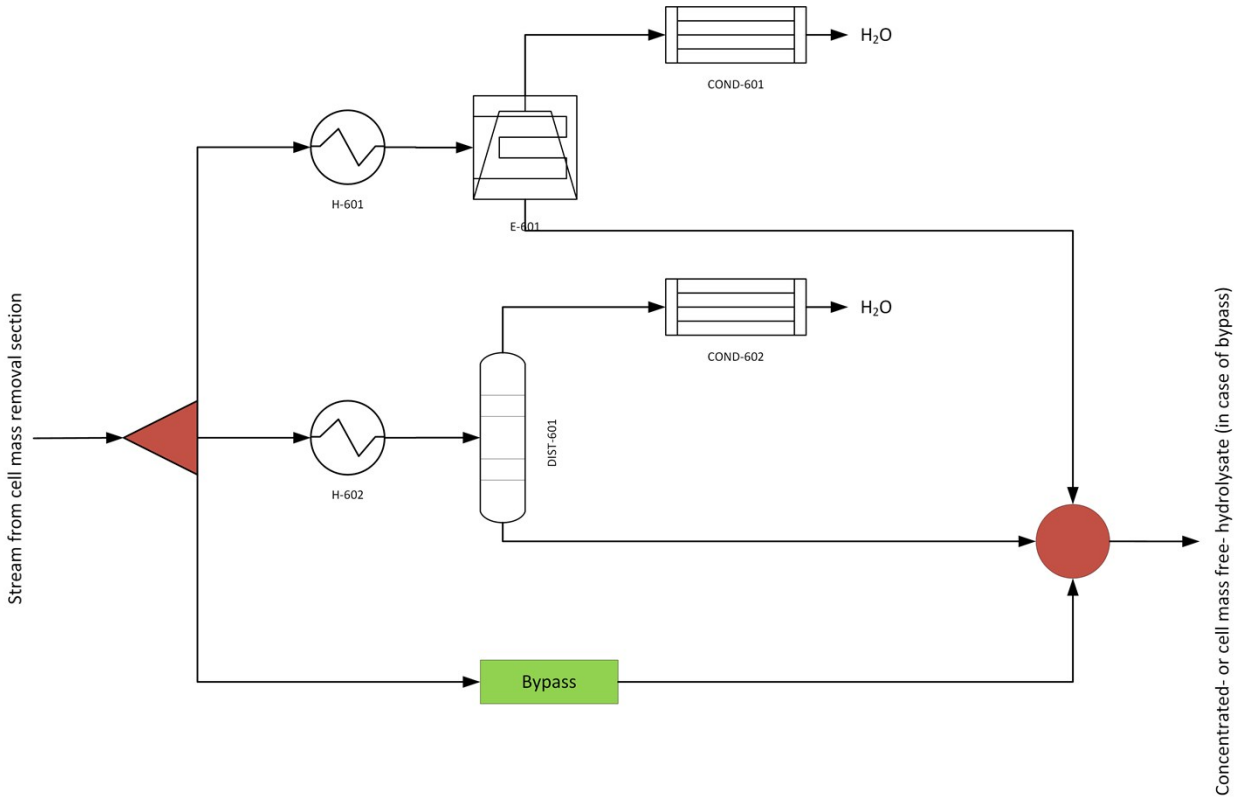


Figure S7 Process flow diagram for bio-SA broth concentration before or after the isolation step.

1.6 Isolation

Isolation is the separation technology that separates SA from the clarified broth. Technologies embedded for this task in the superstructure are shown in Figure S7 and include (1) electrodialysis,[21] (2) direct crystallization,[20] (3) reactive extraction,[22,23] (4) ion-exchange column,[24] (5) reactive crystallization,[25] and (6) membrane technology.[17]

In electrodialysis,[21] bio-SA is separated from the broth using two desalting (DSED-701 and DSED-702) units and one water splitting (WSED-701) unit. Note that this technology cannot handle divalent ions, and therefore, a restriction is applied to the selection of buffering agents that produce divalent ions.[26] In addition, this technology does not require acidified broth containing free acid, and therefore, acidification can be bypassed (Bypass-701), and broth can be directly sent to the electrodialysis units. In this method, succinate salt is separated from the broth (diluting stream) in a concentrating stream using two desalting electrodialysis units working in series. The driving force of the electrodialysis is the electrical potential difference across an alternating series of positively and negatively charged polymer membranes. The concentrated stream containing the enriched succinate salt at 22.8 wt% concentration is sent to water splitting electrodialysis that converts or splits succinate salt to bio-SA and alkali. The recovered

alkali is recycled to the fermentation unit where it is used for pH control. The electricity consumption for both desalting electrodialysis units and the water splitting electrodialysis unit is assumed to be 3.5 and 2.5 kWh/succinate salt, respectively, based on the work of Garde.[27] This technology recovers 88% of the bio-SA at 99.9 wt% purity from the fermentation broth.[21]

In direct crystallization, the clarified fermentation broth is first acidified (V-701) with 10% molar excess sulfuric acid, producing free bio-SA and salt.[20] The salt is separated via filtration (F-701), and permeate is fed to the evaporator to concentrate it to 200 g/L. Afterward, the concentrated stream is slowly cooled (C-701) to 4 ^{o}C and sent to a crystallizer (CRYS-701) to produce a suspension of bio-SA crystals. This method recovers 57% of the bio-SA at 90 wt% purity from the fermentation broth.[23]

Reactive solvent extraction consists of three extraction columns (EC-701–703) that work in series and a back extraction column (BEC-701).[22] In the former, bio-SA is extracted from the aqueous phase (acidified broth) into an organic phase consisting of 1-octanol (87 wt%) and trioctylamine (13 wt%). Equal volumetric flows of the organic and aqueous phase are used to achieve an efficiency of 86% per column.[22] The organic phase from the extraction columns are mixed and fed to the back extraction column that re-extracts the bio-SA from the organic phase into the aqueous phase using a solvent consisting of 25% trimethylamine and 75% water. As suggested by Kurzrock *et al.*[22] 9.3 mol of trimethylamine per mol bio-SA is used to achieve a separation efficiency of 1. A solvent loss of 0.21 wt% of 1-octanol and 0.1% of trimethylamine is considered for the extraction- and back extraction column which is constantly fed via the makeup stream.[26] This technology recovers 99.7% of the bio-SA at 99.9 wt% purity from the fermentation broth.[26]

In the ion-exchange column,[24] the clarified broth containing succinate salt is first acidified (V-701) with sulfuric acid, producing bio-SA and salt. The acidified broth is then purified through two resin bed trains consisting of a strong cation exchanger (CIEC-701) to separate all positively charged ions and a weak anion exchanger (AIEC-701) to separate all negatively charged ions; Ionic impurities (e.g., magnesium, hydroxide) are separated into raffinate and free acid broth into the extract. For each ion-exchanger, two units (CIEC-702 and AIEC-702) are installed in parallel to allow for the regeneration of the resins, which is expected to be required every 48 hours.[24] The strong cation and weak anion exchanger contain Dowex 50 WX8 resin and Rohm and Hass Amberlite IRA-96 resin, as suggested by Datta *et al.*[24] The cation and anion exchanger are regenerated using 8 wt% hydrogen chloride and 8 wt% sodium hydroxide, respectively.[24] The waste material produced during the regeneration cycles is routed to the wastewater treatment plant. 71% of the bio-SA can be recovered from the fermentation broth at 99.9 wt% purity using this technology.

The reactive crystallization technology is developed based on the work of Yedur *et al.*[25] The method involves the formation of diammonium succinate using ammonia as a buffering agent in the fermenter. The clarified broth is acidified with ammonium bisulfate in a reactive crystallizer (CRYS-702) to produce free acid and ammonium sulfate as a byproduct. The ammonium sulfate crystals are removed by filtration (F-702), and the free acid broth is forwarded for further purification. The ammonium sulfate crystals may contain residual SA, which can be removed by adding methanol in the crystallizer (CRYS-703) that produces pure ammonium sulfate crystals while impurities will remain dissolved in methanol. The pure crystals are filtered (F-703) and thermally cracked (TC-701) at a temperature of 300 ^{o}C to produce ammonium bisulfate and ammonia gas, which are recycled to the reactive crystallizer (CRYS-702) and the fermenter, respectively. The methanol containing impurities is fed to the distillation column (DIST-701) that separates methanol in the distillate, which is recycled to the process (M-701), while impurities are separated in the bottom and sent to a wastewater treatment plant. 93.3% of the bio-SA can be recovered from the fermentation broth at 99.9 wt% purity using this technology.

The membrane technology is a combination of microfiltration and nanofiltration membranes. The design of the membrane technology system is based on the results of a pilot-scale study reported by Thuy et al.[17] The microfiltration membrane removes the dead cells and clarified broth, which is then followed by an acidification step to produce free acid and salt. The broth is then filtered and pumped through a spiral-wound membrane that removes color impurities and macromolecules. The acidified broth free from dead cells and color impurities can be further purified using crystallization or solvent purification to recover 86.53% of the bio-SA at 99.2 wt% purity.[17] Note that in the superstructure, microfiltration is embedded in the cell mass removal process interval (see Section 1.4), and therefore, the first separation task will be performed here. The second task, in which nanofiltration membranes are used to remove color and other impurities, occurs in the color removal processing interval (see Section 1.8).

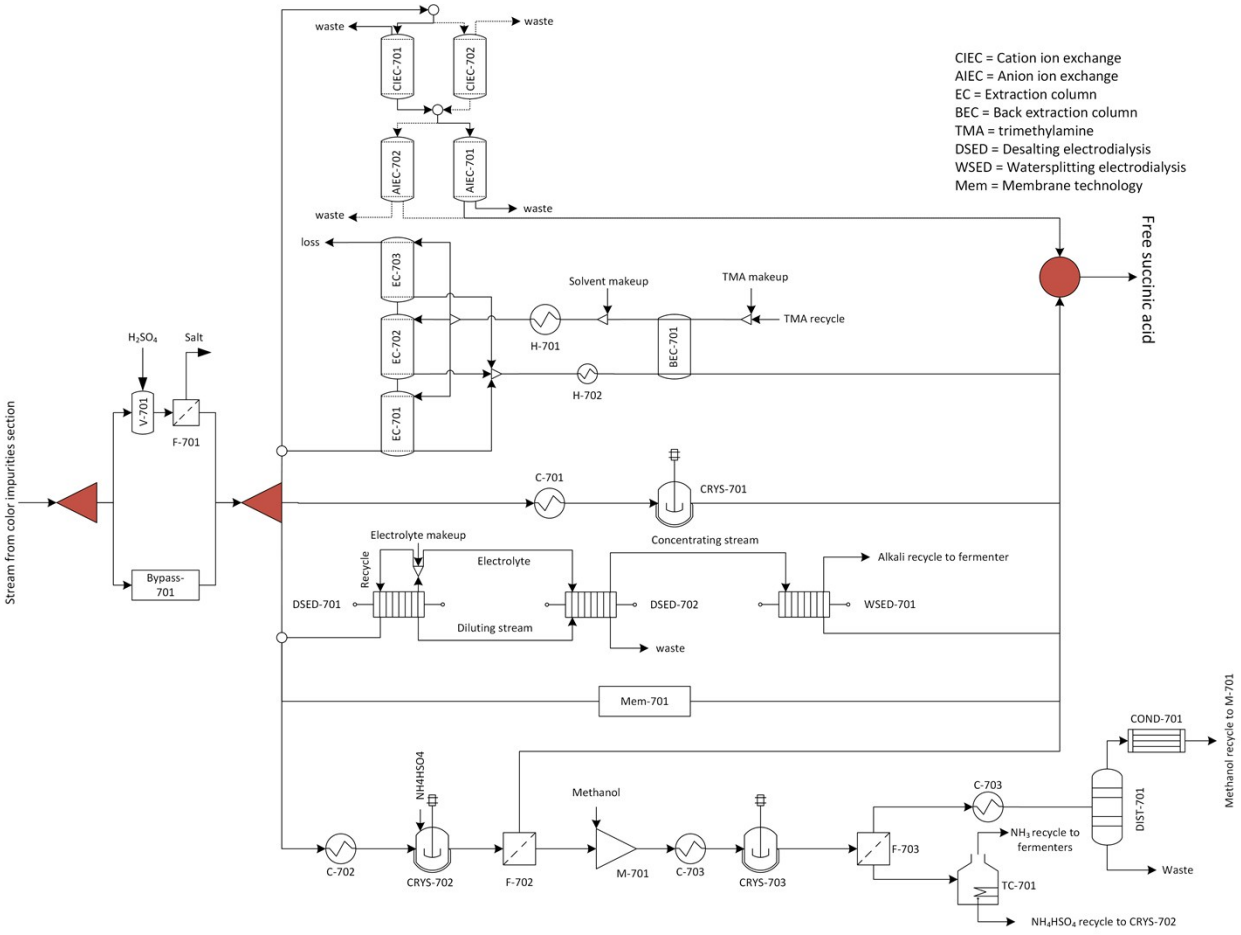


Figure S8 Process flow diagram for bio-SA isolation from succinate (salt) to free acid.

