Hollow-fiber membrane bioelectrochemical reactor for domestic wastewater treatment
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Abstract

BACKGROUND: Microbial fuel cells (MFCs) are potentially advantageous as an energy-efficient approach to wastewater treatment; however, the quality of the MFC effluent has not been well addressed. In this study, a membrane bioelectrochemical reactor (MBER) was developed through integrating hollow-fiber ultrafiltration membranes into a tubular MFC to improve the effluent quality.

RESULTS: This MBER was operated with an acetate solution or domestic wastewater (primary effluent) for more than 200 days. The MBER removed 43–58% of total chemical oxygen demand (COD) from the acetate solution and achieved 30–36% coulombic efficiency. When treating the wastewater, the MBER was able to maintain almost 90% COD removal and an effluent turbidity <1 NTU. A strategy of periodic backwash and membrane relaxation led to a slow increase in the transmembrane pressure (TMP) from zero to 15 kPa in more than 40 days at hydraulic retention time (HRT) 36 h. However, both lower HRTs and high organic loading rates rapidly increased the transmembrane pressure.

CONCLUSION: A proof of concept of an MBER was presented and shown to be effective in contaminant removal. Preliminary energy analysis suggests that the MBER could theoretically produce sufficient energy from the acetate solution to support the pumping system. These results demonstrate the feasibility of the MBER concept and the challenges for further development of the MBER system.

INTRODUCTION

Microbial fuel cells (MFCs) have been widely studied as a potential approach for simultaneous wastewater treatment and bioenergy production,1 and research has focused primarily on improving electricity generation in MFCs by investigating their architecture and operation, electrode and membrane materials, substrates, electrochemistry, microbiology, and new functions.2–6 The primary function of an MFC is wastewater treatment; however, the issues surrounding MFC effluent quality have not been sufficiently addressed, possibly because most studies fed the MFCs with synthetic solutions that are ‘cleaner’ than real wastewater. A few studies that used real wastewater demonstrated effective organic removal in MFCs,7 but information about the quality of the treated effluent, such as turbidity, is very limited. MFCs are designed to replace the secondary treatment,7 which usually requires a tertiary treatment such as filtration and/or disinfection to further purify the treated effluent.8 Producing a high-quality effluent for either direct discharge or water reuse will make MFCs more competitive as a wastewater treatment technology.

Existing wastewater treatment can achieve high-quality effluents using membrane bioreactors (MBR), which integrate membrane filtration using microfiltration (MF) or ultrafiltration (UF) membranes with suspended growth bioreactors. The MBR technology has been commercialized for municipal or industrial wastewater treatment.9,10 The filtration elements can be either submerged (internal) in a bioreactor such as an activated sludge tank, or installed as a sidestream (external). To recover energy from wastewater, an anaerobic MBR (AnMBR) is developed with the production of methane gas.11 Additional details about MBR/AnMBR technology are provided in several in-depth review publications.12–16

Filtration membranes have also been studied in MFCs, mostly as separators between the anode and the cathode, instead of filtration media.17 The use of MF/UF filtration membranes greatly improved power production, because they have a larger pore size, which facilitates ion transport;18,19 however, the increased permeation of oxygen and the loss of substrate across filtration membranes constrain their applications in MFCs.17 Recently developed osmotic MFCs use a forward osmosis membrane as both the separator for electrodes and the medium for water filtration (driven by an osmotic pressure),20,21 but the recovered water needs post-treatment such as reverse osmosis.

To polish the effluent using filtration membranes in MFCs, a possible strategy is to use MF/UF membranes as the filtration media, installed either inside or outside the anode compartment,
similar to that in MBRs. Ion exchange membranes are still required to separate the electrodes unless a membrane-less MFC is employed. In this study, a membrane bioelectrochemical reactor (MBER) was developed by installing hollow-fiber ultrafiltration membranes inside a tubular MFC. This design is different from recent studies of bioelectrochemical membrane reactors, which used biofilm formed on stainless steel mesh or nylon mesh as the filtration material in MFCs and relied on aerobic cathode for most COD removal. The MBER employed a passive oxygen supply by exposing the cathode electrode in the air. The MBER was operated with either a synthetic solution or actual wastewater for more than 200 days and examined its electricity generation performance, transmembrane pressure (TMP), and effluent quality.

