Supplementary Information

We did a quantitatively techno-economic analysis (TEA) for HCOOH/formate production in MEA and flow cell systems, based on a modified model from previous reports, incorporating the performance obtained from our experimental results (*ACS Energy Lett*. 2026, 11, 1880; *Nat Catal*. 2025, 8, 968; *Nat Catal*. 2026, 9, 492).

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(1) TEA analysis of the CO_2RR-to-HCOOH process in MEA (200 mA cm^{-2})

The chemical equations for the formation of HCOOH are as follows:

$$\text{Cathode reaction: } 2CO_2+4e^-+4H^+ \rightarrow 2HCOOH$$
$$\text{Anode reaction: } 2H_2O-4e^- \rightarrow O_2+4H^+$$
$$\text{Overall reaction: } 2CO_2+2H_2O \rightarrow 2HCOOH+O_2$$

For HCOOH production, the production of 1 mole of HCOOH necessitates the consumption of 1 mole of CO_2 and 1 mole of H_2O. We assumed a cost analysis for the production of 100,000 kg of HCOOH *per* day with a capacity factor of 0.9 (32,850 tons *per* year=100000*0.9*365/1000), considering a factory lifespan of 20 years (*Nat Catal*. 2025, 8, 968; *Nature*. 2024, 626, 86).
For this estimation, we used experimental data from our previous publications (*Nat. Synth*. 2025, 4, 799). For HCOOH production in a MEA electrolyzer, the process was operating at a total current of 200 mA cm^{-2} with HCOOH FE of 90% with a cell voltage of 3 V. There is no need for catholyte, while 1 M KOH solution was used as the anolyte.

To illustrate the calculation details, we herein adopt the CO_2 electroreduction process for formate synthesis within a MEA as a representative case.

For HCOOH production, the partial current needed is:

$$I_{HCOOH}=100{,}000\frac{kg}{day}*\frac{day}{86{,}400\ s}*1{,}000\frac{g}{kg}*\frac{mol}{46.03}*2\ e^-*96{,}485\frac{C}{mol}=4{,}852{,}159.62\ A$$

The total current needed is the partial current divided by the faradaic efficiency:

$$I_{total}=\frac{4{,}852{,}159.62\ A}{0.9}=5{,}391{,}288.47\ A$$

The electrolyzer area needed is the total current divided by the current density:

$$Total\ electrolyzer\ area=\frac{5{,}391{,}288.47\ A}{0.2\ \frac{A}{cm^2}}*\frac{m^2}{10{,}000\ cm^2}=2695.64\ m^2$$

The power needed is given from P=VI:

$$Power=3\ V*5{,}391{,}288.47\ A*\frac{k\text{W}}{1{,}000\ W}=16{,}173.87\ kW$$

The CO_2 flow rate needed is:

$$CO_2\ flow=100{,}000\frac{kg}{day}*1{,}000\frac{g}{kg}*\frac{mol}{46.03}=2{,}172{,}496.20\ \frac{mol}{day}=95{,}589.83\ \frac{kg}{day}$$

According to 65% single pass conversion efficiency of CO_2, the inlet flow rate and outlet flow rate of CO_2 are:

$$CO_2\ inlet\ flow=2{,}172{,}496.20\frac{mol}{day}*\frac{day}{24\ \text{h}}*\frac{1}{0.65}=139{,}262.58\ \frac{mol}{\text{h}}=6{,}127.55\ \frac{kg}{\text{h}}$$

$$CO_2\ \text{outlet flow}=139{,}262.58\ \frac{mol}{\text{h}}*(1\text{-}0.65)=48{,}741.90\ \frac{mol}{\text{h}}=2{,}144.64\ \frac{kg}{\text{h}}$$

$$CO_2\ Outlet\ flow\ rate=\frac{CO_2\ outlet\ flow*R*T}{P}=\frac{48{,}741.90\ \frac{mol}{\mathrm{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=1{,}195.66\ \left(\frac{m^3}{\mathrm{h}}\right)$$

The byproduct is hydrogen (FE=10%), and the flow rate of hydrogen is:

$$H_2\ production\ rate=5{,}391{,}288.47*3{,}600\frac{s}{\mathrm{h}}*0.1*\frac{1}{2\ e^-*96{,}485\frac{C}{mol}}=10{,}057.85\ \frac{mol}{\mathrm{h}}$$

$$H_2\ flow\ rate=\frac{H_2\ production\ rate*R*T}{P}=\frac{10{,}057.85\ \frac{mol}{\mathrm{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=246.72\left(\frac{m^3}{\mathrm{h}}\right)$$

The water flow rate for the anodic OER reaction is:

$$Anodic\ H_2O\ flow\ rate=5{,}391{,}288.47\ A*3{,}600\frac{s}{\mathrm{h}}*\frac{1}{2\ e^-*96{,}485\frac{C}{mol}}*0.018\frac{kg}{mol}=1810.41\ \frac{kg}{\mathrm{h}}$$

It is assumed that the aqueous solution is recirculated until the total volume concentration of HCOOH reaches 10% before initiating liquid separation procedure (*ACS Energy Lett*. 2026, 11, 1880; *Nat Energy*. 2022, 7, 170; *Nat Commun.* 2025, 16, 9237).
The liquid HCOOH product flow rate is:

$$Product\ flow\ rate=\frac{Product\ rate\ (kg\frac{product}{day})*1000\frac{L}{m^3}}{Product\ density\ (\frac{kg}{m^3})*24\frac{\mathrm{h}}{day}}=\frac{100{,}000\ (\frac{kg}{day})*1000\frac{L}{m^3}}{1200\ (\frac{kg}{m^3})*24\frac{\mathrm{h}}{day}}=3{,}472.22\frac{L}{\mathrm{h}}$$

$$liquid\ product\ flow\ rate=\frac{Product\ rate\ (\frac{L}{\mathrm{h}})}{Product\ concentration}=\frac{3{,}472.22\frac{L}{\mathrm{h}}}{10\%}=34{,}722.20\frac{L}{\mathrm{h}}$$

1. Electolyzer cost

We referenced the analysis re-calculated based on the stack cost from DOE Current Central H2A PEM electrolysis. The reference electrolyzer cost for the stack component was 550 USD kW^{-1}, and it operated at 1.75 V and 0.4 A cm^{-2} (*Nat Commun*. 2026, 17, 2641; *Nature*. 2024, 626, 86; *Nat Sustain* 4. 2021, 911). In PEM systems, the precious Ir anode catalyst (2 mg cm^{-2}) dominates electrode cost; our electrolyzer is inferred to have a similar cost structure. To compare configurations with different power densities fairly, costs were recalculated per area (USD m^{-2}) rather than per power (USD kW^{-1}). Based on the DOE H2A PEM model, the electrolyzer cost *per* m^2 is calculated as follows:

$$Electrolyzer\ cost_{\text{per m}^2}=Reported\ cost\ \left(\frac{USD}{kW}\right)*\ Reported\ power\ density\ \left(\frac{kW}{m^2}\right)$$

$$=550\left(\frac{USD}{kW}\right)*\frac{0.4\ A}{cm^2}*1.75V*\frac{kW}{10^3\ W}*\frac{10^4\ cm^2}{m^2}=3{,}850.00\frac{USD}{m^2}$$

According to the factory operation capacity factor of 0.9, which represents the ratio of practical production to the maximum potential production capacity, the electrolyzer cost can be calculated as below (*Nat Catal*. 2025, 8, 968).

$$Electrolyzer\ cost=Total\ electrolyzer\ area*Electrolyzer\ cost_{per\ \text{m}^2}$$

$$=2695.64\ m^2*3{,}850.00\frac{USD}{m^2}=10{,}378{,}214.00\ USD$$

Assuming a capital discount rate (i) of 5%, the capital recovery factor (CRF) can be calculated as below:

$$CRF_{electrolyzer}=\frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1}=\frac{0.05*(1+0.05)^{20}}{(1+0.05)^{20}-1}=0.08$$

$$Electrolyzer\ cost_{per\ ton\ HCOOH}=10{,}378{,}214.00*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=25.27\frac{USD}{ton}$$

2. Balance of plant cost

According to the DOE H2A analysis for distributed grid electrolysis, the Balance of Plant (BOP) cost, which including both mechanical components BoP and electrical BoP, is 39% of the total cost, while the stack is 61% (*ACS Energy Lett*. 2026, 11, 1880). Thus, the total BoP capital cost is:

$$BoP\ cost=Electrolyzer\ cost\ (USD)*\frac{0.39}{0.61}=10{,}378{,}214.00\ USD*\frac{0.39}{0.61}=6{,}635{,}251.57\ \text{USD}$$

Herein, the BOP capital cost for production of *per* ton HCOOH can be calculated as below:

$$BoP\ cost_{per\ ton\ HCOOH}=6{,}635{,}251.57\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=16.16\ \frac{USD}{ton}$$

3. Electricity cost

We estimate the cost of electricity using the average industrial electricity price of US in 2024, which is 0.08 USD *per* kilowatt-hour as informed by data from the Department of Energy (DOE) (*Nat Sustain*. 2026, 9, 523). To calculate the electricity cost *per* ton of formic acid, we divided the total electricity cost by the daily output of production.