1.7 Concentration post-isolation

The objective and working principle of all alternatives embedded in this processing interval are similar to that explained in the interval concentration pre-isolation (Section 1.5). The major decision variable for these two sections is to determine where the concentration of the broth should be performed, *i.e.*, before or after the isolation section. If the concentration step is implemented before the isolation step then it may remove a large quantity of water to achieve SA concentration of up to 200 g/L

maximum that may decrease the capital and operating cost of any further downstream technology. However, evaporating such a large quantity of water also consumes a considerable amount of steam, which may increase the operating cost more than the saving expected to be achieved in the downstream process. Therefore, in the superstructure, concentration before or after the isolation step is embedded as a decision variable related to the topology of the biorefinery. Note that both concentration steps *i.e.*, before or after the isolation can also be avoided in the superstructure if other processing pathways are found more promising than these two. Regarding operation, this process interval works similarly to that described in Section 1.5.

1.8 Color removal

The process flow diagram of this section is shown in Figure S8, which aims to separate pigments and proteins from the succinate broth using a nanofiltration system or an adsorption column. In the former approach, feed is pumped (P-801) at 1.25 MPa through the spiral-wound membrane with a cutoff of 300 Da that operates in crossflow mode. In this work, two membrane systems work in parallel to ensure continuous processing during the cleaning of one membrane system, which is required every 84 hours. In the adsorption column, the feed is mixed with 1.8-2.5% (w/v) activated carbon in a column (V-801) with a holdup time of 1 hr to remove pigments and proteins. About 1.8-3.8 wt% of the SA could be lost here, depending on the concentration of activated carbon used. Alternatively, if the aforementioned technologies are not optimal, the color removal stage can also bypass.

1.9 Purification

As shown in Figure S9, two technologies are considered in the superstructure to get pure crystals: (1) crystallization and (2) solvent-based purification. In the former, concentrated broth free from cells and color impurities is slowly cooled (C-901) to saturation temperature to get a suspension of crystals (CRYS-907), which is then filtered (F-901). In the solvent-based purification technology (V-901), methanol is used to dissolve SA and leave impurities behind. The impurities are separated using filtration (F-902) and are then discarded.

1.10 Drying

Drying of crystals is the last step in the superstructure that removes moisture from the crystals. In Figure S9, the wet crystals obtained from (F-901) can be sent to the dryer (D-901), which are then cooled (C-902) and stored, whereas the vent of the dryer is condensed (COND-901) and sent to the wastewater treatment unit. In the case of solved-based purification technology, the methanol solution containing bio-SA obtained from the filter (F-902) is sent to a distillation column (Dist-901), where methanol is recovered in the distillate and recycled to the vessel (V-901). Pure bio-SA crystals are obtained in the bottom of the distillation column, which are then cooled (C-903) and stored.

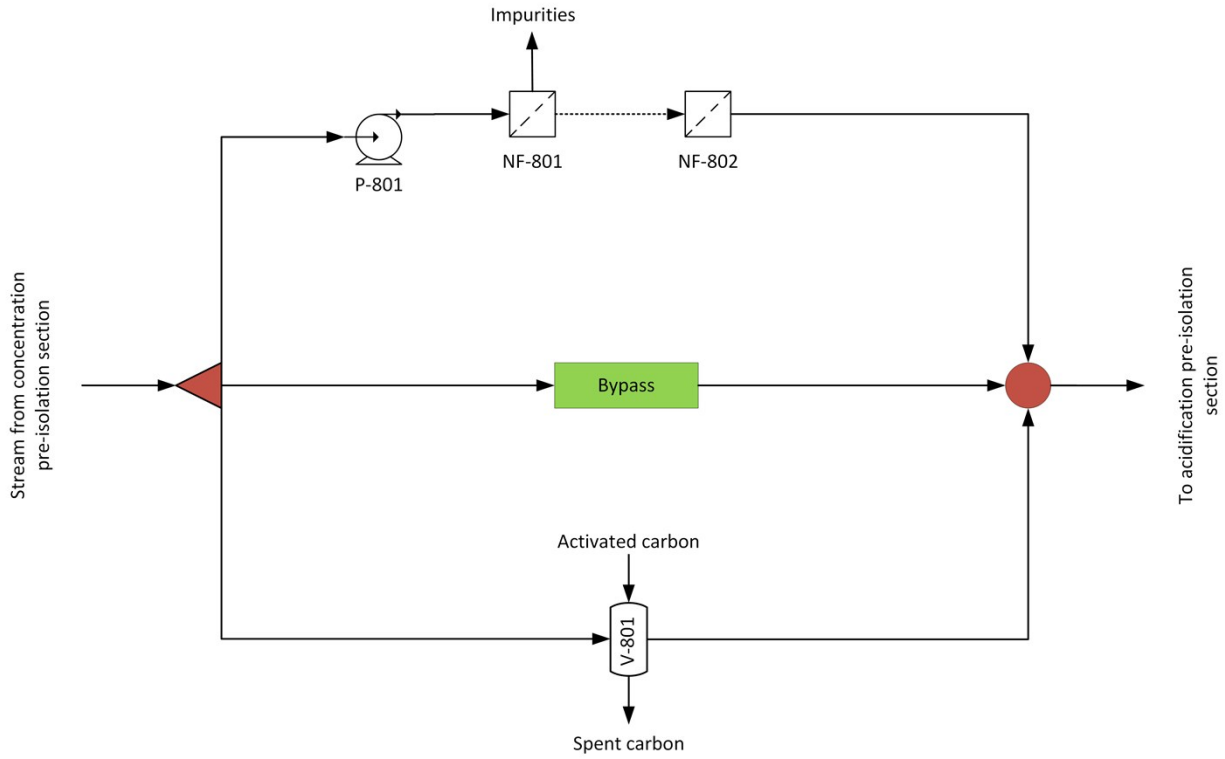


Figure S9 Process flow diagram for color impurities removal from broth containing bio-SA.

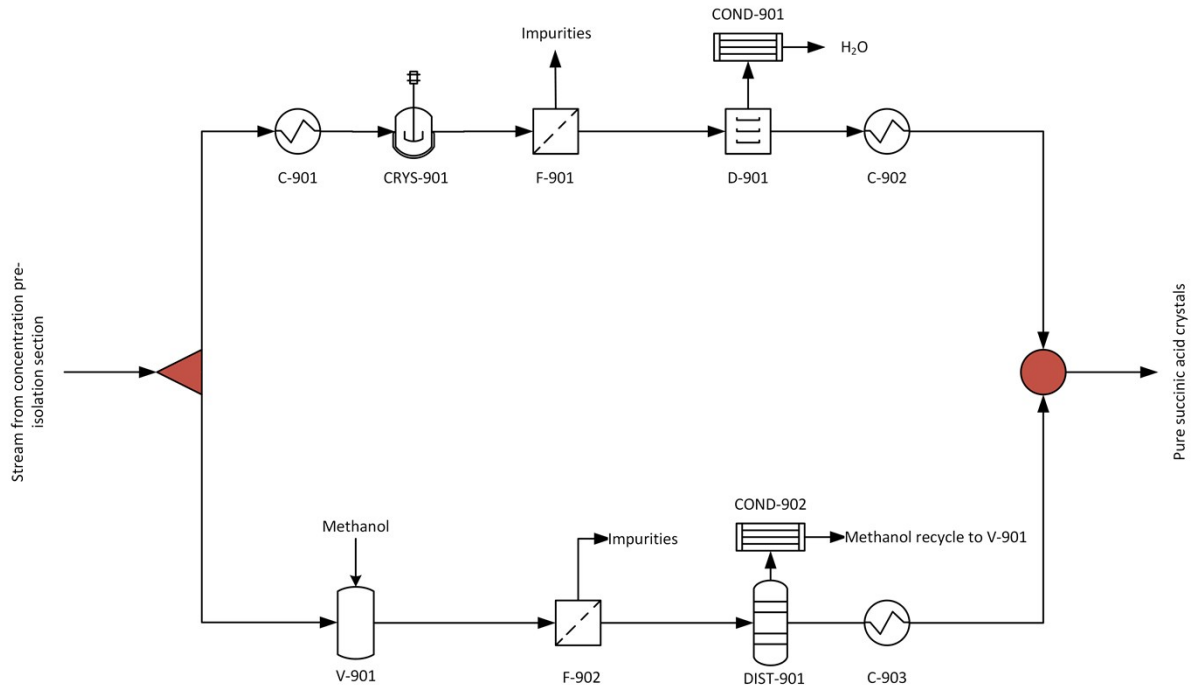


Figure S10 Process flow diagram for bio-SA purification and drying.

2. Mathematical model

The mathematical model of the superstructure is formulated as a mixed-integer linear-programming (MILP) model by considering the mass balance, energy balance, capital cost, and operating cost constraints.

Indices/sets

The indices are used throughout the mathematical model are listed below.

b : Feedstock index

k : Process stream index

i : Component index

j : Stage index

g : Processing alternatives in stage *j*

l : Utility index

The set of all elements, *b*, includes corn stover, glucose, glycerol, and *S. japonica*.

$b \in B = \{corn\ stover, glucose, glycerol, S.\ japonica\}$

The set of all elements, *k*, includes 1 to 235.

$k \in K = \{1 \times 235\}$

Note that notation *k1* and *k2* represent inlet and outlet stream at any stage *j* in the bio-SA process.

The set of all elements, *i*, is listed in Table S4 and defined formally in below Equation.

$i \in I = \{Complete\ set\ of\ species\ listed\ in\ Table\ S4\}$

Table S4 Species present in the bio-SA synthesis problem.

Species Name	Species Index	Species Name	Species Index	Species Name	Species Index
Water	H2O	Hydroxymethylfurfural	HMF	Carbon dioxide	CO2
Extractives	EXT	Sodium hydroxide	NAOH	Oxygen	O2
Sucrose	SUC	Magnesium carbonate	MGCO3	Anthraquinone	ANQU
Cellulose	CELL	Magnesium hydroxide	MGOH2	Hydrogen	H2
Xylan	XYL	Magnesium succinate	MGSA	Acetic	AAC
Arabinan	ARA	Sodium carbonate	NA2CO3	Lignin	LIG1
Mannitol	MAN	Sodium succinate	NASA	Furfural	FUR
Galactan	GAL	Diammonium succinate	DASA	Mannose	MANO
Lignin	LIG	Diammonium sulfate	DASUL	Galactose	GALO
Acetate	ACE	Ammonium acetate	AMA	Arabinose	ARAO
Ash	ASH	Ammonium sulfate	AMS	Ammonia	NH3
Protein	PRO	Ammonium bisulfate	AMBS	Malic acid	MAL
Laminarian	LAM	Sodium sulfate	NASUL	Enzymes	ENZ
Alginate	ALG	Magnesium sulfate	MGSUL	Succinic acid	SA
Mannitol	MANN	Tri-octylamine	TRI-OCTY	Lactic acid	LA
Fucoidan	FUC	Trimethylamine	TMA	Formic acid	FA
Lipids	LIP	Phosphoric acid	H3PO4	Dead cells	DCW
Glucose	GLU	Activated carbon	AC	Ethanol	ETH
Glycerol	GLY	Sodium acetate	SACE	Octanol	OCTANOL
Acid	H2SO4	Propanoic acid	PROP	Methanol	METHA
Xylose	XYLO	Pyruvic acid	PYR	Tar	TAR

The set of all elements, *j*, is listed in Table S5 and defined formally in below Equation.