MATERIALS AND METHODS
MBER construction
The MBER was constructed with a tube made of cation exchange membrane (CEM - Ultrex CMI7000, Membranes International, Inc., Glen Rock, NJ) (Fig. 1). The CEM was used to separate the anode and the cathode: the CEM tube formed an anode compartment with a liquid volume of ~920 mL and the cathode (electrode) was directly exposed in the air to avoid active aeration. Ten 20-cm long PVDF hollow-fiber (HF) membranes (150 000 Dalton, Litree Purifying Technology Co., Ltd., China) were installed around the carbon brush (Gordon Brush Mfg. Co., Inc., Commerce, CA), which was used as the anode electrode and pretreated by heating at 450 °C for 30 min. The HF membranes have pore size of 0.02 μm and a total surface area of about 0.011 m². The cathode electrode was one layer of carbon cloth (PANEX®30-PW03, Zoltek, Corporation, St. Louis, MO). Pt/C powder (10%, Etek, Somerset, NJ, USA) was mixed with Nafion solution and deionized water, and then applied to the surface (the side facing CEM) of the carbon cloth with a loading rate of ~0.3 mg Pt cm⁻². The cathode catalysts were air-dried for 24 h at room temperature before operation. Both anode and cathode electrodes were connected by copper wires to an external circuit across a resistance decade box (Tenma 7270, Springboro, OH, USA). The external resistance for the MBER operation was determined by polarization tests to achieve high power output.

Operating conditions
The MBER was inoculated with anaerobic digester sludge from a local wastewater treatment facility (South Shore, Oak Creek, WI) and operated at a room temperature of ~20 °C. The MBER was first operated with a synthetic solution that was fed into the anode compartment from the bottom of the MBER and the permeated water (effluent) flow was controlled by a peristaltic pump. The synthetic solution contained (g L⁻¹ tap water): sodium acetate, 0.75; NH₄Cl, 0.15; NaCl, 0.5; MgSO₄, 0.015; CaCl₂, 0.02; NaHCO₃, 0.1; KH₂PO₄, 0.53; K₂HPO₄, 1.07; and 1 mL of trace elements. The HF membranes were operated intermittently to reduce membrane fouling (4 min flow and 1 min rest). The flow rate was 0.7–1.4 mL min⁻¹, resulting in a hydraulic retention time (HRT) of 27–19 h. The anode solution was recirculated at 50 mL min⁻¹. After about 75 days, the synthetic solution was changed to the wastewater collected from the South Shore Wastewater Treatment Plant. Tap water was used as catholyte to rinse the cathode electrode from the top to the bottom at 6.2 mL min⁻¹ for 110 days and then the catholyte was changed to the anode effluent at the same flow rate.

During the operation, the permeation flux was kept constant while the transmembrane pressure (TMP) across the membrane was measured by vacuum pressure meter. According to the instructions of the HF membrane manufacturer, the preferred operating TMP should be below 35 kPa. For off-line membrane cleaning, the membranes were removed from the MBER, soaked in pH 12 solution (500 ppm NaOH and 100–500 ppm NaOCl) for 30 min, and then backwashed for 30 min using the same chemical solution.