$$Cost_{electricity}=\frac{Power\times 24\ \frac{\text{h}}{day}\times price_{electricity}}{HCOOH\ production}=\frac{16{,}173.87\ kW\times 24\ \frac{\text{h}}{day}\times\frac{0.08\ USD}{kW\text{h}}}{100\left(\frac{ton}{day}\right)}=310.54\ \frac{USD}{ton}$$

4. Pressure-swing adsorption (PSA) cost

To enhance the recirculation of unreacted CO_2, byproduct H_2 is separated from the CO_2 outlet stream. To estimate the costs of gaseous separation, we refer to data from a reference pressure-swing adsorption (PSA) system (*Nat Catal*. 2025, 8, 968). The model uses a reference cost of \$1,989,043 for a flowrate capacity of 1000 $m^3\ h^{-1}$ with a scaling factor of 0.7 and an energy consumption of 0.25 kWh m^{-3} (*Nat Energy*. 2022, 7, 170).

$$PSA\ capital\ cost=reference\ cost*\left(\frac{total\ flow\ rate}{reference\ capacity}\right)^{scale\ factor}=1{,}989{,}043\ USD*\left(\frac{1{,}195.66\ \frac{m^3}{\text{h}}+346.72\ \frac{m^3}{\text{h}}}{1000\ \frac{m^3}{\text{h}}}\right)^{0.7}$$

$$=2{,}693{,}879.35\ USD$$

$$PSA\ capital\ cost_{per\ ton\ HCOOH}=2{,}693{,}879.35\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\ \frac{ton}{day}}=6.56\ \frac{USD}{ton}$$

$$PSA\ operating\ cost=\frac{total\ flow\ rate*\ reference\ PSA\ cost*electricity\ price}{target\ output\ production}=$$

$$\frac{\left(1{,}195.66\ \frac{m^3}{\text{h}}+346.72\ \frac{m^3}{\text{h}}\right)*24\ \frac{\text{h}}{day}*0.25\ \frac{kWh}{m^3}*0.08\ \frac{USD}{kWh}}{100\ \frac{ton}{day}}=7.40\ \frac{USD}{ton}$$

$$PSA\ total\ cost=PSA\ capital\ cost+PSA\ operating\ cost=13.96\ \frac{USD}{ton}$$

5. Liquid product separation cost

The distillation costs are calculated by scaling the reference costs to the flow rates. (*Nature*. 2023, 617, 724; *Nat Catal*. 2026, 9, 492)

For HCOOH:

$$Distillation\ capital\ cost=Distillation\ ref\ cost*\left(\frac{Product\ flow\ rate\ \left(\frac{L}{\text{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{scale\ factor}=$$

$$4{,}162{,}240\ USD*\left(\frac{34{,}722.20\ \frac{L}{\text{h}}}{1{,}000\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{0.7}=2{,}838{,}221.64\ USD$$

$$Distillation\ capital\ cost_{per\ ton\ HCOOH}=2{,}838{,}221.64\ USD*\frac{0.080}{365}*\frac{1}{100\ \frac{ton}{day}}=6.22\ \frac{USD}{ton}$$

$$Distillation\ operating\ cost=\ ref\ operating\ cost*\left(\frac{Product\ flow\ rate\ \left(\frac{L}{\text{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{scale\ factor}=$$

$$32037.71\ \frac{USD}{day}*\frac{1}{100\frac{ton}{day}}*\left(\frac{34{,}722.20\ \frac{L}{\text{h}}}{1{,}000\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{0.7}=218.46\ \frac{USD}{ton}$$

$$Distillation\ Cost_{per\ ton\ HCOOH}=6.22\ \frac{USD}{ton}+218.46\ \frac{USD}{ton}=224.68\ \frac{USD}{ton}$$

6. Catalyst and membrane cost

As mentioned above, the Ir catalyst used for the anodic water oxidation reaction (WOR) constitutes a significant portion of the electrolyzer cost. It is assumed that the costs of the catalyst and membrane account for 50% of the total electrolyzer cost, and that the catalyst and membrane have a lifetime of 5 years. (*Nature*. 2024, 626, 86; *Nat Catal*. 2026, 9, 492).

$$CRF_{C\&M}=\frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1}=\frac{0.05(1.05)^5}{(1.05)^5-1}=0.23$$

$$Cost_{C\&M}=\frac{CRF_{C\&M}*\ Cost_{Total\ Electrolyzer}\ (USD)*50\%}{Capacity\ factor*365\left(\frac{days}{year}\right)*production\left(\frac{ton}{day}\right)}=\frac{0.23*10{,}378{,}214.00\ USD*50\%}{0.9*365\left(\frac{days}{year}\right)*100\left(\frac{ton}{day}\right)}=36.33\ \frac{USD}{ton}$$

7. Input chemical cost

The cost of input CO_2 can be calculated by multiply CO_2 consumption by a market price for CO_2 (40 USD ton^{-1}) (*ACS Energy Lett*. 2026, 11, 1880; *Nature*. 2024, 626, 86).

$$Input\ CO_2\ cost=40\frac{USD}{ton\ CO_2}*95{,}589.83\ \frac{kg\ CO_2}{day}*\frac{ton}{1{,}000\ kg}*\frac{1}{100\frac{ton}{day}}=38.24\ \frac{USD}{ton}$$

Water consumption can be approximated by the anodic H_2O flow rate, given that the water from the cathodic liquid product flow rate is recyclable. Therefore, the cost of input H_2O can be calculated by multiply water consumption by a market price for water (5 USD ton^{-1}) (*Nat Catal*. 2025, 8, 968).

$$Input\ H_2O\ cost=5\frac{USD}{ton\ H_2O}*1810.41\ \frac{kg\ H_2O}{\text{h}}*\frac{ton}{1{,}000\ kg}*24\frac{\text{h}}{day}*\frac{1}{100\frac{ton}{day}}=2.17\ \frac{USD}{ton}$$

For HCOOH production, the electrolyte is 0.5 mol L^{-1} KOH, which can be recycled. Assuming electrolyte lifetime is one year, it can be calculated that the CRF is 1.07 (*Nature*. 2024, 626, 86). By using a fixed volume ratio of 100 L of electrolyte *per* square meter of electrolyzer for WOR, we can find the total volume of electrolyte needed (*Nat Catal*. 2025, 8, 968; *Nature*. 2024, 626, 86). The price of KOH is based on the data reported in this literature (*Angew. Chem. Int. Ed.* 2025, 64, e202504782)

$$Volume_{\text{electrolyte}}=electrolyzer\ area\ (m^2)*100\ \left(\frac{L}{m^2}\right)=2695.64\ m^2*100\frac{L}{m^2}=269{,}564.00\ L$$

$$Mass_{KOH}=molarity_{KOH}\left(\frac{mol}{L}\right)*Volume_{electrolyte}\ (L)\times molecular\ weight\ \left(\frac{g}{mol}\right)=0.5\frac{mol}{L}*269{,}564.00\ L*56.11\frac{g}{mol}$$

$$=7.56\ \text{ton}$$

$$KOH\ cost_{per\ ton\ HCOOH}=\frac{CRF_{electrolyzer}*Mass_{KOH}*KOH\ price}{Capacity\ factor*365\left(\frac{days}{year}\right)*production\left(\frac{ton}{day}\right)}=\frac{1.07*7.56\ ton*970\frac{USD}{ton}}{0.9*365\left(\frac{days}{year}\right)*100\left(\frac{ton}{day}\right)}=0.24\ \frac{USD}{ton}$$

8. Other operational cost

Other operational cost, including pre-treat, labor and maintenance et al., are presumed to constitute 2% of the electricity's cost (*Nat Catal*. 2025, 8, 968; *Nat. Synth*. 2026, 5, 885). Other operating costs are set differently for the flow cell and the MEA, with the flow cell accounting for the increased operating costs caused by frequent membrane replacement. (*Nature*. 2024, 626, 86; *Nat Catal*. 2026, 9, 492)

$$Other\ cost_{per\ ton\ HCOOH}=310.54\ \frac{USD}{ton}*0.02=6.21\ \frac{USD}{ton}$$

9. Total cost for one ton of Product:

By summing all the above costs, the cost for producing one ton of formic acid can be obtained:

$$Cost_{HCOOH}=Cost_{electrolyzer}+Cost_{BoP}+Cost_{electricity}+Cost_{PSA}+Cost_{Liquid\ product\ separation}+Cost_{catalyst\ and\ membrane}+Cost_{CO_2}$$

$$+Cost_{water}+ Cost_{Electrolyte\ KOH}+Cost_{other}$$

$$=25.27+16.16+310.54+13.96+224.68+36.33+38.24+2.17+0.24+6.21=673.80\ \frac{USD}{ton}$$

10. Profit:

The price of HCOOH is based on the data reported in this literature (*Nat Catal*. 2026, 9, 492).

Profit=HCOOH $_{per\ ton\ HCOOH}$-$Cost_{HCOOH}$

$=740\ \frac{USD}{ton}$ - $673.80\frac{USD}{ton}$= $66.20\ \frac{USD}{ton}$

(2) TEA analysis of the CO_2RR-to-HCOOH process in flow cell (200 mA cm^{-2})

For HCOOH production in a flow cell, the process was operating at a total current of 200 mA cm^{-2} with HCOOH FE of 95.6% with a cell voltage of 4.95 V (SPCE=70%) (*ACS Energy Lett.* 2026, 11, 1880-1888). There is no need for catholyte, while 1 M KOH solution was used as the anolyte.