$j \in J = \{Complete\ set\ of\ process\ units\ listed\ in\ Table\ S5\}$

Table S5 Process units present in the bio-SA synthesis problem.

Stage index	Stage name	Stage index	Stage name
FSH	Feedstock storage and handling	EVAP	Evaporation
MXR	Mixer	DIST	Distillation
SPL1	Topology splitter	BYPASS	Bypass
SPL2	Fractional splitter	AC	Activated carbon
HEH	Heater	ACID	Acidification
HEC	Cooler	RE	Reactive extraction
PRER	Pretreatment reactor	BK	Back extraction
COND	Conditioning vessel	ED	Electrodialysis
SCR	Saccharification reactor	WSED	Water splitting electrodialysis
FERM	Fermentation	CRYS	Crystallization
DEACE	Deacetylation	FIL	Filtration
SEPA	Separator	IEC	Ion-exchange column
WASH	Washer	RC	Reactive crystallization
MFIL	Microfiltration	MT	Membrane technology
BFP	Belt filter press	TC	Thermal cracker
CF	Centrifuge	SOL	Solvent purification
NFIL	Nanofiltration	DRY	Dryer

2.1 Mass balance constraints

Where applicable, the unit of mass flow rate is in kg per second in subsequent equations.

The component mass flow rate of feedstock *b* in the stream *k* can be modeled as follows:

$$F_{b,i}^{k} = F_{b}^{k} \times x_{b,i}^{k}, \forall b \in B, \forall i \in I \forall, k \in K, \quad (1)$$

where $F_{b,i}^{k}$ is the mass flow rate of component *i* in the stream *k*, F_{b}^{k} is the overall mass flow rate in the stream *k*, and $x_{b,i}^{k}$ is the feedstock composition of component *i* in the stream *k*.

The overall mass balance of feedstock *b* in the stream *k* is given by:

$$F_{b}^{k} = \sum_{i} f_{b,i}^{k}, \forall b \in B, \forall k \in K \quad (2)$$

.

The logical constraint to select feedstock *b* is modeled as:

$$F_{b}^{k} \leq \Phi \times y1_{b}^{k}, \forall b \in B, \forall k \in K, \quad (3)$$

where Φ represents upper bound for mass flow rate according to the big formulation method and $y1_{b}^{k}$ represent binary variables to select feedstock *b*. These binary variables must select one optimal feedstock.

The logical constraint to select multiple feedstocks is modeled as:

$$\sum_{b} y1_{b}^{k} \leq 4, \forall b \in B, \forall k \in K \quad . \tag{4}$$

Splitters are used for optimizing the topology of biorefinery by selecting option *g* from stage *j*. The constraints pertaining to the splitters are given by:

$$\sum_{k1} F_{b,i,j}^{k1} - \sum_{k2} F_{b,i,j}^{k2} = 0, \forall b \in B, \forall i \in I, \forall j \in J \quad , \tag{5}$$

and

$$F_{b,j}^{k2} \leq \Phi \times y2_{b,j}^{k2}, \forall b \in B, \forall k2 \in K, \forall j \in J \quad , \tag{6}$$

where $F_{b,i,j}^{k1}$ and $F_{b,i,j}^{k2}$ are the mass flow rate of component *i* in the inlet (*k1*) and the outlet (*k2)* stream of stage *j* when utilizing feedstock *b,* and $y2_{b,j}^{k2}$ are the binary variables for the selection of option *g* from stage *j*.

The constraint that enforces the selection of only one technology is given by:

$$\sum_{b} y2_{b,j}^{k} \leq 1, \forall k \in K, \forall j \in J \quad . \tag{7}$$

Disjunctions are used for implementing the concentration step either before or after the isolation step as follow:

$$\sum_{k2=1}^{2} y2_{b,j1}^{k2} \leq 1, \forall b \in B, \forall j1 \in J \quad , \tag{8}$$

where $y2_{b,j1}^{k2}$ are the binary variables pertaining to the outlet stream *k2 of* concertation stage *j1* in the superstructure.

To find a realistic processing pathway from the superstructure, logical constraints are used to ensure a feasible match of various processing stages. For instance, electrodialysis cannot deal with the divalent ions such as Mg^{+2} and Ca^{+2}, therefore, in fermentation the feasible match should be monovalent pH control reagent *i.e.*, sodium hydroxide that will generate monovalent ion such as Na^{+1}. Likewise, the acidification of broth before electrodialysis would be an infeasible match. These logical conditions are modeled as:

$$\sum_{k2} y2_{b,j1}^{k2} - y2_{b,j2}^{k2} \leq 0, \forall b \in B, \forall j1 and\ j2 \in J \quad , \tag{9}$$

where $y2_{b,j1}^{k2}$ and $y2_{b,j2}^{k2}$ are the binary variables corresponding to the outlet stream *k2* of stage *j1* and *j2*.

The mass balance equation for reactors and purification technologies such as the pretreatment, deacetylation, enzymatic hydrolysis, fermenter, conditioning vessel, acidification vessel, water splitting electrodialysis, reactive crystallization, and thermal cracker, where the reactant *r* is converted to the product *p* is given by

$$F_{b,p,j}^{k2} = F_{b,r,j}^{k1} \times \Phi_{b,p,r,j}^{k2} + F_{b,p,j}^{k1}, \forall b \in B, \forall k1\ and\ k2 \in K, \forall p\ and\ r \in I, \tag{10}$$

where $F_{b,p,j}^{k2}$ is the mass flow rate of product *p* in the outlet stream *k2* of stage *j* when utilizing feedstock *b*, $F_{b,r,j}^{k1}$ is the mass flow rate of reactant *r* in the inlet stream *k1* of stage *j* when utilizing feedstock *b*, $F_{b,p,j}^{k1}$ is the mass flow rate of product *p* in the inlet stream of stage *j* when utilizing feedstock *b*, and $\Phi_{b,p,r,j}^{k2}$ is the yield of product *p* from reactant *r* in the outlet stream *k* of stage *j* when utilizing feedstock *b*.

The mass balance constraint for the feedstock storage and handling, mixers, pumps, bypass, and heat exchangers are given by

$$F_{b,i}^{k2} = \sum_{k1=1}^{n_k} F_{b,i}^{k1}, \forall i \in I, \forall k2 \in K \quad , \tag{11}$$

where $F_{b,i}^{k1}$ is the mass flow rate of component *i* in the inlet stream *k* when utilizing feedstock *b* and $F_{b,i}^{k2}$ is the mass flowrate of component *i* in the outlet stream *k2* when utilizing feedstock *b*.

The amount of solids at any stage *j* is controlled by

$$F_{b,i,j}^{k} \leq \alpha_{b,i,j}^{k} \times F_{b,j}^{k}, \forall b \in B, \forall k \in K, \forall i \in I, \forall j \in J \quad , \tag{12}$$

and

$$F_{b,j}^{k} = \sum_{i=1}^{n_i} F_{b,i,j}^{k}, \forall b \in B, \forall k \in K, \forall j \in J \quad , \tag{13}$$

where $F_{b,i,j}^{k}$, $F_{b,j}^{k}$, and $\alpha_{b,i,j}^{k}$ is the mass flow rate of component *i-*, the total mass flow rate-, and the mass fraction of component *i-* in the stream *k* of stage *j* when utilizing feedstock *b,* respectively.

The catalyst loading at any stage *j* is controlled by

$$F_{b,j}^{k} \leq \sum_{i'} F_{b,i,j}^{k} \times \beta_{b,j}^{k}, \forall k \in K, \forall j \in J \quad , \tag{14}$$

where $\beta_{b,j}^{k}$ is the catalyst loading per kg of incoming feed in the stream *k* of stage *j* when utilizing feedstock *b*.

The mass balance of component *i* in the outlet key-stream *k* in the separator, washer, microfiltration, belt filter press, centrifuge, nanofiltration, evaporation, distillation, activated carbon column, reactive extraction, back extraction, electrodialysis, crystallization, ion-exchange column, solvent purification, flash column, and dryer is given by

$$F_{b,i,j}^{k2} = F_{b,i,j}^{k1} \times \zeta_{b,i,j}^{k2}, \forall b \in B, \forall k1 \; and \; k2 \in K, \forall i \in I, \forall j \in J \quad , \tag{15}$$

$$F_{b,i,j}^{k1} = F_{b,i,j}^{k1} \times \left(1 - \zeta_{b,i,j}^{k2}\right), \forall b \in B, \forall k1 \in K, \forall i \in I, \forall j \in J \quad , \tag{16}$$

where $\zeta_{b,i,j}^{k2}$ represents the recovery of component *i* in the outlet stream *k* when utilizing feedstock *b,* $F_{b,i,j}^{k1}$ is the mass flowrate of component *i* in the inlet stream *k* when utilizing feedstock *b* and $F_{b,i,j}^{k2}$ is the mass flowrate of component *i* in the outlet stream *k2* when utilizing feedstock *b*.

The feedstock purchase is bounded by its availability (*Θ)* and minimum purchase amount (*Y*):

$$\Theta \geq Feed \geq Y. \tag{17}$$

2.2 Energy balance constraints

Where applicable, the units of power, heating, and cooling in subsequent equations are in kW. The power ($P_{b,j1}$) consumed during the processing of feedstock *b* in stage *j1* is given by

$$P_{b,j1} = \sum_{k1} F_b^{k1} \times \emptyset_{j1}, \forall b \in B, \forall j1 \in J \tag{18}$$

where *J = {FSH, PRER, DEACE, FERM, BFP, CF,}* and $\emptyset_{j1}$is the power required per kg of feed rate in stage *j1.*

The power consumed during the processing of feedstock *b* in desalting electrodialysis ($P_{b,DED}$) and water splitting electrodialysis ($P_{b,WSED}$) is given by

$$P_{b,DED} = \sum_{k1} F_{b,i}^{k1} \times \emptyset_{DED}, \forall b \in B, \forall i1 \in I \tag{19}$$

and

$$P_{b,WSED} = \sum_{k1} F_{b,i}^{k1} \times \emptyset_{WSED}, \forall b \in B, \forall i1 \in I, \tag{20}$$

where *I = {NASA}* and $\emptyset_{DED}$ and $\emptyset_{WSED}$are power required per kg of sodium succinate in desalting electrodialysis and water splitting electrodialysis, which are ~3.5 and ~2.5 kWh per kg of sodium succinate, respectively.[26]

The power consumed during the processing of feedstock *b* in crystallizer ($P_{b,CRYS}$) is given by

$$V_{b,CRYS} = \sum_{(k1,i)} \frac{F_{b,i}^{k1}}{\rho_{b,i}^{k1}} \times \tau_{CRYS}, \forall b \in B \tag{21}$$

and

$$P_{b,CRYS} = V_{b,CRYS} \times \emptyset_{CRYS}, \forall b \in B, \tag{22}$$

where $V_{b,CRYS}$ is the volume of crystallizer in m^3, $\rho_{b,i}^{k1}$ is the density of component *i* in the inlet stream *k1*, τ_{CRYS} residence time (1200 s), and $\emptyset_{CRYS}$is the power required in a crystallizer, which is 2 kW per m^3.[28]

The power consumed during the processing of feedstock *b* in filtration ($P_{b,FIL}$) is given by

$$A_{b,FIL} = \frac{\sum_{(k1,i)} \frac{F_{b,i}^{k1}}{\rho_{b,i}^{k1}}}{\psi_{FIL}}, \forall b \in B \tag{23}$$

and

$$P_{b,FIL} = A_{b,FIL} \times \emptyset_{FIL}, \forall b \in B, \tag{24}$$

where $A_{b,FIL}$ is the area of rotary vacuum filter in m^2, ψ_{FIL} is flux, which is 400 $L.m^{-2}.h^{-1}$, and $\emptyset_{FIL}$is the power required (0.8 kW per m^2) in crystallizer.[28]

The energy ($E_{b,j1}$) required in kWh per m^2 during the processing of feedstock *b* is determined by a relation proposed by[29] and as follows:

$$E_{b,j1} = \frac{\beta_{j1}}{(\psi_{j1} \times \eta_{j1})}, \forall b \in B, \forall j1 \in J \quad (25)$$

where *J = {MF, NF},* β_{j1} is the energy required at membrane surface, which is 50 W per m^2, ψ_{j1} is membrane flux, which is 20 L.m^{-2}.h^{-1} for microfiltration and 50 L.m^{-2}.h^{-1} for nanofiltration, and η_{j1} is membrane efficiency (50%).