Measurement and analysis
The cell voltage was recorded every 3 min by a digital multimeter (2700, Keithley Instruments, Inc., Cleveland, OH). The pH was measured using a benchtop pH meter (Oakton Instruments, Vernon Hills, IL, USA). The concentrations of chemical oxygen demand (COD), ammonium, nitrite, and nitrate were measured using a colorimeter according to the manufacturer’s procedure (Hach DR/890, Hach Company, Loveland, CO, USA). The turbidity was measured using turbidimeter (DRT 100B, HF Scientific, Inc., Fort Myers, FL, USA). The polarization tests were performed by a potentiostat (Reference 600, Gamry Instruments, Warminster, PA, USA) at a scan rate of 0.2 mV s⁻¹. The power density and current density were calculated based on the anode liquid volume. The water flux from the HF membrane filtration was calculated through pumping flow rate and expressed as liter per surface area of the membrane per hour (L m⁻² h⁻¹ - LMH). Elemental analysis was performed on an X-ray energy dispersive spectrometer (EDS) (TopCon ABT-32). The EDS was conducted with an accelerating voltage of 15 kV and a working distance of 30 mm. The target elements included nitrogen, oxygen, sodium, magnesium, aluminum, silicon, phosphate, calcium, and copper.
Fourier transform infrared spectroscopy (FTIR) measurements were conducted on an infrared spectrophotometer (Bruker Corp VECTOR 22) equipped with diamond Attenuated Total Reflection within a range 4000–600 cm⁻¹. The efficiency of organic-to-electricity was expressed by coulombic recovery (CR) and coulombic efficiency (CE), which were calculated according to the following equations:

\[ CR = \frac{Q_{output}}{Q_{input}} = \frac{\sum I(A) t(s)}{96485 \left( \frac{C}{mol\ e^-} \right) \times COD_{total} (mol) \times 4 \left( \frac{mol\ e^-}{mol\ O_2} \right)} \]

\[ CE = \frac{Q_{output}}{Q_{input-r}} = \frac{\sum I(A) t(s)}{96485 \left( \frac{C}{mol\ e^-} \right) \times COD_{removed} (mol) \times 4 \left( \frac{mol\ e^-}{mol\ O_2} \right)} \]

where \( Q_{output} \) is the produced charge, \( Q_{input} \) is the total charge available in the added organic compounds, \( Q_{input-r} \) is the total charge available in the removed organic compounds, \( I \) is electric current and \( t \) is time. \( COD_{total} \) is the total COD input to the anode compartment in the period of time \( t \), and \( COD_{removed} \) is the COD removed within time \( t \).

Energy estimation
Energy consumption in the MBER was mainly due to recirculation of the anolyte and the catholyte by pumps. The power required for the recirculation pump was estimated at:

\[ P = \frac{Q \gamma E}{1000} \]

where \( P \) is power requirement (kW), \( Q \) is flow rate (m³ s⁻¹), \( \gamma \) is 9800 N m⁻³, and \( E \) is the hydraulic pressure head (m). For example, for the MBER treating the primary effluent at HRT of 19 h, \( Q \) was 8.33 × 10⁻² m³ s⁻¹ (50 mL min⁻¹) and 1.03 × 10⁻⁷ (6.2 mL min⁻¹) m³ s⁻¹ for anode and cathode recirculation, respectively, and the hydraulic pressure head loss was assumed to be 0.098 m; thus, the total power required for the recirculation pumps was 9 × 10⁻² kW. The energy for membrane filtration was calculated based on average vacuum pressure (22 kPa) and permeate flow (2 × 10⁻⁶ m³ s⁻¹), yielding a suction pumping power requirement of 4.31 × 10⁻⁷ kW (Table S1, Supplementary material). The influent pumping power requirement was 4.73 × 10⁻⁶ kW, according to the hydraulic head loss of 0.35 m and average inflow rate of 1.38 × 10⁻⁶ m³ s⁻¹. In total, the power requirement for pumping was 1.38 × 10⁻⁶ kW. Divided by influent flow rate, the total energy requirement was 2.78 × 10⁻⁷ kWh m⁻³ (or 0.028 in kWh m⁻³ in Table S2).