For HCOOH production, the partial current needed is:

$$I_{HCOOH}=100{,}000\frac{kg}{day}*\frac{day}{86{,}400\ s}*1{,}000\frac{g}{kg}*\frac{mol}{46.03}*2\ e^{-}*96{,}485\frac{C}{mol}=4{,}852{,}159.62\ A$$

The total current needed is the partial current divided by the faradaic efficiency:

$$I_{total}=\frac{4{,}852{,}159.62\ A}{0.956}=5{,}075{,}480.77\ A$$

The electrolyzer area needed is the total current divided by the current density:

$$Total\ electrolyzer\ area=\frac{5{,}075{,}480.77\ A}{0.2\ \frac{A}{cm^2}}*\frac{m^2}{10{,}000\ cm^2}=2{,}537.74\ m^2$$

The power needed is given from P=VI:

$$Power=4.95\ V*5{,}075{,}480.77\ A*\frac{kW}{1{,}000\ W}=25{,}123.63\ kW$$

The CO_2 flow rate needed is:

$$CO_2\ flow=100{,}000\frac{kg}{day}*1{,}000\frac{g}{kg}*\frac{mol}{46.03}=2{,}172{,}496.20\ \frac{mol}{day}=95{,}589.83\ \frac{kg}{day}$$

According to 70% single pass conversion efficiency of CO_2, the inlet flow rate and outlet flow rate of CO_2 are:

$$CO_2\ inlet\ flow=2{,}172{,}496.20\frac{mol}{day}*\frac{day}{24\ \text{h}}*\frac{1}{0.70}=129{,}315.25\ \frac{mol}{\text{h}}=5{,}689.87\ \frac{kg}{\text{h}}$$

$$CO_2\ \text{outlet flow}=129{,}315.25\ \frac{mol}{\text{h}}*(1-0.70)=38{,}794.58\ \frac{mol}{\text{h}}=1{,}706.96\ \frac{kg}{\text{h}}$$

$$CO_2\ Outlet\ flow\ rate=\frac{CO_2\ outlet\ flow*R*T}{P}=\frac{38{,}794.58\ \frac{mol}{\text{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=951.65\ \left(\frac{m^3}{\text{h}}\right)$$

The byproduct is hydrogen (FE=4.4%), and the flow rate of hydrogen is:

$$H_2\ production\ rate=5{,}075{,}480.77*3{,}600\frac{s}{\text{h}}*0.044*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}=4{,}166.22\ \frac{mol}{\text{h}}$$

$$H_2\ flow\ rate=\frac{H_2\ production\ rate*R*T}{P}=\frac{4{,}166.22\ \frac{mol}{\text{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=102.20\left(\frac{m^3}{\text{h}}\right)$$

The water flow rate for the anodic OER reaction is:

$$Anodic\ H_2O\ flow\ rate=5{,}075{,}480.77\ A*3{,}600\frac{s}{\text{h}}*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}*0.018\frac{kg}{mol}=1704.36\ \frac{kg}{\text{h}}$$

The aqueous solution is recirculated until the total volume concentration of HCOOH reaches 10% before initiating liquid separation procedure.

The liquid HCOOH product flow rate is:

$$Product\ flow\ rate=\frac{Product\ rate\ (kg\frac{product}{day})*1000\frac{L}{m^3}}{Product\ density\ (\frac{kg}{m^3})*24\frac{h}{day}}=\frac{100{,}000\ (\frac{kg}{day})*1000\frac{L}{m^3}}{1200\ (\frac{kg}{m^3})*24\frac{h}{day}}=3{,}472.22\frac{L}{h}$$

$$\text{liquid product flow rate}=\frac{\text{Product rate }(\frac{L}{h})}{\text{Product concentration}}=\frac{3{,}472.22\frac{L}{h}}{10\%}=34{,}722.20\frac{L}{h}$$

1. Electolyzer cost

$$Electrolyzer\ cost_{\text{per m}^2}=Reported\ cost\ \left(\frac{USD}{kW}\right)*\ Reported\ power\ density\ \left(\frac{kW}{m^2}\right)$$

$$=550\left(\frac{USD}{kW}\right)*\frac{0.4\ A}{cm^2}*1.75V*\frac{kW}{10^3\ W}*\frac{10^4\ cm^2}{m^2}=3{,}850.00\frac{USD}{m^2}$$

$$Electrolyzer\ cost=Total\ electrolyzer\ area*Electrolyzer\ cost_{\text{per m}^2}$$

$$=2{,}537.74\ m^2*3{,}850.00\frac{USD}{m^2}=9{,}770{,}299.00\ USD$$

$$CRF_{electrolyzer}=\frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1}=\frac{0.05*(1+0.05)^{20}}{(1+0.05)^{20}-1}=0.08$$

$$Electrolyzer\ cost_{per\ ton\ HCOOH}=9{,}770{,}299.00*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=23.79\frac{USD}{ton}$$

2. Balance of plant cost

$$BoP\ cost=Electrolyzer\ cost\ (USD)*\frac{0.39}{0.61}=9{,}770{,}299.00\ \ USD*\frac{0.39}{0.61}=6{,}246{,}584.61\ \text{USD}$$

$$BoP\ cost_{per\ ton\ HCOOH}=6{,}246{,}584.61\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=15.21\ \frac{USD}{ton}$$

3. Electricity cost

$$Cost_{electricity}=\frac{Power\times 24\ \frac{h}{day}\times price_{electricity}}{HCOOH\ production}=\frac{25{,}123.63\ \ kW\times 24\ \frac{h}{day}\times\frac{0.08\ USD}{kW h}}{100\left(\frac{ton}{day}\right)}=482.37\ \frac{USD}{ton}$$

4. Pressure-swing adsorption (PSA) cost

$$PSA\ capital\ cost=reference\ cost*\left(\frac{total\ flow\ rate}{reference\ capacity}\right)^{scale\ factor}=$$

$$1{,}989{,}043\ USD*\left(\frac{951.65\ \frac{m^3}{h}+102.20\ \frac{m^3}{h}}{1000\ \frac{m^3}{h}}\right)^{0.7}=2{,}063{,}359.53\ USD$$

$$PSA\ capital\ cost_{per\ ton\ HCOOH}=2{,}063{,}359.53\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\ \frac{ton}{day}}=5.02\ \frac{USD}{ton}$$

$$PSA\ operating\ cost=\frac{total\ flow\ rate*\ reference\ PSA\ cost*electricity\ price}{target\ output\ production}=$$

$$\frac{\left(951.65\ \frac{m^3}{\text{h}}+102.20\ \frac{m^3}{\text{h}}\right)*24\ \frac{\text{h}}{day}*0.25\ \frac{kW\text{h}}{m^3}*0.08\ \frac{USD}{kW\text{h}}}{100\ \frac{ton}{day}}=5.06\ \frac{USD}{ton}$$

$$PSA\ total\ cost=PSA\ capital\ cost+PSA\ operating\ cost=10.08\ \frac{USD}{ton}$$

5. Liquid product separation cost

$$Distillation\ capital\ cost=Distillation\ ref\ cost*\left(\frac{Product\ flow\ rate\ \left(\frac{L}{\text{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{scale\ factor}=$$

$$4{,}162{,}240\ USD*\left(\frac{34{,}722.20\ \frac{L}{\text{h}}}{1{,}000\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{0.7}=2{,}838{,}221.64\ USD$$

$$Distillation\ capital\ cost_{per\ ton\ HCOOH}=2{,}838{,}221.64\ USD*\frac{0.080}{365}*\frac{1}{100\ \frac{ton}{day}}=6.22\ \frac{USD}{ton}$$

$$Distillation\ operating\ cost=\ ref\ operating\ cost*\left(\frac{Product\ flow\ rate\ \left(\frac{L}{\text{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{scale\ factor}=$$

$$32037.71\ \frac{USD}{day}*\frac{1}{100\frac{ton}{day}}*\left(\frac{34{,}722.20\ \frac{L}{\text{h}}}{1{,}000\left(\frac{L}{min}\right)*60\frac{min}{\text{h}}}\right)^{0.7}=218.46\ \frac{USD}{ton}$$

$$Distillation\ Cost_{per\ ton\ HCOOH}=6.22\ \frac{USD}{ton}+218.46\ \frac{USD}{ton}=224.68\ \frac{USD}{ton}$$

6. Catalyst and membrane cost

$$CRF_{C\&M}=\frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1}=\frac{0.05(1.05)^5}{(1.05)^5-1}=0.23$$

$$Cost_{C\&M}=\frac{CRF_{C\&M}*\ Cost_{Total\ Electrolyzer}\ (USD)*50\%}{Capacity\ factor*365\left(\frac{days}{year}\right)*production\left(\frac{ton}{day}\right)}=\frac{0.23*9{,}770{,}299.00\ USD*50\%}{0.9*365\left(\frac{days}{year}\right)*100\left(\frac{ton}{day}\right)}=34.20\ \frac{USD}{ton}$$

7. Input chemical cost

$$Input\ CO_2\ cost=40\frac{USD}{ton\ CO_2}*95{,}589.83\ \frac{kg\ CO_2}{day}*\frac{ton}{1{,}000\ kg}*\frac{1}{100\frac{ton}{day}}=38.24\ \frac{USD}{ton}$$

$$Input\ H_2O\ cost = 5\frac{USD}{ton\ H_2O} * 1704.36\ \frac{kg\ H_2O}{\text{h}} * \frac{ton}{1{,}000\ kg} * 24\frac{\text{h}}{day} * \frac{1}{100\frac{ton}{day}} = 2.05\ \frac{USD}{ton}$$