Once the energy required is calculated then the power consumption ($P_{b,j2}$) in the membrane can be calculated as follows:

$$A_{b,j1} = \frac{\sum_{(k1,i)} \frac{F_{b,i}^{k1}}{\rho_{b,i}^{k1}}}{\psi_{j1}}, \forall b \in B, \forall, \forall i \in I, \forall j1 \in J \quad (26)$$

and

$$P_{b,j2} = A_{b,j1} \times E_{b,j1}, \forall b \in B, \forall j1 \in J, \quad (27)$$

where $A_{b,j1}$ is the area of micro- and nano- filtration in m^2.

The power ($P_{b,pump}$) consumed in the pump can be calculated as:

$$P_{b,pump} = \frac{\sum_{i=1}^{n_i} \rho_{b,i} \times \left(P_b^{k2} - P_b^{k1}\right)}{\eta_{pump}} \quad (28)$$

where $\rho_{b,i}$ is the volumetric density of component *i*, P_b^{k2} is the outlet pressure, P_b^{k1} is the inlet pressure, η_{pump} is the pump efficiency.

For each unit operation involved in the processing of feedstock *b*, the following energy balance constraint was used:

$$\sum_{i=1}^{n_i} F_{b,i,j}^{k1}.cp_{b,i,j}^{k1}.T_{b,j}^{k1} + Q_{b,j} = \sum_{i=1}^{n_i} F_{b,i,j}^{k2}.cp_{b,i,j}^{k2}.T_{b,j}^{k2}, \ \forall b \in B, \forall j \in J, \forall k1 \ and \ k2 \in K \quad (29)$$

where $Q_{b,j}$ is the heat duty of stage *j*, $cp_{b,i,j}^{k1}$and $cp_{b,i,j}^{k2}$ are the specific heat of component *i* at the inlet (*k1*) and outlet (*k2*) conditions of stage *j*, respectively. $T_{b,j}^{k1}$ and $T_{b,j}^{k2}$are the temperature of the inlet and outlet conditions of stage *j*.

Heat balance in the reboiler is determined by a relation proposed by [30] and rearranged as:

$$Q_{b,j1} = (1+R)\sum_{i=1}^{n_i} f_{b,i}^{btm}\lambda_i \quad (30)$$

The cooling heat load needed for the condenser is given by:

$$Q_{b,j1} = -(1+R)\sum_{i=1}^{n_i} f_{b,i}^{dis}\lambda_i \quad (31)$$

where $f_{b,i}^{dis}$ and $f_{b,i}^{btm}$ are component molar flow rate in distillate and bottom, respectively, and λ_i is the latent heat of component *i*.

2.3 Economic analysis constraints

A techno-economic model was formulated to conduct multiperiod economic analysis, which includes constraints for estimating the total capital investment (TCI) and the total cost of manufacturing (TCOM).

2.3.1 Total capital investment

As shown in Table S6, the TCI consists of the total direct costs (TDC), total indirect costs (TIDC), working capital costs, and land costs. The TDC includes the costs related to the equipment, site development, and warehouse, along with additional expenses for the piping and instrumentation. Similarly, the TIDC includes portable and field expenses, project contingency costs, and home and office construction costs. The equipment cost was determined using the relevant economic data from literature, which are summarized in Table S7.

The power law was used to estimate the capacity-adjusted capital cost $C_{b,j}$ as follows:

$$C_{b,j} = C_{j,o} \times \left(\frac{Q_j}{Q_{j,o}}\right)^{a_j} \times I_j, \forall b \in B,\ \forall j \in J \quad (32)$$

where $C_{j,o}$ is the cost of the baseline equipment or stage j, Q_j and $Q_{j,o}$ are the adjusted and baseline capacities of equipment j, a_j is a scaling exponent, which varies depending on the type of equipment j, and $I_{j,o}$ is the installation factor of equipment j. Once the capacity-adjusted equipment cost has been determined, the capital cost of the equipment is then updated to the year of analysis using the chemical engineering plant cost index (CEPCI):

$$UC_{b,j} = C_{b,j}\left(\frac{CEPCI_{2020}}{CEPCI_{ref}}\right),\ \forall b \in B,\ \forall j \in J \quad (33)$$

where $UC_{b,j}$ is the updated equipment cost in the year of interest, and $CEPCI_{2020}$ and $CEPCI_{ref}$ are the index values corresponding to the year 2020 and baseline year, respectively.

The total installation cost is estimated as follows:

$$TIC_b = \sum_j UC_{b,j},\ \forall b \in B \quad (34)$$

The factor approach (Table S6) is used to calculate the TDC and TIDC after the total installation cost of equipment has been determined in the year of interest.

The total direct- and indirect- installed costs are estimated as follows:

$$TDC_b = TIC_b + TIC_b(x2 + x3 + x4),\ \forall b \in B \quad (35)$$

and

$$TIDC_b = TDC_b \times (y1 + y2 + y3 + y4 + y5),\ \forall b \in B \quad (36)$$

where *TDC* is the total direct cost, *TIDC* is the total indirect cost, x2, x3, x4, y1, y2, y3, y4, and y5 are the cost factors listed in Table S6.

Once TDC and TIDC costs have been estimated, the above-mentioned costs are summed to yield the fixed capital investment (FCI).

$$FCI_b = TDC_b + TIDC_b, \ \forall b \in B, \tag{37}$$

The working capital (WC_b), at 5% of the fixed capital investment, and land cost (L_b), at 6% of the total installation costs are subsequently added to obtain the TCI.

$$WC_b = FCI_b \times 5\%, \ \forall b \in B, \tag{38}$$

$$L_b = TIC_b \times 6\%, \forall b \in B, \tag{39}$$

$$TCI_b = FCI_b + WC_b + L_b, \forall b \in B, \tag{40}$$

2.3.2 Total cost of manufacturing

The total cost of manufacturing (TCOM) consists of the total direct manufacturing cost (TDMC), total fixed manufacturing cost (TFMC), and general manufacturing cost (GMC). The TDMC includes the costs of the raw materials (CRM_b), utilities (CUT_b), operating labor (COL_b), waste treatment (CWT_b), maintenance and repairs, and patents and royalties. Fixed manufacturing costs include depreciation, local taxes, insurance, and plant overhead. General costs are related to the administration, distribution, and selling expenses in addition to the research and development costs. The cost of the raw materials and utilities can be estimated using the mass and energy balance constraints detailed in Sections 2.1 and 2.2. The unit price of chemicals, utilities, and wastewater treatment are summarized in Table S8 and S9. The COL_b is determined as follows:

$$COL_b = TIC_b \times 1.6\%, \ \forall b \in B, \tag{41}$$

A CWT_b of \$2.5 per 1000 gal was applied. $TCOM_b$ can be estimated as follows:

$$TCOM_b = f_1 COL_b + f_2 FCI_b + f_3 (CUT_b + CRM_b + CWT_b), \ \forall b \in B, \tag{42}$$

where f_1, f_2, and f_3 are multipliers given in Table S6, and F_{CI} is the fixed capital investment.

The total process revenues (*Rev*) from the sale of bio-SA and dry distiller solid (DDS) are given by

$$Rev_b = \sum_{p=1}^{n_p} F_p P_p, \tag{43}$$

where n_p is the number of products, F_p is the mass flow rate of product *p*, and P_p is the wholesale price of product *p*.

The non-discounted cash flow, $NCF_{b,n}$ for the year *n* is given as:

$$NCF_{b,n} = -r_n TCI_b + a_n WC_b + (Rev_b - TCOM_b)(1 - tax) + D \cdot tax, \tag{44}$$

where r_n is the ratio of total capital investment consumed during year *n*, *D* is depreciation, and W_C is working capital. a_n is a parameter equal to -1 during the year 3, 1 during the last year of the project, and zero for all other years.

The net present value is defined as:

$$NPV_b = \sum_{n=0}^{20} \frac{NCF_{b,n}}{(1+r)^n} . \quad (45)$$

Table S6 Methodology to determine total capital investment and total cost of manufacturing.[1,31]

Parameters	**Value**
x_1, Total installation cost	100%
x_2, Warehouse	4% of Inside battery limits (ISBL)
x_3, Site development	9% of ISBL
x_4, Additional piping	5% of ISBL
Total direct costs (TDC)	$\sum_{i=1}^{4} x_i$
y_1, Prorateable costs	10% of TDC
y_2, Field expenses	10% of TDC
y_3, Home office & construction fee	20% of TDC
y_4, Project contingency	10% of TDC
y_5, Other costs (start-up, permits, etc.)	10% of TDC
Total indirect costs (TIDC)	$\sum_{i=1}^{5} y_i$
Fixed capital investment (FCI)	TDC + TIDC
Land	6% of installation costs
Working capital	5% of FCI
Total capital investment (TCI)	FCI + Land + Working capital
f1	2.2 times the cost of labor
f2	1.1 times of FCI
f3	1.05 times the cost of utility and raw material

Table S7 Equipment cost quoted by vendors and literature.[1,9,28,31–33]