RESULTS
The MBER treating synthetic solution
The MBER started as a conventional MFC in which a synthetic solution was actively fed into the bottom of the reactor and flowed out from the top. In the first 20 days, the external resistance was gradually decreased from 1000 to 10 ohm, and the current generation increased to ~16 mA. Then, the operation was switched to the ‘MBER’ mode and the hollow-fiber membranes began water extraction. At an HRT of 19 h and water flux of 5.45 LMH, the current generation varied mostly between 10 and 17 mA (Fig. S1) and the COD removal was about 45% at a loading rate of 0.61 kg m⁻³ d⁻¹. Depending on the current output, the CE was between 30 and 40% and the CR was 14–20%. The TMP increased very rapidly to 40–50 kPa within 20 h, indicating the occurrence of membrane fouling (Fig. S2). Backwash was conducted using the anode effluent for 1 h when the TMP was above 40–50 kPa; however, the increasing trend was not obviously changed (Fig. S2).

To alleviate the TMP increase and membrane fouling, the HRT was increased to 27 h on day 39, and thus water flux decreased to 3.82 LMH. Figure 2(A) shows the overall electricity generation as a polarization curve. The operating current was maintained around 10 mA, lower than that at HRT 19 h because fewer organic compounds were supplied to the anode (Fig. 3(A)). The pace of the TMP increase clearly slowed, and it took about 1 week to increase the TMP from 5 to 26 kPa (Fig. 3(B)). Then, the membranes were taken out of the MBER and rinsed with DI water to remove the fouling layer. Before resuming the operation, the MBER was backwashed for 1 h at water flux of 5.45 LMH (arrow a on Fig. 3(B)). After physical cleaning, the TMP increased from zero to 37 kPa in six days. Then, a chemical cleaning of the membranes was conducted (arrow b on Fig. 3(B)), in which membranes were soaked in pH 12 NaOH solution for 2 h followed by a 30-min backwash with DI water. When the operation resumed, the TMP increased from zero to 32 kPa in 8 days, indicating that chemical cleaning restored membrane function better than physical cleaning. At an organic loading rate of 0.51 kg m⁻³ d⁻¹, the MBER removed 58% of the COD and produced an effluent containing 246±39 mg COD L⁻¹ (Fig. 3(C) and Table 1). The CE and the CR were about 30% and 17%, respectively. The turbidity of the effluent was 0.41 ± 0.19 NTU.
The TCOD concentration in the MBER effluent slightly decreased ammonium concentration from 20 mg L⁻¹ to 17 mg L⁻¹ (Table 1). The soluble COD concentration could be as low as 20 mg L⁻¹ after major storms (arrow a in Fig. 4(C)).

In Phase 1, the HRT was reduced to 36 h. Before the new operation, the membranes were removed from the MBER and membrane relaxation was adopted to slow down TMP increase and membrane fouling: the MBER was backwashed for 1 min at 4 mL min⁻¹ every 29 min. The TMP dramatically increased from 15 to 24 kPa on day 152 and continued to rise to 39 kPa over the next 10 days. Then, the hollow-fiber membranes were removed for cleaning (Table S3). At restart of the MBER TMP increased an increase to 45–50 kPa in 14–15 days. The effluent turbidity of about 0.2 mg NO₃⁻-N/L. The total phosphate was ∼11% and the CR was about 9% (based on TCOD). On day 110, the catholyte was changed from tap water (without any pH control) to the anode effluent, and current generation was not obviously affected (arrow b in Fig. 4(A)). The MBER slightly decreased ammonium concentration from 20±6 to 17±3 mg NH₄⁺-N/L, probably due to the ammonia loss through bacterial growth and/or cation exchange membrane. The nitrate concentration in both the influent and effluent of the MBER was 20±7 mg L⁻¹. The CE was ∼11% and the CR was about 9% (based on TCOD).