For HCOOH production, the electrolyte is 0.5 mol L^{-1} KOH, which can be recycled. Assuming electrolyte lifetime is one year, it can be calculated that the CRF is 1.07. By using a fixed volume ratio of 200 L of electrolyte *per* square meter of electrolyzer for WOR, we can find the total volume of electrolyte needed.

$$Volume_{\text{electrolyte}} = electrolyzer\ area\ (m^2) * 200\ \left(\frac{L}{m^2}\right) = 2{,}537.74\ m^2 * 200\frac{L}{m^2} = 507{,}548.00\ L$$

$$Mass_{KOH} = molarity_{KOH}\left(\frac{mol}{L}\right) * Volume_{electrolyte}\ (L) \times molecular\ weight\ \left(\frac{g}{mol}\right) = 0.5\frac{mol}{L} * 507{,}548.00\ L * 56.11\frac{g}{mol}$$

$$= 14.24\ \text{ton}$$

$$KOH\ cost_{per\ ton\ HCOOH} = \frac{CRF_{electrolyzer} * Mass_{KOH} * KOH\ price}{Capacity\ factor * 365\left(\frac{days}{year}\right) * production\left(\frac{ton}{day}\right)} = \frac{1.07 * 14.24\ ton * 970\frac{USD}{ton}}{0.9 * 365\left(\frac{days}{year}\right) * 100\left(\frac{ton}{day}\right)} = 0.45\ \frac{USD}{ton}$$

8. Other operational cost

$$Other\ cost_{per\ ton\ HCOOH} = 482.37\frac{USD}{ton} * 0.085 = 41.00\ \frac{USD}{ton}$$

9. Total cost for one ton of Product:

$$Cost_{HCOOH} = Cost_{electrolyzer} + Cost_{BoP} + Cost_{electricity} + Cost_{PSA} + Cost_{Liquid\ product\ separation} + Cost_{catalyst\ and\ membrane} + Cost_{CO_2} + Cost_{water} + Cost_{Electrolyte\ KOH} + Cost_{other}$$

$$= 23.79 + 15.21 + 482.37 + 10.08 + 224.68 + 34.20 + 38.24 + 2.05 + 0.45 + 41.00 = 872.07\ \frac{USD}{ton}$$

10. Profit:

$$Profit = HCOOH_{per\ ton\ HCOOH} - Cost_{HCOOH}$$

$$= 740\ \frac{USD}{ton} - 872.07\ \frac{USD}{ton} = -132.07\frac{USD}{ton}$$

(3) TEA analysis of the CO_2RR-to-HCOOCs process in MEA (300 mA cm^{-2})

For HCOOCs production, we used experimentally derived data from our long-term experiments to approximate the operating conditions of a potential industrial electrolyzer. For HCOOCs production in a MEA electrolyzer, the process was operating at a total current of 300 mA cm^{-2} with FE of 92.0% with a cell voltage of 3 V. There is no need for catholyte, while 20 mM Cs_2CO_3 solution was used as the anolyte.

For HCOOCs production, the partial current needed is:

$$I_{HCOOCs}=100{,}000\frac{kg}{day}*\frac{day}{86{,}400\ s}*1{,}000\frac{g}{kg}*\frac{mol}{177.92}*2\ e^{-}*96{,}485\frac{C}{mol}=1{,}255{,}310.86\ A$$

The total current needed is the partial current divided by the faradaic efficiency:

$$I_{total}=\frac{1{,}255{,}310.86\ A}{0.92}=1{,}364{,}468.33\ A$$

The electrolyzer area needed is the total current divided by the current density:

$$Total\ electrolyzer\ area=\frac{1{,}364{,}468.33\ A}{0.3\ \frac{A}{cm^2}}*\frac{m^2}{10{,}000\ cm^2}=454.82\ m^2$$

The power needed is given from P=VI:

$$Power=3\ V*1{,}364{,}468.33\ A*\frac{k\mathrm{W}}{1{,}000\ W}=4{,}093.40\ kW$$

The CO_2 flow rate needed is:

$$CO_2\ flow=100{,}000\frac{kg}{day}*1{,}000\frac{g}{kg}*\frac{mol}{177.92}=562{,}050.36\ \frac{mol}{day}=24{,}730.22\ \frac{kg}{day}$$

According to 65% single pass conversion efficiency of CO_2, the inlet flow rate and outlet flow rate of CO_2 are:

$$CO_2\ inlet\ flow=562{,}050.36\frac{mol}{day}*\frac{day}{24\ \mathrm{h}}*\frac{1}{0.65}=36{,}028.87\ \frac{mol}{\mathrm{h}}=1{,}585.27\ \frac{kg}{\mathrm{h}}$$

$$CO_2\ \text{outlet flow}=36{,}029.87\ \frac{mol}{\mathrm{h}}*(1-0.65)=12{,}610.10\ \frac{mol}{\mathrm{h}}=554.84\ \frac{kg}{\mathrm{h}}$$

$$CO_2\ Outlet\ flow\ rate=\frac{CO_2\ outlet\ flow*R*T}{P}=\frac{12{,}610.10\ \frac{mol}{\mathrm{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=309.33\ \left(\frac{m^3}{\mathrm{h}}\right)$$

The byproduct is hydrogen (FE=8%), and the flow rate of hydrogen is:

$$H_2\ production\ rate=1{,}364{,}468.33*3{,}600\frac{s}{\mathrm{h}}*0.08*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}=2{,}036.41\ \frac{mol}{\mathrm{h}}$$

$$H_2\ flow\ rate=\frac{H_2\ production\ rate*R*T}{P}=\frac{2{,}036.41\ \frac{mol}{\mathrm{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=49.95\left(\frac{m^3}{\mathrm{h}}\right)$$

The water flow rate for the anodic OER reaction is:

$$Anodic\ H_2O\ flow\ rate=1{,}364{,}468.33\ A*3{,}600\frac{s}{\mathrm{h}}*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}*0.018\frac{kg}{mol}=458.19\ \frac{kg}{\mathrm{h}}$$

The liquid HCOOCs product flow rate is:

$$liquid\ product\ flow\ rate = 100{,}000\frac{kg}{day} * \frac{day}{24\ \text{h}} * 1{,}000\frac{g}{kg} * \frac{mol}{177.92} * \frac{1}{0.02\ \frac{mol}{L}} = 1{,}170{,}938.25\ \frac{L}{\text{h}}$$

1. Electolyzer cost

$$Electrolyzer\ cost_{\text{per m}^2} = Reported\ cost\ \left(\frac{USD}{kW}\right) * Reported\ power\ density\ \left(\frac{kW}{m^2}\right)$$

$$= 550\left(\frac{USD}{kW}\right) * \frac{0.4\ A}{cm^2} * 1.75V * \frac{kW}{10^3\ W} * \frac{10^4\ cm^2}{m^2} = 3{,}850.00\frac{USD}{m^2}$$

$$Electrolyzer\ cost = Total\ electrolyzer\ area * Electrolyzer\ cost_{\text{per m}^2}$$

$$= 454.82\ m^2 * 3{,}850.00\frac{USD}{m^2} = 1{,}751{,}057.00\ USD$$

$$CRF_{electrolyzer} = \frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1} = \frac{0.05*(1+0.05)^{20}}{(1+0.05)^{20}-1} = 0.08$$

$$Electrolyzer\ cost_{per\ ton\ HCOOCs} = 1{,}751{,}057.00 * \frac{0.08}{365*0.9} * \frac{1}{100\left(\frac{ton\ \text{HCOOCs}}{day}\right)} = 4.26\frac{USD}{ton}$$

2. Balance of plant cost

$$BoP\ cost = Electrolyzer\ cost\ (USD) * \frac{0.39}{0.61} = 1{,}751{,}057.00\ USD * \frac{0.39}{0.61} = 1{,}119{,}528.25\ \text{USD}$$

$$BoP\ cost_{per\ ton\ HCOOCs} = 1{,}119{,}528.25\ USD * \frac{0.08}{365*0.9} * \frac{1}{100\left(\frac{ton\ \text{HCOOCs}}{day}\right)} = 2.73\ \frac{USD}{ton}$$

3. Electricity cost

$$Cost_{electricity} = \frac{Power \times 24\ \frac{\text{h}}{day} \times price_{electricity}}{HCOOCs\ production} = \frac{4{,}093.40\ kW \times 24\ \frac{\text{h}}{day} \times \frac{0.08\ USD}{kW\text{h}}}{100\left(\frac{ton\ \text{HCOOCs}}{day}\right)} = 78.59\ \frac{USD}{ton}$$