Equipment	Purchase cost ($)	Scaling exponent	Installation factor	Scaling value	Capacity	Year
Pump	22,500	0.8	2.3	402,194 kg/h	74.57 kW	2009
Flash	511,000	0.7	2	264,116 kg/h	23' X 48' – 416 m^3	2009
Mechanical separator	3,294,700	0.8	1.7	31,815 kg/h	384 m^2	2010
Condenser	34,000	0.7	2.2	2 Gcal/h	2 Gcal/h	2010
Heater	92,000	0.7	2.2	-8 Gcal/h	- 8 Gcal/h	2010
Cooler	85,000	0.7	2.2	8 Gcal/h	8 Gcal/h	2010
Acid thermal hydrolysis (ATH) reactor	19,812,400	0.6	1.5	83,333 kg/h	9' x 30' – 2 min. residence time	2009
ATH reactor after deacetylation	24,600,000	0.6	1.5	63,166 kg/h	9.8'x 33.7'	2013
Alkaline hydrolysis reactor	614,000	0.7	2.2	50 m^3	50 m^3	2018
Conditioning vessel	236,000	0.7	2	410,369 kg/h	447 m^3, 1hr residence time	2009
Hot water washing reactor	3,840,000	0.7	2	421,776 kg/h	946 m^3	2009
Saccharification reactor	3,840,000	0.7	2	421,376 kg/h	946 m^3	2009
Belt filter press	3,294,700	0.8	1.7	31,815 kg/h	384 m^2	2010
Fermenter	590,000	1	1.45	757 m^3	757 m^3	2007
Deacetylation vessel 1	780,000	0.7	1.7	277,167 kg/h	14' x 30' vessel	2013
Deacetylation vessel 2	110,000	0.8	1.7	277,167 kg/h	14' x 30' vessel	2013
Micro- and nano- filtration	1000/m^2	1	1	-	-	-
Centrifuge	170,000	1	1.45	96 m^3/h	96 m^3/h	1990
Evaporator	3,801,095	0.6	1	393,100 kg/h	368 gpm	2010
Vacuum distillation	511,000	0.7	2	264,116 kg/h	14' dia. x 76' tall, 32 trays	2009
Activate carbon vessel	614,000	0.7	1	50 m^3	50 m^3	2018
Acidification vessel	614,000	0.7	1	50 m^3	50 m^3	2018
Extraction column	511,000	0.7	2	264,116 kg/h	23' X 48' – 416 m^3	2009
Back extraction column	511,000	0.7	2	264,116 kg/h	23' X 48' – 416 m^3	2009
Ion-exchange column	5,250,000	0.9	1.8	53,204 kg/h	53,204 kg/h	2014
Electrodialysis	1,410,000	0.7	1	208 m^3/h	200 m^3/h	1993
Water splitting electrodialysis	1,410,000	0.7	1	208 m^3/h	200 m^3/h	1993
Vacuum rotary filters	671,000	0.65	1.5	50 m^3	50 m^3	2014
Crystallizer	428,200	0.675	2	25 m^3	25 m^3	2014
Reactive crystallizer	428,200	0.675	2	25 m^3	25 m^3	2014
Thermal cracker	241,400	0.7	1.5	2 Gcal/h	2.42 Gcal/h	2011
Solvent purification vessel	614,000	0.7	1	50 m^3	50 m^3	2018
Dryer	10,500,000	0.6	1	54,431 kg/h	100 m^3	1990
Washer	614,000	0.7	1	50 m^3	50 m^3	2018

Table S8 Deterministic price of chemicals.

Chemicals	Price (USD/t)
Corn stover[34]	80
S. japonica[35]	68
Steam[31]	12.68
Glucose[36]	988
Glycerol[37,38]	750
Acid[1]	87.78
Ammonia[38]	550
Enzymes[39]	5000
Succinic acid[40]	2000
Carbon dioxide[38]	30
Sodium hydroxide[1]	149.16
Magnesium carbonate[41]	480
Magnesium hydroxide[38]	270
Sodium carbonate[42]	300
Activated carbon[43]	1300
Octanol[44]	5000
Tri-octylamine[44]	1000
Tri-methyl amine	1000
Methanol[44]	547
Ammonium bisulfate[38]	260
Phosphoric acid[38]	420
Dry distiller solids[36]	50

Table S9 Deterministic price of utility and wastewater treatment.

Utility	Price
Electricity[45]	0.07 USD/kWh
Steam[45]	6.13 USD/GJ
Cooling water[46]	0.28 USD/GJ
Chilled water[31]	5 USD/GJ
Wastewater[31]	0.041 USD/m^3

Table S10 Inventory data for conversion, transportation, and end-use phase obtained from optimal processing pathways (OPP) shown in Figure 4 of the manuscript through robust (stochastic) optimization. Data is scaled based on one kg bio-SA.

Item	OPP of glucose	OPP of glycerol	OPP of corn Stover	OPP of *S. Japonica*
Feedstock				
Glucose (kg)	1.051	-	-	-
Glycerol (kg)	-	1.136	-	-
Corn sotver (kg)	-	-	3.482	-
S. Japonica (kg)	-	-	-	8.323
Pretreatment				
Water (kg)	-	-	14.181	22.888
Acid (kg)	-	-	0.077	-
Ammonia (kg)	-	-	0.080	-
Enzyme (kg)	-	-	0.022	0.01
Electricity (kWh)	-	-	0.473	1.339
Heating, LP-steam (kWh)	-	-	0.467	1.387
Fermentation				
Process water (kg)	10.638	19.528	14.809	25.896
Ammonia (kg)	0.016	0.031	0.424	0.006
CO2 added (kg)	0.215	0.525	0.251	0.354
NaOH (kg)	0.783	-	-	0.954
$Mg(OH)_2$	-	0.695	-	-
Electricity (kWh)	0.715	1.174	1.077	1.642
Heating, LP-steam (kWh)	0.114	0.201	-	-
Wastewater	-	-	0.006	0.01
DDS Purification				
Electricity (kWh)	-	-	0.123	0.211
Heating, LP-steam (kWh)	-	-	5.21	8.721
Purification				
Acid (kg)	0.922	1.146	1.146	1.146
NaOH (kg)	0.006	-	-	-
H3PO4 (kg)	0.006	-	-	-
Electricity (kWh)	0.212	0.062	0.067	0.088
Heating, LP-steam (kWh)	6.712	2.049	2.049	2.049
Wastewater	0.013	0.022	0.027	0.046
Oxygen to air (kg)	0.398	0.334	0.516	1.087
Ammonia (kg)	-	-	0.036	-
transportation				
Each raw material/ Product transport (km)	300	300	300	300
Substituted products				
Dried distillers' grains with solubles (kg)	-	-	1.474	6.249
Petroleum-based succinic acid (kg)	1.0	1.0	1.0	1.0

Table S11 Inventory data for SA production via hydrogenation of maleic anhydride. Data is scaled based on one kg bio-SA.[47]

Item	Input
Maleic anhydride (kg)	0.89
Hydrogen (kg)	0.25
Water (kg)	0.30
Nitrogen (kg)	0.073
Palladium Catalyst (kg)	1.00E-03
Natural gas (kg)	0.10
Electricity (kWh)	0.36
Waste treatment	0.32
Transportation (km)	300

Table S12 Ecoinvent 3.6 and USLCI databases selected for life cycle impact assessment of bio-SA production process.

Processes	Databases
Sugar, from sugar beet {GLO}| market for | APOS, S	Ecoinvent 3.6
Glycerine {RoW}| treatment of waste cooking oil, purified, esterification | Cut-off, S	Ecoinvent 3.6
Corn stover, at conversion plant, 2022/ton/RNA	USLCI
S. Japonica (this feedstock is not available in SimaPro databases, the life cycle inventory of seaweed cultivation and harvesting was estimated from.[48])	-
Water, Tap water {GLO}| market group for | APOS, S	Ecoinvent 3.6
Sulfuric acid {RoW}| market for sulfuric acid | APOS, S	Ecoinvent 3.6
Ammonia, liquid {RoW}| market for | APOS, S	Ecoinvent 3.6
Enzyme, Cellulase, Novozyme Celluclast/kg/RER	USLCI
Sodium hydroxide, production mix, at plant/RNA	USLCI
Wastewater from anaerobic digestion of whey {GLO}| market for | Conseq, S	Ecoinvent 3.6
Phosphoric acid, fertiliser grade, without water, in 70% solution state {GLO}| market for | APOS, S	Ecoinvent 3.6
Transport, freight, lorry 16-32 metric ton, euro5 {RoW}| market for transport, freight, lorry 16-32 metric ton, EURO5 | APOS, S	Ecoinvent 3.6
Distillers dried grains with solubles, 2022/kg/RNA	USLCI
Electricity, medium voltage {US}| market group for | APOS, S	Ecoinvent 3.6
Steam, in chemical industry {RoW}| market for steam, in chemical industry | APOS, S	Ecoinvent 3.6

Table S13 The definition of environmental indicators.

Impact Category (Units per kg emission)	**Definition**	**Relevant LCI Data**
Abiotic Depletion Potential (kg Sb eq)	Collective quantification of impact caused by extraction of minerals due to inputs in the system	Extraction of mineral resources
Abiotic Fossil Fuel Depletion Potential (MJ)	Surplus energy (lower hating value) per extracted MJ per kg or m^3 fossil fuel	Extraction of fossil fuel resources
Global Warming Potential (kg CO_2 eq)	GWP potential for time horizon 100 years: Impact caused by emissions of greenhouse gases	CO_2, NO_2, CH_4, CFCS, HCFCS, CH_3BR
Ozone Depletion Potential (kg CFC-11 eq)	Thinning of stratospheric ozone layer due to anthropogenic emissions due to greenhouse gases	CFCS, HCFCS, CH_3BR, Halons
Human Toxicity Potential (1,4-Dichlorobenzene eq)	Potential impacts of toxic substances on human health present in environment	Human toxic substances
Fresh Water Aquatic Ecotoxicity Potential (1,4-Dichlorobenzene eq)	Potential impact of toxic substances on aquatic ecosystems, as a result of emissions of toxic substances to air, water and soil.	Toxic substances with a reported lethal concentration to fish
Marine Aquatic Ecotoxicity Potential (1,4-Dichlorobenzene eq)	Potential impacts of toxic substances on marine ecosystems, as a result of emissions of toxic substances to air, water, and soil	Reported toxic substances
Terrestrial Ecotoxicity Potential (1,4-Dichlorobenzene eq)	Potential impacts of toxic substances on terrestrial ecosystems (land-dependent organisms), as a result of emissions of toxic substances to air, water and soil	Reported toxic substances
Photochemical Oxidation Potential (kg C_2H_4 eq)	Formation of reactive chemical compounds like ozone, by the action of sunlight on emissions of air pollutants	PM10, NH_3, SO_2, NO_X, and NMVOC
Acidification Potential (kg SO_2 eq)	Formation of acidic compounds as a result of manufacturing process	SO_X
Eutrophication Potential (kg PO_4 eq)	Collective quantification of formation of phosphorus compounds	Nitrogen and Phosphorus compounds

Table S14 Uncertainties in chemical prices based on historical prices. Uncertainties presented in color cells applicable to both local sensitive analysis and stochastic analysis. In comparison, uncertainties presented in white cells only applicable to stochastic analysis. The mean values of chemical prices are reported in Table S8.

Chemicals	**Min**	**Max**
Corn Stover[49]	31.3%	37.5%
S. Japonica[50]	15.0%	50.0%
Steam	19.6%	8.8%
Glucose[36]	41.3%	15.0%
Glycerol[37]	20.0%	15.0%
Acid[38]	15.0%	16.2%
Ammonia[36]	58.7%	54.7%
Enzymes	15.0%	15.0%
Succinic acid[51]	20.0%	20.0%
Carbon dioxide	15.0%	15.0%
NaOH[52]	15.0%	75.0%
$MgCO_3$	15.0%	15.0%
$Mg(OH)_2$	15.2%	14.8%
Na_2CO_3[52]	51.3%	0.0%
Activated carbon[52]	0.0%	15.4%
Octanol	15.0%	15.0%
Trioctylamine	15.0%	15.0%
Trimethylamine	15.0%	15.0%
Methanol[52]	48.4%	0.0%
Ammonium bisulfate[52]	36.5%	15.4%
H_3PO_4[52]	48.1%	7.1%
Dry distillery solids[36]	16.0%	57.1%

Note that ±15 variation is assumed for chemicals that don't have historical cost data available in literature.

Table S15 Uncertainties in the process economic parameters.[1,50,51] The data presented in this table is applicable to both local sensitive analysis and stochastic analysis. In sensitivity analysis, the impact of each variable on economics was estimated by varying its maxima and minima while keeping all other variables constant. Whereas, in stochastic analysis, uniform distribution function is used to select random number between min and max values for 5000 scenarios.

Process economic parameters	**Min**	**Max**
Equipment costs	20%	50%
Utility costs	20%	20%
Plant capacity	20%	20%
Operating hours	-10%	5%
Discount rate	20%	20%
Income tax rate	20%	20%

Mean value for utility costs is reported in Table S9, whereas mean value is 30 kt/y, 10%, and 30% for plant capacity, discount rate, and income tax, respectively.