4. Pressure-swing adsorption (PSA) cost

$$PSA\ capital\ cost = reference\ cost * \left(\frac{total\ flow\ rate}{reference\ capacity}\right)^{scale\ factor} = 1{,}989{,}043\ USD * \left(\frac{309.33\ \frac{m^3}{\text{h}} + 49.95\ \frac{m^3}{\text{h}}}{1000\ \frac{m^3}{\text{h}}}\right)^{0.7}$$

$$= 971{,}510.06\ USD$$

$$PSA\ capital\ cost_{per\ ton\ HCOOCs} = 971{,}510.06\ USD * \frac{0.08}{365*0.9} * \frac{1{,}000\ \frac{kg}{ton}}{100{,}000\ \frac{kg}{day}} = 2.37\ \frac{USD}{ton}$$

$$PSA\ operating\ cost = \frac{total\ flow\ rate * reference\ PSA\ cost * electricity\ price}{target\ output\ production} =$$

$$\frac{\left(309.33\ \frac{m^3}{\text{h}} + 49.95\ \frac{m^3}{\text{h}}\right) * 24\ \frac{\text{h}}{day} * 0.25\ \frac{kW\text{h}}{m^3} * 0.08\ \frac{USD}{kW\text{h}}}{100\ \frac{ton\ HCOOCs}{day}} = 1.72\ \frac{USD}{ton}$$

$$PSA\ total\ cost = PSA\ capital\ cost + PSA\ operating\ cost = 4.09\ \frac{USD}{ton}$$

5. Liquid product separation cost

Membrane concentration coupled with vacuum fractional crystallization is a low-energy separation alternative to distillation for CO_2RR-derived cesium formate. The mixed electrolyte containing cesium formate and residual cesium carbonate first undergoes reverse osmosis to remove bulk water, followed by MVP vacuum crystallization. Selective separation of the two salts is achieved via their distinct solubility gaps under vacuum, yielding ≥98% pure solid cesium formate. Distillation is infeasible for this inorganic salt mixture due to the absence of usable vapor-liquid separation properties. Carbonate-rich mother liquor is fully recycled to the electrolyzer, cutting electrolyte loss and fitting streamlined techno-economic evaluation (*Chem. Eng. J.* 2025, 519, 165511; *Nat. Commun.* 2026, 17, 2099; *Nature Catalysis*. 2026, 9, 9).
RO cost:

The liquid HCOOCs product flow rate is:

$$liquid\ product\ flow\ rate = 100{,}000\ \frac{kg}{day} * \frac{day}{24\ \text{h}} * 1{,}000\ \frac{g}{kg} * \frac{mol}{177.92} * \frac{1}{0.02\ \frac{mol}{L}} = 1{,}170{,}938.25\ \frac{L}{\text{h}}$$

Based on the average flux for RO membrane of 50 L m^{-2} h^{-1} (Nature Catalysis 9 (2026) 9-17). Area of RO:

$$Area = \frac{1{,}170{,}938.25\ \frac{L}{\text{h}}}{50\ \text{L m}^{-2}\ \text{h}^{-1}} = 23418.765\ m^2$$

The cost of commercial RO modules was estimated at 380 USD m^{-2}, based on pricing data from DuPont Filmtec SW30-2521 (AUD 475 *per* unit with 0.7432 m^2 membrane area; ~428 USD m^{-2}) (*Nature Catalysis*. 2026, 9, 9).

$$RO\ capital\ cost = 23418.765\ m^2\ * 380 \frac{USD}{m^2} + \text{BOP} = 8{,}899{,}130.70 * 2\ \frac{USD}{m^2}$$

$$= 17{,}798{,}261.40\ \frac{USD}{m^2}$$

$$CRF_{RO\text{-}M} = \frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1} = \frac{0.05*(1+0.05)^5}{(1+0.05)^5-1} = 0.23$$

$$CRF_{RO\text{-}BOP} = \frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1} = \frac{0.05*(1+0.05)^{10}}{(1+0.05)^{10}-1} = 0.13$$

$$\text{RO}\ capital\ cost_{per\ ton\ HCOOCs} = 8{,}899{,}130.70\ USD * \frac{0.23}{365*0.9} * \frac{1}{100\left(\frac{ton}{day}\right)} + 8{,}899{,}130.70\ USD * \frac{0.13}{365*0.9} * \frac{1}{100\left(\frac{ton}{day}\right)}$$

$$= 62.31 + 35.21\ \frac{USD}{ton} = 97.52 \frac{USD}{ton}$$

According to the reference (*Nature Catalysis*. 2026, 9, 9), the electricity consumption for RO is 4.5 kWh to produce 1,000 kg of RO water.

The liquid HCOOCs product flow rate is:

$$liquid\ product\ flow\ rate = 100{,}000\frac{kg}{day} * \frac{day}{24\ \text{h}} * 1{,}000\frac{g}{kg} * \frac{mol}{177.92} * \frac{1}{0.02\ \frac{mol}{L}} = 1{,}170{,}938.25\ \frac{L}{\text{h}}$$

The feed flow rate of dilute cesium formate solution is 1,170,938.25 L/h, with an initial cesium formate concentration of 40 mM (0.04 mol/L), and the target concentration after RO concentration treatment is 1 mol/L. The total mass of solute remains constant throughout the concentration process in accordance with material balance.

Taking the approximate density of water as 1 kg/L for dilute aqueous solutions to convert volume to mass, the mass of water required to be evaporated/removed from the system *per* hour is determined to be:

$$M_{H_2O} = \rho_{H_2O} * V_{liquid\ product\ flow\ rate} \frac{C_{outlet} - C_{input}}{C_{outlet}} = 1\frac{kg}{L} * 1{,}170{,}938.25\frac{L}{\text{h}} * \frac{1\frac{mol}{L} - 0.04\frac{mol}{L}}{1\frac{mol}{L}} = 1{,}124{,}100.72\ \frac{kg}{h}$$

RO electricity = $\frac{liquid\ product\ flow\ rate}{1000}$ *Electricity $_{\text{RO module}}$ (kwh day^{-1})*Price

$$= 1{,}124{,}100.72\ \frac{kg}{h} * 4.5\ \frac{\text{kwh}}{1{,}000\ kg} * 0.08\frac{USD}{kW\text{h}} = 404.68\ \frac{USD}{\text{h}}$$

$$\text{RO electricity}_{per\ ton\ HCOOCs} = 404.68\ \frac{USD}{\text{h}} * 24\ \frac{h}{\text{day}} * \frac{1}{100\left(\frac{ton}{day}\right)} = 97.12\frac{USD}{ton}$$

According to the reference (*Nature Catalysis*. 2026, 9, 9), RO maintenance rate is 2% year^{-1}.

RO maintenance$_{per\ ton\ HCOOCs}$ = *RO capital cost* * RO maintenance rate

$$= 17{,}798{,}261.40\ \text{USD} * \frac{0.02}{365*0.9} * \frac{1}{100\left(\frac{ton}{day}\right)} = 10.84\ \text{USD/ton}$$

$$RO\ operating\ Cost_{per\ ton\ HCOOCs} = \text{RO electricity}_{per\ ton\ HCOOCs} + \text{RO maintenance}_{per\ ton\ HCOOCs}$$

$$= 97.12\ \frac{USD}{ton} + 10.84\ \frac{USD}{ton} = 107.96\frac{USD}{ton}$$

$$RO\ Cost_{per\ ton\ HCOOCs} = \text{RO}\ capital\ cost_{per\ ton\ HCOOCs} + RO\ operating\ Cost_{per\ ton\ HCOOCs}$$

$$= 97.52\ \frac{USD}{ton} + 107.96\ \frac{USD}{ton} = 205.48\ \frac{USD}{ton}$$

Distillation cost:

According to the reference (*Chem. Eng. J.* 2025, 519, 165511), the capital cost of evaporative crystallization equipment is quantified via the simplified fitted linear expression derived from the equipment cost-capacity curve (*Chem. Eng. J.* 2025, 519, 165511): Y=64.8+59.5X. where Y denotes the equipment price in ten thousand Chinese Yuan (CNY), and X represents the water treatment capacity in ton *per* hour (t/h).

$$water\ treatment\ capacity_{Evaporative\ crystallization} = \frac{\rho * C_{input} * V}{1000 * C_{outlet}} = \frac{1\frac{kg}{L} * 0.04\ \frac{mol}{L} * 1{,}170{,}938.25\ \frac{L}{\text{h}}}{1000 * 1\ \frac{mol}{L}} = 46.84\ \frac{ton}{h}$$

Evaporative crystallization capital cost = 64.8+59.5*46.84=*2851.78CNY/6.8 USD =419.38 USD*

$Evaporative\ crystallization_{capital\ cost\text{-}BOP}$=419.38 USD *5%=20.97 USD

$Evaporative\ crystallization_{capital\ cost}$ = 419.38+20.97=440.35 USD

$$Evaporative\ crystallization\ cost_{per\ ton\ HCOOCs}=440.35\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=0.0011\ \frac{USD}{ton}$$

Evaporative crystallization Operating cost= 96.37 CNY/t=14.17 USD/t

$$Evaporative\ crystallization\ Cost_{per\ ton\ HCOOCs}=0.0011\ \frac{USD}{ton}+14.17\ \frac{USD}{ton}=14.17\ \frac{USD}{ton}$$

Liquid product separation cost = $RO\ Cost_{per\ ton\ HCOOCs}$*2 + $Evaporative\ crystallization\ Cost_{per\ ton\ HCOOCs}$