Table S16 Uncertainties in the biorefinery processing (technical) parameters.[1,50,51] The data presented in this table is applicable to both local sensitive analysis and stochastic analysis. In sensitivity analysis, the impact of each variable on economics was estimated by varying its maxima and minima while keeping all other variables constant. Whereas, in stochastic analysis, uniform distribution function is used to select random number between min and max values for 5000 scenarios.

Technical parameters	Min	Max
Yield in pretreatment reactors		
1. Acid thermal hydrolysis of corn stover	10%	5%
2. Acid thermal hydrolysis of *S. japonica*	10%	2%
3. Deacetylation	10%	10%
4. Alkaline pretreatment	10%	10%
5. Hot water washing	10%	10%
6. Enzymatic hydrolysis	10%	5%
Yield in Fermentation		
1. F1 – Separate hydrolysis and fermentation of corn stover using *A. succinogen*	10%	10%
2. F2 – Batch fermentation of corn stover using *A. succinogen*	10%	10%
3. F3 – Simultaneous fermentation and saccharification of corn stover using *A. succinogen*	10%	10%
4. F4 – Dual phase fermentation of *S. japonica* using *E. coli*	10%	10%
5. F5 – Separate hydrolysis and fermentation of *S. japonica* using *A. succinogen*	10%	10%
6. F6 – Dual phase fermentation of glucose using *E. coli*	10%	2%
7. F7 – Batch fermentation of glucose using *A. succinogen*	10%	10%
8. F8 – Fed-batch fermentation of glycerol using glycerol	10%	2%
9. F9 – Fed-batch fermentation of glycerol using glycerol	10%	10%
Fermentation titer		
F1 – Separate hydrolysis and fermentation of corn stover using *A. succinogen*	10%	10%
F2 – Batch fermentation of corn stover using *A. succinogen*	10%	10%
F3 – Simultaneous fermentation and saccharification of corn stover using *A. succinogen*	10%	10%
F4 – Dual phase fermentation of *S. japonica* using *E. coli*	10%	10%
F5 – Separate hydrolysis and fermentation of *S. japonica* using *A. succinogen*	10%	10%
F6 – Dual phase fermentation of glucose using *E. coli*	10%	10%
F7 – Batch fermentation of glucose using *A. succinogen*	10%	10%
F8 – Fed-batch fermentation of glycerol using glycerol	10%	10%
F9 – Fed-batch fermentation of glycerol using glycerol	10%	10%
Efficiency in purification technology		
Electrodialysis	10%	10%
Direct crystallization	10%	10%
Reactive extraction	10%	-
Ion-exchange column	10%	10%

Reactive crystallization	10%	-
Membrane technology	10%	10%

The mean values of pretreatment yield, fermentation yield, and titer are reported in Tables S2 and S3. Whereas, the mean values of purification technologies are reported in Section 1.6.

Mean value for utility costs is reported in Table S9, whereas mean value is 30 kt/y, 10%, and 30% for plant capacity, discount rate, and income tax, respectively.

Table S17 Processing pathways and process economic indicators of the top 100 topologies.

Ranking	Processing pathway	NPV (M$)	MPSP ($/kg)	TCI (M$)	COM (M$/y)
1	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	57.18	1.60	71.33	39.53
2	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	56.19	1.61	72.54	39.56
3	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	54.25	1.62	33.80	43.54
4	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	53.49	1.62	34.82	43.55
5	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	51.67	1.64	74.84	40.17
6	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	50.48	1.65	38.45	43.64
7	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	49.86	1.65	36.63	44.04
8	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	47.89	1.66	73.23	41.21
9	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	47.14	1.67	62.72	43.82
10	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,2),(9,1),(10,1),(11,1)	46.90	1.67	74.44	41.24
11	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	45.97	1.68	64.07	43.86
12	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	44.96	1.68	35.70	45.21
13	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,2),(9,1),(10,1),(11,1)	44.20	1.69	36.73	45.22
14	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	43.48	1.69	81.15	40.96
15	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	43.08	1.70	41.81	44.70
16	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	42.48	1.70	82.36	40.98
17	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,1),(9,1),(10,1),(11,1)	42.38	1.70	76.74	41.85
18	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	42.31	1.70	42.84	44.71
19	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,6),(7,1),(8,1),(9,1),(10,1),(11,1)	41.49	1.71	40.11	45.29
20	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	41.26	1.71	72.30	42.35
21	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	40.69	1.71	66.79	44.57
22	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,1),(9,1),(10,1),(11,1)	40.57	1.72	38.53	45.72
23	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	38.69	1.73	44.64	45.20
24	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	37.96	1.73	84.65	41.60
25	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	37.85	1.73	64.62	45.49
26	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,2),(9,1),(10,1),(11,1)	36.68	1.74	65.97	45.53
27	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	34.19	1.76	83.05	42.63
28	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	33.79	1.76	43.71	46.37
29	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,4),(7,3),(8,2),(9,1),(10,1),(11,1)	33.19	1.77	84.26	42.66
30	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,4),(7,3),(8,2),(9,1),(10,1),(11,1)	33.02	1.77	44.74	46.38
31	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,6),(7,1),(8,1),(9,1),(10,1),(11,1)	32.43	1.77	73.84	43.98
32	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,1),(9,1),(10,1),(11,1)	31.40	1.78	68.69	46.24
33	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	31.35	1.78	74.02	45.46
34	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	30.18	1.79	75.38	45.50
35	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,4),(7,3),(8,1),(9,1),(10,1),(11,1)	29.40	1.79	46.54	46.88
36	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,4),(7,3),(8,1),(9,1),(10,1),(11,1)	28.67	1.80	86.55	43.27
37	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	24.90	1.83	78.09	46.21
38	(1,4),(3a,4),(3b,6),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	23.78	1.83	32.35	50.18
39	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	22.07	1.85	66.50	47.98
40	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	22.06	1.85	75.93	47.13

41	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,4),(7,3),(8,2),(9,1),(10,1),(11,1)	20.89	1.85	77.28	47.18
42	(1,1),(2,1),(3a,5),(3b,1),(4,2),(5,3),(6,5),(7,1),(8,3),(9,1),(10,1),(11,1)	18.51	1.87	79.54	45.96
43	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,4),(7,3),(8,1),(9,1),(10,1),(11,1)	15.61	1.89	79.99	47.88
44	(1,2),(2,5),(3a,4),(3b,5),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	14.99	1.89	89.44	52.89
45	(1,4),(3a,4),(3b,6),(4,1),(5,3),(6,6),(7,1),(8,1),(9,1),(10,1),(11,1)	14.85	1.90	33.97	51.82
46	(1,2),(2,5),(3a,4),(3b,5),(4,2),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	13.30	1.91	91.21	52.98
47	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,6),(7,1),(8,1),(9,1),(10,1),(11,1)	13.19	1.91	68.07	49.62
48	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,2),(7,2),(8,2),(9,2),(10,1),(11,1)	12.80	1.91	62.95	49.71
49	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,2),(7,2),(8,3),(9,2),(10,1),(11,1)	12.40	1.91	61.78	49.97
50	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,2),(7,2),(8,2),(9,2),(10,1),(11,1)	11.70	1.92	31.59	52.84
51	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,2),(7,2),(8,3),(9,2),(10,1),(11,1)	11.10	1.92	30.58	53.11
52	(1,1),(2,1),(3a,5),(3b,1),(4,2),(5,3),(6,5),(7,1),(8,2),(9,1),(10,1),(11,1)	11.02	1.92	81.84	47.26
53	(1,1),(2,1),(3a,5),(3b,1),(4,2),(5,3),(6,5),(7,1),(8,1),(9,1),(10,1),(11,1)	10.18	1.93	83.00	47.24
54	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,2),(7,2),(8,1),(9,2),(10,1),(11,1)	8.76	1.94	64.68	50.30
55	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,2),(7,2),(8,1),(9,2),(10,1),(11,1)	8.38	1.94	32.94	53.34
56	(1,1),(2,1),(3a,5),(3b,1),(4,1),(5,3),(6,5),(7,1),(8,3),(9,1),(10,1),(11,1)	8.30	1.94	86.84	47.03
57	(1,2),(2,5),(3a,4),(3b,5),(4,2),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	5.70	1.96	91.34	54.56
58	(1,2),(2,5),(3a,4),(3b,5),(4,2),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	5.66	1.96	95.27	53.98
59	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,2),(7,2),(8,2),(9,2),(10,1),(11,1)	4.94	1.97	55.82	53.14
60	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,2),(7,2),(8,3),(9,2),(10,1),(11,1)	4.68	1.97	54.53	53.39
61	(1,2),(2,5),(3a,4),(3b,5),(4,2),(5,3),(6,4),(7,3),(8,2),(9,1),(10,1),(11,1)	4.01	1.97	93.11	54.65
62	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,2),(7,2),(8,2),(9,1),(10,1),(11,1)	3.61	1.97	64.77	51.38
63	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,2),(7,2),(8,3),(9,1),(10,1),(11,1)	3.20	1.98	63.60	51.64
64	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,2),(7,2),(8,2),(9,2),(10,1),(11,1)	2.69	1.98	38.05	53.77
65	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,2),(7,2),(8,2),(9,1),(10,1),(11,1)	2.51	1.98	33.41	54.50
66	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,2),(7,2),(8,3),(9,2),(10,1),(11,1)	2.09	1.99	37.04	54.05
67	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,2),(7,2),(8,3),(9,1),(10,1),(11,1)	1.91	1.99	32.40	54.78
68	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,2),(7,2),(8,2),(9,2),(10,1),(11,1)	1.75	1.99	70.87	50.86
69	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,2),(7,2),(8,3),(9,2),(10,1),(11,1)	1.34	1.99	69.70	51.12
70	(1,1),(2,1),(3a,5),(3b,1),(4,1),(5,3),(6,5),(7,1),(8,2),(9,1),(10,1),(11,1)	0.48	2.00	89.38	48.35
71	(1,3),(3a,5),(3b,8),(4,2),(5,3),(6,5),(7,2),(8,3),(9,1),(10,1),(11,1)	0.37	2.00	48.08	52.76
72	(1,1),(2,3),(3a,5),(3b,3),(4,2),(5,3),(6,5),(7,1),(8,3),(9,1),(10,1),(11,1)	0.34	2.00	72.70	51.50
73	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,2),(7,2),(8,1),(9,2),(10,1),(11,1)	0.29	2.00	57.89	53.81
74	(1,4),(3a,4),(3b,6),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	0.11	2.00	30.57	55.43
75	(1,1),(2,1),(3a,5),(3b,1),(4,1),(5,3),(6,5),(7,1),(8,1),(9,1),(10,1),(11,1)	-0.24	2.00	90.45	48.32
76	(1,4),(3a,4),(3b,6),(4,2),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	-0.37	2.00	31.35	55.42
77	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,2),(7,2),(8,1),(9,1),(10,1),(11,1)	-0.44	2.00	66.51	51.97
78	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,2),(7,2),(8,1),(9,2),(10,1),(11,1)	-0.64	2.00	39.40	54.27
79	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,2),(7,2),(8,1),(9,1),(10,1),(11,1)	-0.82	2.01	34.76	55.00
80	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,2),(7,2),(8,1),(9,2),(10,1),(11,1)	-2.30	2.02	72.61	51.45
81	(1,4),(3a,4),(3b,6),(4,2),(5,3),(6,4),(7,3),(8,1),(9,2),(10,1),(11,1)	-2.94	2.02	32.58	55.78
82	(1,2),(2,5),(3a,4),(3b,5),(4,2),(5,3),(6,4),(7,3),(8,1),(9,1),(10,1),(11,1)	-3.63	2.03	97.17	55.66
83	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,2),(7,2),(8,2),(9,1),(10,1),(11,1)	-4.26	2.03	57.64	54.81