=205.48*2 $\frac{USD}{ton}$ + 14.17 $\frac{USD}{ton}$

=425.13 $\frac{USD}{ton}$

6. Catalyst and membrane cost

$$CRF_{C\&M}=\frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}\text{-}1}=\frac{0.05(1.05)^5}{(1.05)^5\text{-}1}=0.23$$

$$Cost_{C\&M}=\frac{CRF_{C\&M}*\ Cost_{Total\ Electrolyzer}\ (USD)*50\%}{Capacity\ factor*365\left(\frac{days}{year}\right)*production\left(\frac{ton}{day}\right)}=\frac{0.23*1{,}751{,}057.00\ USD*50\%}{0.9*365\left(\frac{days}{year}\right)*100\left(\frac{ton}{day}\right)}=6.13\ \frac{USD}{ton}$$

7. Input chemical cost

$$Input\ CO_2\ cost=40\frac{USD}{ton\ CO_2}*24{,}730.22\ \frac{kg\ CO_2}{day}*\frac{ton}{1{,}000\ kg}*\frac{1}{100\frac{ton}{day}}=9.89\ \frac{USD}{ton}$$

$$Input\ H_2O\ cost=5\frac{USD}{ton\ H_2O}*458.19\ \frac{kg\ H_2O}{\text{h}}*\frac{ton}{1{,}000\ kg}*24\frac{\text{h}}{day}*\frac{1}{100\frac{ton}{day}}=0.55\ \frac{USD}{ton}$$

For HCOOCs production, the electrolyte is 0.02 mol L^{-1} Cs_2CO_3. The market price of Cs_2CO_3 and HCOOCs are taken as 294,117.65 and 411,764.71 USD ton^{-1}, respectively, with price data from 1688.com. As mentioned above, the production of 1 mole of HCOOCs necessitates the consumption 1 mole of Cs^+, the electrolyte Cs_2CO_3 cost can be calculated as:

$$Mass_{Cs_2CO_3}=1\ ton\ HCOOCs*\frac{1{,}000{,}000\ g}{ton}*\frac{mol}{177.92\ g}*\frac{325.82\ g}{2\ mol\ Cs^+}*\frac{ton}{1{,}000{,}000\ g}=0.92\ ton$$

$$Cs_2CO_3\ cost_{per\ ton\ HCOOCs}=0.92\ ton*294{,}117.65\frac{USD}{ton}=270{,}588.24\ \frac{USD}{ton}$$

8. Other operational cost

$$Other\ cost_{per\ ton\ HCOOCs}=78.59\ \frac{USD}{ton}*0.02=1.57\ \frac{USD}{ton}$$

9. Total cost for one ton of Product:

$$Cost_{HCOOCs}=Cost_{electrolyzer}+Cost_{BoP}+Cost_{electricity}+Cost_{PSA}+Cost_{Liquid\ product\ separation}+Cost_{catalyst\ and\ membrane}+Cost_{CO_2}+Cost_{water}+Cost_{Electrolyte\ Cs_2CO_3}+Cost_{other}$$

$$=4.26+2.73+78.59+4.09+425.13+6.13+9.89+0.55+270{,}588.24+1.57=271{,}121.18\ \frac{USD}{ton}$$

10. Profit:

$Profit=HCOOCs_{\ per\ ton\ HCOOCs}\text{-}Cost_{HCOOCs}$

$=411{,}764.71\ \frac{USD}{ton}\ -\ 271{,}121.18\ \frac{USD}{ton}=140{,}643.53\ \frac{USD}{ton}$

(4) TEA analysis of the CO_2RR-to-HCOOK process in MEA (300 mAcm^{-2})

For HCOOK production, the partial current needed is:

$$I_{HCOOCs}=100{,}000\frac{kg}{day}*\frac{day}{86{,}400\ s}*1{,}000\frac{g}{kg}*\frac{mol}{84.12}*2\ e^{-}*96{,}485\frac{C}{mol}=2{,}655{,}074.98\ A$$

The total current needed is the partial current divided by the faradaic efficiency:

$$I_{total}=\frac{2{,}655{,}074.98A}{0.92}=2{,}885{,}951.07\ A$$

The electrolyzer area needed is the total current divided by the current density:

$$Total\ electrolyzer\ area=\frac{2{,}885{,}951.07\ A}{0.3\ \frac{A}{cm^2}}*\frac{m^2}{10{,}000\ cm^2}=961.98\ m^2$$

The power needed is given from P=VI:

$$Power=3\ V*2{,}885{,}951.07\ A*\frac{kW}{1{,}000\ W}=8{,}657.85\ kW$$

The CO_2 flow rate needed is:

$$CO_2\ flow=100{,}000\frac{kg}{day}*1{,}000\frac{g}{kg}*\frac{mol}{84.12}=1{,}188{,}777.94\ \frac{mol}{day}=52{,}306.23\ \frac{kg}{day}$$

According to 65% single pass conversion efficiency of CO_2, the inlet flow rate and outlet flow rate of CO_2 are:

$$CO_2\ inlet\ flow=1{,}188{,}777.94\frac{mol}{day}*\frac{day}{24\ \text{h}}*\frac{1}{0.65}=76{,}203.71\ \frac{mol}{\text{h}}=3{,}352.96\ \frac{kg}{\text{h}}$$

$$CO_2\ \text{outlet flow}=76{,}203.71\ \frac{mol}{\text{h}}*(1-0.65)=26{,}671.30\ \frac{mol}{\text{h}}=1{,}173.54\ \frac{kg}{\text{h}}$$

$$CO_2\ Outlet\ flow\ rate=\frac{CO_2\ outlet\ flow*R*T}{P}=\frac{26{,}671.30\ \frac{mol}{\text{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=654.26\ \left(\frac{m^3}{\text{h}}\right)$$

It is assumed the byproduct is hydrogen (FE=8%), the flow rate of hydrogen is:

$$H_2\ production\ rate=2{,}885{,}951.07\ A*3{,}600\frac{s}{\text{h}}*0.08*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}=4{,}307.17\ \frac{mol}{\text{h}}$$

$$H_2\ flow\ rate=\frac{H_2\ production\ rate*R*T}{P}=\frac{4{,}307.17\ \frac{mol}{\text{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=105.66\left(\frac{m^3}{\text{h}}\right)$$

The water flow rate for the anodic OER reaction is:

$$Anodic\ H_2O\ flow\ rate=2{,}885{,}951.07\ A*3{,}600\frac{s}{\text{h}}*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}*0.018\frac{kg}{mol}=969.11\ \frac{kg}{\text{h}}$$

The liquid HCOOK product flow rate is:

$$liquid\ product\ flow\ rate=100{,}000\frac{kg}{day}*\frac{day}{24\ \text{h}}*1{,}000\frac{g}{kg}*\frac{mol}{84.12}*\frac{1}{0.5\ \frac{mol}{L}}=99{,}064.83\ \frac{L}{\text{h}}$$

1. Electolyzer cost

$$Electrolyzer\ cost_{\text{per m}^2}=Reported\ cost\ \left(\frac{USD}{kW}\right)*\ Reported\ power\ density\ \left(\frac{kW}{m^2}\right)$$

$$=550\left(\frac{USD}{kW}\right)*\frac{0.4\ A}{cm^2}*1.75V*\frac{kW}{10^3\ W}*\frac{10^4\ cm^2}{m^2}=3{,}850.00\frac{USD}{m^2}$$

$$Electrolyzer\ cost=Total\ electrolyzer\ area*Electrolyzer\ cost_{\text{per m}^2}$$

$$=961.98\ m^2*3{,}850.00\frac{USD}{m^2}=3{,}703{,}623.00\ USD$$

$$CRF_{electrolyzer}=\frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1}=\frac{0.05*(1+0.05)^{20}}{(1+0.05)^{20}-1}=0.08$$

$$Electrolyzer\ cost_{per\ ton\ HCOOK}=3{,}703{,}623.00*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=9.02\frac{USD}{ton}$$

2. Balance of plant cost

$$BoP\ cost=Electrolyzer\ cost\ (USD)*\frac{0.39}{0.61}=3{,}703{,}623.00\ USD*\frac{0.39}{0.61}=2{,}367{,}912.86\ \text{USD}$$

$$BoP\ cost_{per\ ton\ HCOOK}=2{,}367{,}912.86\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=5.77\ \frac{USD}{ton}$$

3. Electricity cost

$$Cost_{electricity}=\frac{Power\times 24\ \frac{\text{h}}{day}\times price_{electricity}}{HCOOK\ production}=\frac{8{,}657.85\ kW\times 24\ \frac{\text{h}}{day}\times\frac{0.08\ USD}{kW\text{h}}}{100\left(\frac{ton}{day}\right)}=166.23\ \frac{USD}{ton}$$

4. Pressure-swing adsorption (PSA) cost

$$PSA\ capital\ cost=reference\ cost*\left(\frac{total\ flow\ rate}{reference\ capacity}\right)^{scale\ factor}=1{,}989{,}043\ USD*\left(\frac{654.26\ \frac{m^3}{\text{h}}+105.66\ \frac{m^3}{\text{h}}}{1000\ \frac{m^3}{\text{h}}}\right)^{0.7}$$

$$=1{,}641{,}276.25\ USD$$

$$PSA\ capital\ cost_{per\ ton\ HCOOK}=1{,}641{,}276.25\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\ \frac{ton}{day}}=4.00\ \frac{USD}{ton}$$

$$PSA\ operating\ cost=\frac{total\ flow\ rate*\ reference\ PSA\ cost*electricity\ price}{target\ output\ production}=$$

$$\frac{\left(654.26\ \frac{m^3}{\text{h}}+105.66\ \frac{m^3}{\text{h}}\right)*24\ \frac{\text{h}}{day}*0.25\ \frac{kW\text{h}}{m^3}*0.08\ \frac{USD}{kW\text{h}}}{100\ \frac{ton}{day}}=3.65\ \frac{USD}{ton}$$

$$PSA\ total\ cost=PSA\ capital\ cost+PSA\ operating\ cost=7.65\ \frac{USD}{ton}$$