84	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,2),(7,2),(8,3),(9,1),(10,1),(11,1)	-4.51	2.03	56.35	55.06
85	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,2),(7,2),(8,2),(9,1),(10,1),(11,1)	-6.51	2.05	39.88	55.44
86	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,2),(7,2),(8,3),(9,1),(10,1),(11,1)	-7.11	2.05	38.87	55.72
87	(1,3),(3a,5),(3b,8),(4,2),(5,3),(6,5),(7,2),(8,2),(9,1),(10,1),(11,1)	-7.20	2.05	49.58	54.14
88	(1,3),(3a,5),(3b,8),(4,2),(5,3),(6,5),(7,2),(8,1),(9,1),(10,1),(11,1)	-7.23	2.05	50.60	53.99
89	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,2),(7,2),(8,2),(9,1),(10,1),(11,1)	-7.45	2.05	72.69	52.52
90	(1,2),(2,5),(3a,4),(3b,5),(4,1),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	-7.46	2.05	105.52	55.22
91	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,2),(7,2),(8,2),(9,2),(10,1),(11,1)	-7.80	2.05	64.94	54.47
92	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,2),(7,2),(8,3),(9,1),(10,1),(11,1)	-7.86	2.06	71.53	52.78
93	(1,3),(3a,5),(3b,8),(4,1),(5,3),(6,5),(7,2),(8,3),(9,1),(10,1),(11,1)	-7.87	2.06	53.98	53.62
94	(1,4),(3a,4),(3b,6),(4,1),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	-7.96	2.06	36.35	56.27
95	(1,1),(2,3),(3a,5),(3b,3),(4,2),(5,3),(6,5),(7,1),(8,2),(9,1),(10,1),(11,1)	-8.04	2.06	75.02	53.00
96	(1,1),(2,3),(3a,4),(3b,3),(4,1),(5,3),(6,2),(7,2),(8,3),(9,2),(10,1),(11,1)	-8.06	2.06	63.65	54.71
97	(1,4),(3a,4),(3b,6),(4,1),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	-8.43	2.06	37.14	56.25
98	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,2),(7,2),(8,1),(9,1),(10,1),(11,1)	-8.91	2.06	59.71	55.48
99	(1,2),(2,5),(3a,4),(3b,5),(4,1),(5,3),(6,4),(7,3),(8,2),(9,2),(10,1),(11,1)	-9.15	2.06	107.30	55.31
100	(1,4),(3a,4),(3b,6),(4,2),(5,3),(6,4),(7,3),(8,3),(9,1),(10,1),(11,1)	-9.18	2.06	32.47	57.11

Table S18 Distribution of unique optimal processing pathways for glycerol, corn stover, and glucose using stochastic optimization for 5000 scenarios.

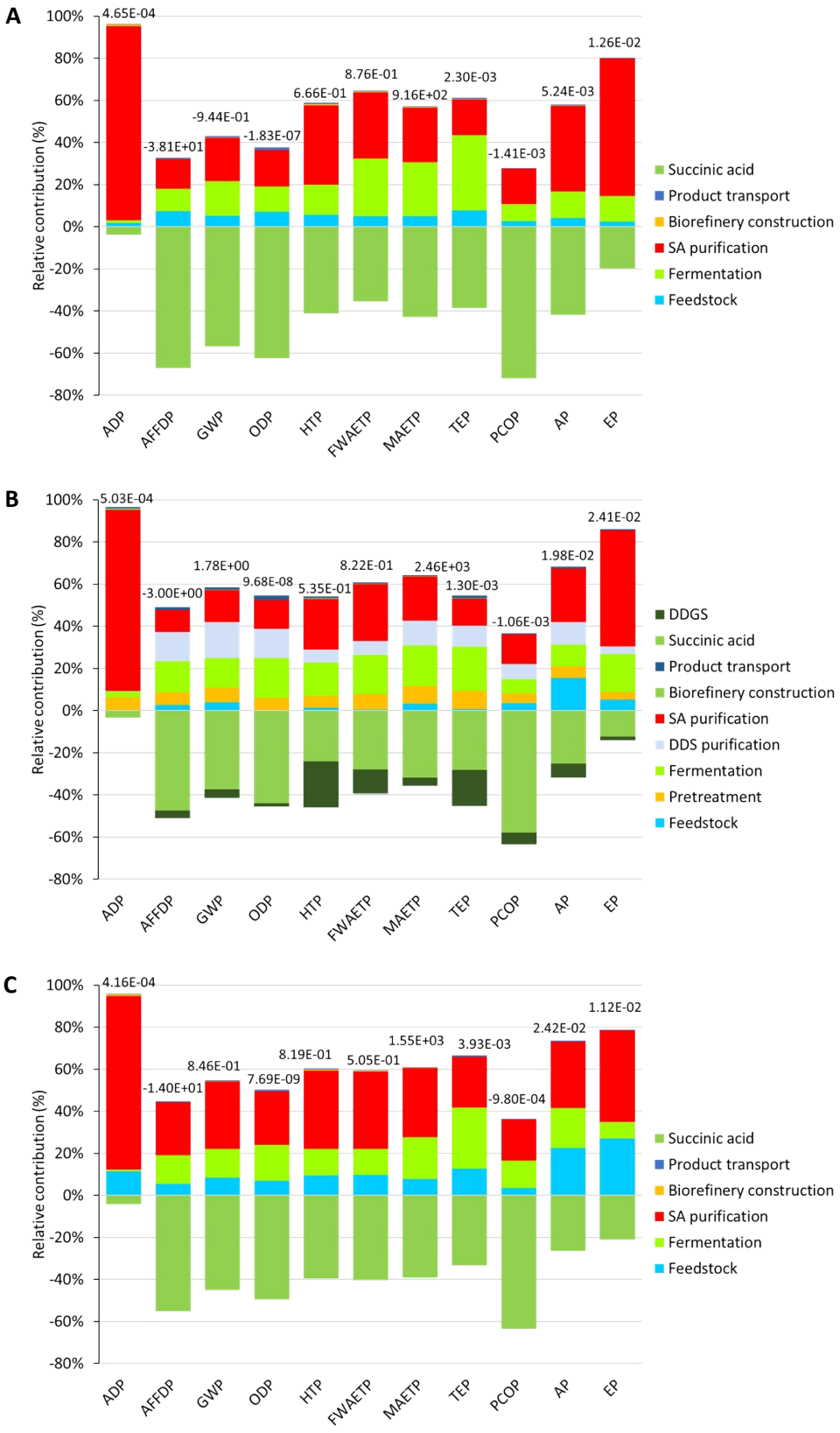
Pathway no.	Processing pathways for glycerol	Frequency	Relative distribution
1	(1,3),(3a,2),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	585	22.0%
2	(1,3),(3a,4),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	537	20.2%
3	(1,3),(3a,5),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	485	18.3%
4	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	207	7.8%
5	(1,3),(3a,5),(3b,8),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	203	7.6%
6	(1,3),(3a,2),(3b,8),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	192	7.2%
7	(1,3),(3a,5),(3b,8),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	162	6.1%
8	(1,3),(3a,4),(3b,8),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	144	5.4%
9	(1,3),(3a,2),(3b,8),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	126	4.7%
10	(1,3),(3a,3),(3b,8),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	6	0.2%
11	(1,3),(3a,3),(3b,8),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	4	0.2%
12	(1,3),(3a,3),(3b,8),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	3	0.1%
	Total	2654	100%
Pathway no.	**Processing pathways for corn stover**	**Frequency**	**Relative distribution**
1	(1,1),(2,1),(3a,5),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	403	29.5%
2	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	377	27.6%
3	(1,1),(2,1),(3a,2),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	349	25.5%
4	(1,1),(2,3),(3a,4),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	82	6.0%
5	(1,1),(2,3),(3a,2),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	56	4.1%
6	(1,1),(2,3),(3a,5),(3b,3),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	22	1.6%
7	(1,1),(2,1),(3a,5),(3b,1),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	15	1.1%
8	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	10	0.7%
9	(1,1),(2,1),(3a,5),(3b,1),(4,2),(5,3),(6,5),(7,2),(8,3),(9,1),(10,1),(11,1)	10	0.7%
10	(1,1),(2,1),(3a,2),(3b,1),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	10	0.7%
11	(1,1),(2,1),(3a,5),(3b,1),(4,2),(5,3),(6,5),(7,1),(8,3),(9,1),(10,1),(11,1)	9	0.7%
12	(1,1),(2,1),(3a,2),(3b,1),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	9	0.7%
13	(1,1),(2,1),(3a,4),(3b,1),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	6	0.4%
14	(1,1),(2,1),(3a,3),(3b,1),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	4	0.3%
15	(1,1),(2,1),(3a,5),(3b,1),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	4	0.3%
16	(1,1),(2,1),(3a,4),(3b,1),(4,2),(5,3),(6,2),(7,1),(8,3),(9,2),(10,1),(11,1)	1	0.1%
	Total	1367	100%
Pathway no.	**Processing pathways for glucose**	**Frequency**	**Relative distribution**
1	(1,4),(3a,4),(3b,6),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	183	18.7%
2	(1,4),(3a,2),(3b,6),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	161	16.4%
3	(1,4),(3a,5),(3b,6),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	154	15.7%
4	(1,4),(3a,4),(3b,6),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	153	15.6%
5	(1,4),(3a,5),(3b,6),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	147	15.0%
6	(1,4),(3a,2),(3b,6),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	130	13.3%
7	(1,4),(3a,4),(3b,6),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	19	1.9%
8	(1,4),(3a,5),(3b,6),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	16	1.6%
9	(1,4),(3a,2),(3b,6),(4,2),(5,3),(6,4),(7,3),(8,3),(9,2),(10,1),(11,1)	12	1.2%
10	(1,4),(3a,3),(3b,6),(4,1),(5,3),(6,6),(7,1),(8,1),(9,2),(10,1),(11,1)	3	0.3%
11	(1,4),(3a,3),(3b,6),(4,1),(5,3),(6,6),(7,2),(8,1),(9,2),(10,1),(11,1)	1	0.1%
	Total	979	100%

Table S19 Life cycle indicator results of bio-SA production through the optimal processing pathway (OPP) of glycerol, corn stover, glucose, and *S. japonica*. Indicator results are obtained by substituting equivalent amount of fossil fuel based succinic acid and dry distillery solids. Negative sign represents environmental savings while positive represent environmental burden.

Impact categories	OPP of glycerol	OPP of corn Stover	OPP of glucose	OPP of *S. Japonica*
ADP (kg Sb eq.)	4.65E-04	5.03E-04	4.16E-04	5.24E-04
AFFDP (MJ)	-3.81E+01	-3.00E+00	-1.40E+01	9.94E+01
GWP100 (kg CO2 eq.)	-9.44E-01	1.78E+00	8.46E-01	8.77E+00
ODP (kg CFC-11 eq.)	-1.83E-07	9.68E-08	7.69E-09	5.97E-07
HTP (kg 1,4-DB eq.)	6.66E-01	5.35E-01	8.19E-01	4.72E+00
FWAETP (kg 1,4-DB eq.)	8.76E-01	8.22E-01	5.05E-01	4.11E+00
MAETP (kg 1,4-DB eq.)	9.16E+02	2.46E+03	1.55E+03	1.13E+04
TEP (kg 1,4-DB eq.)	2.30E-03	1.30E-03	3.93E-03	2.36E-02
PCOP (kg C2H4 eq.)	-1.41E-03	-1.06E-03	-9.80E-04	2.89E-04
AP (kg SO2 eq.)	5.24E-03	1.98E-02	2.42E-02	3.93E-02
EP (kg PO4 eq.)	1.26E-02	2.41E-02	1.12E-02	-1.21E-01

Table S20 Life cycle indicator results of SA production through the optimal processing pathway (OPP) of fossil-based SA via maleic anhydride, glycerol, corn stover, glucose, and *S. japonica*.

Impact categories	Maleic anhydride-based SA	OPP of glycerol	OPP of corn Stover	OPP of glucose	OPP of *S. Japonica*
ADP (kg Sb eq.)	1.86E-05	4.84E-04	5.22E-04	4.35E-04	5.42E-04
AFFDP (MJ)	7.46E+01	3.65E+01	7.73E+01	6.06E+01	1.98E+02
GWP100 (kg CO2 eq.)	3.90E+00	2.95E+00	6.11E+00	4.74E+00	1.45E+01
ODP (kg CFC-11 eq.)	4.63E-07	2.80E-07	5.76E-07	4.71E-07	1.13E-06
HTP (kg 1,4-DB eq.)	1.54E+00	2.21E+00	3.47E+00	2.36E+00	1.21E+01
FWAETP (kg 1,4-DB eq.)	1.07E+00	1.94E+00	2.31E+00	1.57E+00	6.97E+00
MAETP (kg 1,4-DB eq.)	2.74E+03	3.66E+03	5.53E+03	4.29E+03	1.54E+04
TEP (kg 1,4-DB eq.)	3.92E-03	6.22E-03	7.62E-03	7.85E-03	3.77E-02
PCOP (kg C2H4 eq.)	2.30E-03	8.93E-04	1.46E-03	1.32E-03	3.52E-03
AP (kg SO2 eq.)	1.36E-02	1.88E-02	3.69E-02	3.78E-02	6.80E-02
EP (kg PO4 eq.)	4.10E-03	1.67E-02	2.88E-02	1.53E-02	-1.15E-01


A
B
C
Relative contribution (%)
100%
80%
60%
40%
20%
0%
-20%
-40%
-60%
-80%
ADP
AFFDP
GWP
ODP
HTP
FWAETP
MAETP
TEP
PCOP
AP
EP
4.65E-04
-3.81E+01
-9.44E-01
-1.83E-07
6.66E-01
8.76E-01
9.16E+02
2.30E-03
-1.41E-03
5.24E-03
1.26E-02
Succinic acid
Product transport
Biorefinery construction
SA purification
Fermentation
Feedstock
5.03E-04
-3.00E+00
1.78E+00
9.68E-08
5.35E-01
8.22E-01
2.46E+03
1.30E-03
-1.06E-03
1.98E-02
2.41E-02
DDGS
Succinic acid
Product transport
Biorefinery construction
SA purification
DDS purification
Fermentation
Pretreatment
Feedstock
4.16E-04
-1.40E+01
8.46E-01
7.69E-09
8.19E-01
5.05E-01
1.55E+03
3.93E-03
-9.80E-04
2.42E-02
1.12E-02
Succinic acid
Product transport
Biorefinery construction
SA purification
Fermentation
Feedstock

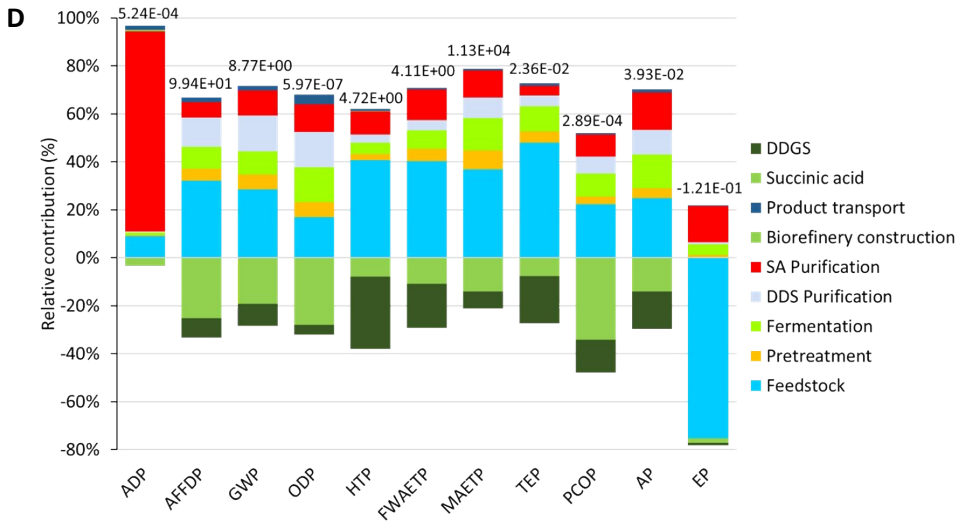


Figure S11 Life cycle profile of a bio-SA production through optimal processing pathway (OPP) of glycerol (A), corn stover (B), glucose (C), and *S. japonica* (D). Data at the top of each bar show net value for each environmental category. Abbreviations: ADP = Abiotic depletion potential; AFFDP = Fossil fuel depletion potential; GWP = Global warming potential 100 years; ODP = Ozone depletion potential; HTP = Human toxicity potential; FWAETP = Fresh water aquatic ecotoxicity potential; MAETP = Marine aquatic ecotoxicity potential; TEP = Terrestrial ecotoxicity potential; PCOP = Photo Chemical oxidation potential; AP = Acidification potential; EP = Eutrophication potential.

BioAmber Plant

BioAmber Sarina Plant was simulated to investigate bankruptcy causes. The process flow diagram shown in Figure S12 is based on patents released by BioAmber.[53–55] The feedstock is corn syrup, which is first heated to fermentation temperature and sent to the batch fermenter to yield succinic acid.[53,54] The fermentation operating conditions, yield, and concentration are given in Table S21. After fermentation, the broth is clarified using centrifugation or filtration to remove insoluble solids. According to the patent,[53] disc stack centrifuge, ultrafiltration, microfiltration, and depth filtration are promising technologies to remove insoluble solids. Thus, in the simulation, microfiltration and nanofiltration are assumed to remove insoluble solids and pigments. The clarified broth is pumped to a distillation column to remove water and ammonia, while distillate is cooled to saturation temperature (crystallization) to get a suspension of crystals, which is then filtered, washed, and dried.[53] It is important to mention that exact information regarding BioAmber's process is not publicly available, even not in the patents released by the company. Therefore, 25-50% error could be expected in the cost estimates accuracy. The assumptions in the techno-economic model are listed in Table S22. In reality, BioAmber signed several loan agreements to fund or secure the plant's construction.[56] It is not possible to consider all these loans in the model. Hence, we assume that fixed capital investment debt is 70%, with a loan period of 7 years and 8% loan interest.

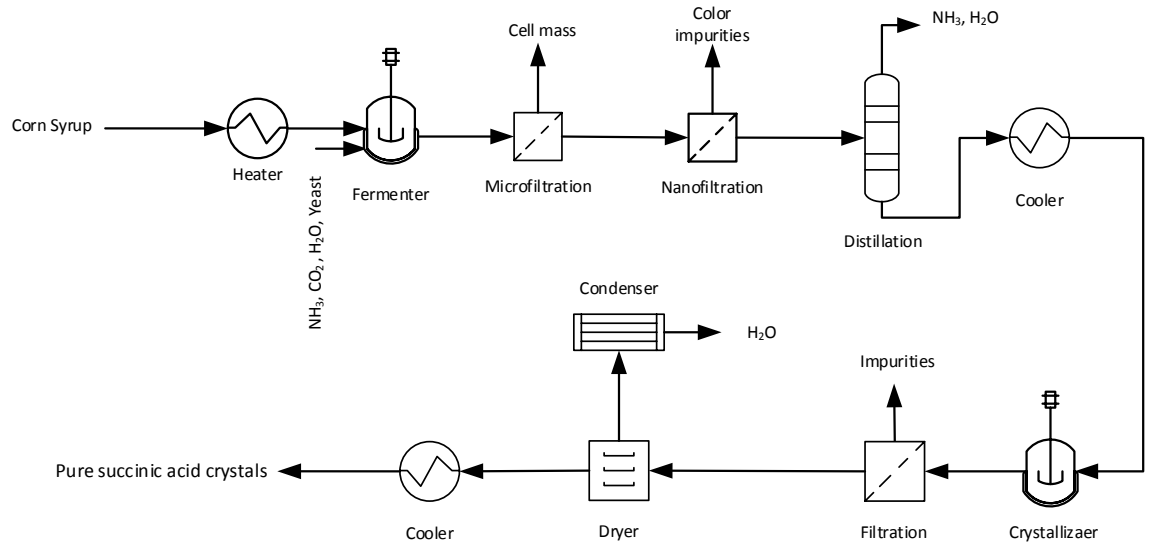


Figure S12 Process flow diagram of bio-SA production through BioAmber process.

Table S21 Fermentative conditions, yield, and titer values used to simulate the BioAmber process.[53,54]

Fermentation: operative conditions	**Yield (wt%)**	**Titer (g/**L)
Yeast (*C. krusei);* batch; O_2 and CO_2; 72h; 30°C; pH 3.0	0.45	48.2

Table S22 Techno-economic model assumptions to calculate the economics of simulated processes.

Assumptions	Values
Analysis year	2015
Production capacity	30 kt/y succinic acid
Project life	20 year
Discount rate	10%
Depreciation method	Straight-line
Depreciation years	7 year
Tax rate	30%
Construction time	2 years
Equity	30%
Loan to pay fixed capital investment	70%
Loan period	7 years
Loan interest	8%
Operating hours	8000
Salvage value	0

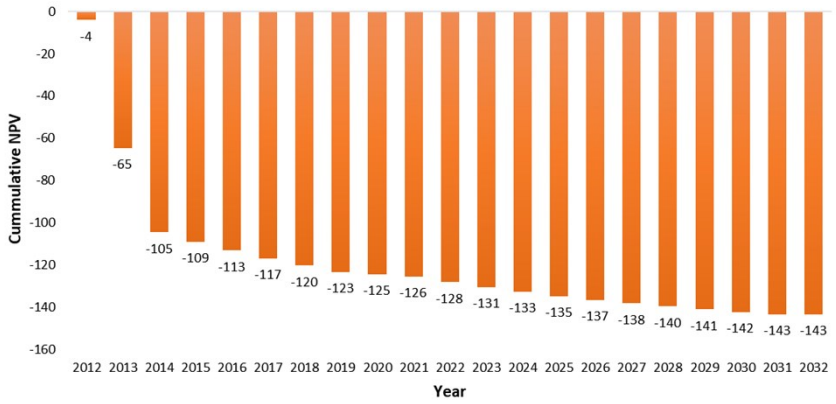


Figure S13 Cash flow vs. project life at a market selling price of 2.5 USD/kg.

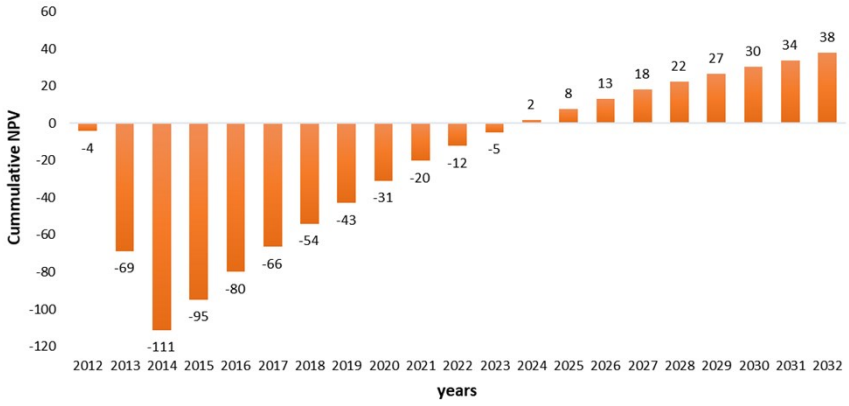


Figure S14 Cash flow vs. project life at a market selling price of 3.5 USD/kg.

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