5. Liquid product separation cost

$$Distillation\ capital\ cost = Distillation\ ref\ cost * \left(\frac{Product\ flow\ rate\ \left(\frac{L}{\text{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right) * 60\frac{min}{\text{h}}} \right)^{scale\ factor}$$

$$= 4{,}162{,}240\ USD * \left(\frac{99{,}064.83\ \frac{L}{\text{h}}}{1{,}000\left(\frac{L}{min}\right) * 60\frac{min}{\text{h}}} \right)^{0.7} = 5{,}941{,}432.93\ USD$$

$$Distillation\ capital\ cost_{per\ ton\ HCOOK} = 5{,}941{,}432.93\ USD * \frac{0.080}{365} * \frac{1}{100\ \frac{ton}{day}} = 14.47\ \frac{USD}{ton}$$

$$Distillation\ operating\ cost = ref\ operating\ cost * \left(\frac{Product\ flow\ rate\ \left(\frac{L}{\text{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right) * 60\frac{min}{\text{h}}} \right)^{scale\ factor}$$

$$= 32037.71\ \frac{USD}{day} * \frac{1}{100\frac{ton}{day}} * \left(\frac{99{,}064.83\ \frac{L}{\text{h}}}{1{,}000\left(\frac{L}{min}\right) * 60\frac{min}{\text{h}}} \right)^{0.7} = 457.33\ \frac{USD}{ton}$$

$$Distillation\ Cost_{per\ ton\ HCOOK} = 14.47\ \frac{USD}{ton} + 457.33\ \frac{USD}{ton} = 471.80\ \frac{USD}{ton}$$

6. Catalyst and membrane cost

$$CRF_{C\&M} = \frac{i(1+i)^{lifetime}}{(1+i)^{lifetime} - 1} = \frac{0.05(1.05)^5}{(1.05)^5 - 1} = 0.23$$

$$Cost_{C\&M} = \frac{CRF_{C\&M} * Cost_{Total\ Electrolyzer}\ (USD) * 50\%}{Capacity\ factor * 365\left(\frac{days}{year}\right) * production\left(\frac{ton}{day}\right)} = \frac{0.23 * 3{,}703{,}623.00\ USD * 50\%}{0.9 * 365\left(\frac{days}{year}\right) * 100\left(\frac{ton}{day}\right)} = 12.97\ \frac{USD}{ton}$$

7. Input chemical cost

$$Input\ CO_2\ cost = 40\frac{USD}{ton\ CO_2} * 52{,}306.23\ \frac{kg\ CO_2}{day} * \frac{ton}{1{,}000\ kg} * \frac{1}{100\frac{ton}{day}} = 20.92\ \frac{USD}{ton}$$

$$Input\ H_2O\ cost = 5\frac{USD}{ton\ H_2O} * 969.11\ \frac{kg\ H_2O}{\text{h}} * \frac{ton}{1{,}000\ kg} * 24\frac{\text{h}}{day} * \frac{1}{100\frac{ton}{day}} = 1.16\ \frac{USD}{ton}$$

For HCOOK production, the electrolyte is 0.5 mol L^{-1} KOH. As mentioned above, the production of 1 mole of HCOOK necessitates the consumption 1 mole of K^+, the electrolyte KOH cost can be calculated as:

$$Mass_{KOH} = 1\ ton\ HCOOK * \frac{1{,}000{,}000\ g}{ton} * \frac{mol}{84.12\ g} * \frac{56.11\ g}{1\ mol\ K^+} * \frac{ton}{1{,}000{,}000\ g} = 0.67\ ton$$

$$KOH\ cost_{per\ ton\ HCOOK} = 0.67\ ton * 970\frac{USD}{ton} = 649.90\ \frac{USD}{ton}$$

8. Other operational cost

$$Other\ cost_{per\ ton\ HCOOK}=166.23\ \frac{USD}{ton}*0.02=3.32\ \frac{USD}{ton}$$

9. Total cost for one ton of Product:

$$Cost_{HCOOK}= Cost_{electrolyzer}+Cost_{BoP}+Cost_{electricity}+Cost_{PSA}+Cost_{Liquid\ product\ separation}+Cost_{catalyst\ and\ membrane}+Cost_{CO_2}+Cost_{water}+ Cost_{Electrolyte\ KOH}+Cost_{other}$$

$$=9.02+5.77+166.23+7.65+471.80+12.97+20.92+1.16+649.90+3.32=1348.74\ \frac{USD}{ton}$$

10. Profit:

The price of HCOOK is based on the data reported in this literature (*Angew. Chem. Int. Ed.* 2025, 64, e202504782).

$$Profit=HCOOK_{per\ ton\ HCOOK} - Cost_{HCOOK}$$

$$=1100\ \frac{USD}{ton} - 1348.74\ \frac{USD}{ton}= -248.74\ \frac{USD}{ton}$$

(5) TEA analysis of the CO_2RR-to-HCOOH process in MEA (300 mAcm^{-2})

For HCOOH production, the partial current needed is:

$$I_{HCOOH}=100{,}000\frac{kg}{day}*\frac{day}{86{,}400\ s}*1{,}000\frac{g}{kg}*\frac{mol}{46.03}*2\ e^{-}*96{,}485\frac{C}{mol}=4{,}852{,}159.62\ A$$

The total current needed is the partial current divided by the faradaic efficiency:

$$I_{total}=\frac{4{,}852{,}159.62\ A}{0.92}=5{,}274{,}086.54\ A$$

The electrolyzer area needed is the total current divided by the current density:

$$Total\ electrolyzer\ area=\frac{5{,}274{,}086.54\ A}{0.3\ \frac{A}{cm^2}}*\frac{m^2}{10{,}000\ cm^2}=1{,}758.03\ \ m^2$$

The power needed is given from P=VI:

$$Power=3\ V*5{,}274{,}086.54\ A*\frac{k\mathrm{W}}{1{,}000\ W}=15{,}822.26\ kW$$

The CO_2 flow rate needed is:

$$CO_2\ flow=100{,}000\frac{kg}{day}*1{,}000\frac{g}{kg}*\frac{mol}{46.03}=2{,}172{,}496.20\ \frac{mol}{day}=95{,}589.83\ \frac{kg}{day}$$

According to 65% single pass conversion efficiency of CO_2, the inlet flow rate and outlet flow rate of CO_2 are:

$$CO_2\ inlet\ flow=2{,}172{,}496.20\frac{mol}{day}*\frac{day}{24\ \mathrm{h}}*\frac{1}{0.65}=139{,}262.58\ \frac{mol}{\mathrm{h}}=6{,}127.55\ \frac{kg}{\mathrm{h}}$$

$$CO_2\ \text{outlet flow}=139{,}262.58\ \frac{mol}{\mathrm{h}}*(1-0.65)=48{,}741.90\ \frac{mol}{\mathrm{h}}=2{,}144.64\ \frac{kg}{\mathrm{h}}$$

$$CO_2\ Outlet\ flow\ rate=\frac{CO_2\ outlet\ flow*R*T}{P}=\frac{48{,}741.90\ \frac{mol}{\mathrm{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=1{,}195.66\ \left(\frac{m^3}{\mathrm{h}}\right)$$

The byproduct is hydrogen (FE=8%), the flow rate of hydrogen is:

$$H_2\ production\ rate=5{,}274{,}086.54\ A*3{,}600\frac{s}{\mathrm{h}}*0.08*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}=7{,}871.36\ \frac{mol}{\mathrm{h}}$$

$$H_2\ flow\ rate=\frac{H_2\ production\ rate*R*T}{P}=\frac{7{,}871.365\ \frac{mol}{\mathrm{h}}*8.314\ \frac{J}{mol\ K}*298\ K}{101\ kPa}=193.09\left(\frac{m^3}{\mathrm{h}}\right)$$

The water flow rate for the anodic OER reaction is:

$$Anodic\ H_2O\ flow\ rate=5{,}274{,}086.54\ A*3{,}600\frac{s}{\mathrm{h}}*\frac{1}{2\ e^{-}*96{,}485\frac{C}{mol}}*0.018\frac{kg}{mol}=1{,}771.06\ \frac{kg}{\mathrm{h}}$$

The aqueous solution is recirculated until the total volume concentration of HCOOH reaches 10% before initiating liquid separation procedure.

The liquid HCOOH product flow rate is:

$$Product\ flow\ rate=\frac{Product\ rate\ (kg\frac{product}{day})*1000\frac{L}{m^3}}{Product\ density\ (\frac{kg}{m^3})*24\frac{\mathrm{h}}{day}}=\frac{100{,}000\ (\frac{kg}{day})*1000\frac{L}{m^3}}{1200\ (\frac{kg}{m^3})*24\frac{\mathrm{h}}{day}}=3{,}472.22\frac{L}{\mathrm{h}}$$

$$liquid\ product\ flow\ rate=\frac{Product\ rate\ (\frac{L}{h})}{Product\ concentration}=\frac{3{,}472.22\frac{L}{h}}{10\%}=34{,}722.20\frac{L}{h}$$

1. Electolyzer cost

$$Electrolyzer\ cost_{per\ m^2}=Reported\ cost\ \left(\frac{USD}{kW}\right)*\ Reported\ power\ density\ \left(\frac{kW}{m^2}\right)$$

$$=550\left(\frac{USD}{kW}\right)*\frac{0.4\ A}{cm^2}*1.75V*\frac{kW}{10^3\ W}*\frac{10^4\ cm^2}{m^2}=3{,}850.00\frac{USD}{m^2}$$

$$Electrolyzer\ cost=Total\ electrolyzer\ area*Electrolyzer\ cost_{per\ m^2}$$

$$=1{,}758.03\ m^2*3{,}850.00\frac{USD}{m^2}=6{,}768{,}415.5\ \ USD$$

$$CRF_{electrolyzer}=\frac{i(1+i)^{lifetime}}{(1+i)^{lifetime}-1}=\frac{0.05*(1+0.05)^{20}}{(1+0.05)^{20}-1}=0.08$$

$$Electrolyzer\ cost_{per\ ton\ HCOOH}=6{,}768{,}415.5*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=16.48\frac{USD}{ton}$$

2. Balance of plant cost

$$BoP\ cost=Electrolyzer\ cost\ (USD)*\frac{0.39}{0.61}=6{,}768{,}415.5\ USD*\frac{0.39}{0.61}=4{,}327{,}347.61\ \text{USD}$$

$$BoP\ cost_{per\ ton\ HCOOH}=4{,}327{,}347.61\ USD*\frac{0.08}{365*0.9}*\frac{1}{100\left(\frac{ton}{day}\right)}=10.54\ \frac{USD}{ton}$$

3. Electricity cost

$$Cost_{electricity}=\frac{Power\times 24\ \frac{h}{day}\times price_{electricity}}{HCOOH\ production}=\frac{15{,}822.26\ kW\times 24\ \frac{h}{day}\times\frac{0.08\ USD}{kWh}}{100\left(\frac{ton}{day}\right)}=303.79\ \frac{USD}{ton}$$

4. Pressure-swing adsorption (PSA) cost

$$PSA\ capital\ cost=reference\ cost*\left(\frac{total\ flow\ rate}{reference\ capacity}\right)^{scale\ factor}=1{,}989{,}043\ USD*\left(\frac{1{,}195.66\ \frac{m^3}{h}+193.09\ \frac{m^3}{h}}{1000\ \frac{m^3}{h}}\right)^{0.7}$$

$$=2{,}503{,}116.27\ USD$$

$$PSA\ capital\ cost_{per\ ton\ HCOOH}=2{,}503{,}116.27\ \ USD*\frac{0.08}{365*0.9}*\frac{1}{100\ \frac{ton}{day}}=6.10\ \frac{USD}{ton}$$

$$PSA\ operating\ cost=\frac{total\ flow\ rate*\ reference\ PSA\ cost*electricity\ price}{target\ output\ production}$$

$$=\frac{\left(1{,}195.66\ \frac{m^3}{h}+193.09\ \ \frac{m^3}{h}\right)*24\ \frac{h}{day}*0.25\ \frac{kWh}{m^3}*0.08\ \frac{USD}{kWh}}{100\ \frac{ton}{day}}=6.67\ \frac{USD}{ton}$$

$$PSA\ total\ cost = PSA\ capital\ cost + PSA\ operating\ cost = 12.77\ \frac{USD}{ton}$$

5. Liquid product separation cost

$$Distillation\ capital\ cost = Distillation\ ref\ cost * \left(\frac{Product\ flow\ rate\ \left(\frac{L}{\mathrm{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right) * 60\frac{min}{\mathrm{h}}} \right)^{scale\ factor}$$

$$= 4{,}162{,}240\ USD * \left(\frac{34{,}722.20\ \frac{L}{\mathrm{h}}}{1{,}000\left(\frac{L}{min}\right) * 60\frac{min}{\mathrm{h}}} \right)^{0.7} = 2{,}838{,}221.64\ USD$$

$$Distillation\ capital\ cost_{per\ ton\ HCOOH} = 2{,}838{,}221.64\ USD * \frac{0.080}{365} * \frac{1}{100\ \frac{ton}{day}} = 6.22\ \frac{USD}{ton}$$

$$Distillation\ operating\ cost = ref\ operating\ cost * \left(\frac{Product\ flow\ rate\ \left(\frac{L}{\mathrm{h}}\right)}{Distillation\ ref\ scale\left(\frac{L}{min}\right) * 60\frac{min}{\mathrm{h}}} \right)^{scale\ factor}$$

$$= 32037.71\ \frac{USD}{day} * \frac{1}{100\frac{ton}{day}} * \left(\frac{34{,}722.20\ \frac{L}{\mathrm{h}}}{1{,}000\left(\frac{L}{min}\right) * 60\frac{min}{\mathrm{h}}} \right)^{0.7} = 218.46\ \frac{USD}{ton}$$

$$Distillation\ Cost_{per\ ton\ HCOOH} = 6.22\ \frac{USD}{ton} + 218.46\ \frac{USD}{ton} = 224.68\ \frac{USD}{ton}$$

6. Catalyst and membrane cost

$$CRF_{C\&M} = \frac{i(1+i)^{lifetime}}{(1+i)^{lifetime} - 1} = \frac{0.05(1.05)^5}{(1.05)^5 - 1} = 0.23$$

$$Cost_{C\&M} = \frac{CRF_{C\&M} * Cost_{Total\ Electrolyzer}\ (USD) * 50\%}{Capacity\ factor * 365\left(\frac{days}{year}\right) * production\left(\frac{ton}{day}\right)} = \frac{0.23 * 6{,}768{,}415.5\ USD * 50\%}{0.9 * 365\left(\frac{days}{year}\right) * 100\left(\frac{ton}{day}\right)} = 23.69\ \frac{USD}{ton}$$

7. Input chemical cost

$$Input\ CO_2\ cost = 40\frac{USD}{ton\ CO_2} * 95{,}589.83\ \frac{kg\ CO_2}{day} * \frac{ton}{1{,}000\ kg} * \frac{1}{100\frac{ton}{day}} = 38.24\ \frac{USD}{ton}$$

$$Input\ H_2O\ cost = 5\frac{USD}{ton\ H_2O} * 1{,}771.06\ \frac{kg\ H_2O}{\mathrm{h}} * \frac{ton}{1{,}000\ kg} * 24\frac{\mathrm{h}}{day} * \frac{1}{100\frac{ton}{day}} = 2.13\ \frac{USD}{ton}$$

For HCOOH production, the electrolyte is 0.5 mol L^{-1} H_2SO_4, which can be recycled. Assuming electrolyte lifetime is one year, it can be calculated that the CRF is 1.07. By using a fixed volume ratio of 100 L of electrolyte *per* square meter of electrolyzer for WOR, we can find the total volume of electrolyte needed. The price of H_2SO_4 is based on the data reported in this literature (*Adv. Sci.* 2025, 12, e08152).

$$Volume_{\text{electrolyte}} = electrolyzer\ area\ (m^2) * 100\ \left(\frac{L}{m^2}\right) = 1{,}758.03\ m^2 * 100\frac{L}{m^2} = 175{,}803.00\ L$$

$$Mass_{H_2SO_4} = molarity_{H_2SO_4}\left(\frac{mol}{L}\right) * Volume_{electrolyte}\ (L) \times molecular\ weight\ \left(\frac{g}{mol}\right) = 0.5\frac{mol}{L} * 175{,}803.00\ L * 98.08\frac{g}{mol}$$

$$= 8.62 \text{ ton}$$

$$H_2SO_4\ cost_{per\ ton\ HCOOH} = \frac{CRF_{electrolyzer} * Mass_{H_2SO_4} * H_2SO_4\ price}{Capacity\ factor * 365\left(\frac{days}{year}\right) * production\left(\frac{ton}{day}\right)} = \frac{1.07 * 8.62\ ton * 185\frac{USD}{ton}}{0.9 * 365\left(\frac{days}{year}\right) * 100\left(\frac{ton}{day}\right)}$$

$$= 0.05\ \frac{USD}{ton}$$

8. Other operational cost

$$Other\ cost_{per\ ton\ HCOOH} = 303.79\frac{USD}{ton} * 0.02 = 6.08\ \frac{USD}{ton}$$

9. Total cost for one ton of Product:

$$Cost_{HCOOH} = Cost_{electrolyzer} + Cost_{BoP} + Cost_{electricity} + Cost_{PSA} + Cost_{Liquid\ product\ separation} + Cost_{catalyst\ and\ membrane} + Cost_{CO_2} + Cost_{water} + Cost_{Electrolyte\ H_2SO_4} + Cost_{other}$$

$$= 16.48 + 10.54 + 303.79 + 12.77 + 224.68 + 23.69 + 38.24 + 2.13 + 0.05 + 6.08 = 638.48\ \frac{USD}{ton}$$

10. Profit:

$$Profit = HCOOH_{per\ ton\ HCOOH} - Cost_{HCOOH}$$

$$= 740\ \frac{USD}{ton} - 638.45\ \frac{USD}{ton} = 101.55\ \frac{USD}{ton}$$