In Phase 2, the HRT was reduced to 26 h. Before the new operation, the membranes were removed from the MBER and cleaned using a pH 12 NaOH solution (Table S3). The TMP increased from zero to 15 kPa in 28 days, faster than Phase 1 (Fig. 4(B)). The current was generally higher than the previous phase due to higher organic loading in a shorter HRT, and largely affected by the actual organic concentration in the wastewater. A high peak of 8 mA was observed, because of a high TCOD (300 mg L⁻¹) and SCOD (152 mg L⁻¹) in the feed at day 130. Despite the higher organic influx, the MBER effluent contained a TCOD concentration of about 20 mg L⁻¹, indicating stable performance of the organic removal. During this period, the CE and the CR based on TCOD was 17% and 15%, respectively. The average turbidity was below 0.5 NTU (Table 1). In Phase 3, operation with a new HRT of 19 h started on day 151 without cleaning the membrane. The TMP dramatically increased from 15 to 24 kPa on day 152 and continued to rise to 39 kPa over the next 10 days. Then, the hollow-fiber membranes were removed for cleaning (Table S3). At restart of the MBER TMP exhibited an increase to 45–50 kPa in 14–15 days. The effluent quality (COD removal and turbidity) was similar to the previous operation (Table 1). A brief Phase 4 was operated at an HRT of 15 h, and a quick increase of TMP to 45 kPa in 7 days was observed; at this rate, the membrane would require frequent cleaning and thus make this operation unsustainable.

### Energy balance

The energy balance was estimated by analyzing energy consumption and production (kWh m⁻³) in the MBER, as summarized in Table S2 and shown in Fig. 5. Energy consumption in the MBER is mainly due to the pumping system, which consists of three parts: influent feed, permeate (vacuum pumping), and recirculation of the anolyte and the catholyte. The power required for the pumps shown in Table S1 was estimated according to a previous study and more details can be found in the Supporting Information. It was found that recirculation consumed the most energy (Table S2), and a longer HRT resulted in more energy consumption, because the recirculating pumps were operated for a longer period of time to treat every cubic meter of wastewater. The estimated energy requirement varied between 0.027 and 0.040 kWh m⁻³ at HRTs of 15–36 h (Table S2). The energy produced in the MBER was calculated by integrating power over time and was affected by organic loading rates. The MBER fed with the acetate solution that had a higher COD produced 0.036–0.038 kWh m⁻³, much higher than 0.003–0.025 kWh m⁻³ with the wastewater (Table S2). According to the estimation of energy consumption, the MBER could theoretically produce enough energy to support the pumping system when treating the acetate solution. The energy production from wastewater was lower than the energy requirement, and at the shortest HRT of 15 h production was close to the consumption (Fig. 5). The low organic content in the wastewater used in this study led to low energy production.

### Table 1. Characterization of the influent and the effluent of the MBER with different substrates and HRTs

<table>
<thead>
<tr>
<th>Substrate</th>
<th>HRT (h)</th>
<th>Rₑₑₑₑ (Ω)</th>
<th>TCOD (mg L⁻¹)</th>
<th>SCOD (mg L⁻¹)</th>
<th>NH₄⁺-N (mg L⁻¹)</th>
<th>Rₑₑₑₑ (%)</th>
<th>CE (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Acetate solution</td>
<td>19</td>
<td>10</td>
<td>485 ± 16</td>
<td>-</td>
<td>-</td>
<td>265 ± 10</td>
<td>45.3</td>
</tr>
<tr>
<td>Acetate solution</td>
<td>27</td>
<td>10</td>
<td>583 ± 5</td>
<td>-</td>
<td>-</td>
<td>246 ± 39</td>
<td>57.8</td>
</tr>
<tr>
<td>Primary effluent</td>
<td>36</td>
<td>60</td>
<td>152 ± 38</td>
<td>83 ± 22</td>
<td>20 ± 6</td>
<td>20 ± 7</td>
<td>6.55</td>
</tr>
<tr>
<td>Primary effluent</td>
<td>26</td>
<td>45</td>
<td>158 ± 106</td>
<td>72 ± 53</td>
<td>24 ± 3</td>
<td>16 ± 6</td>
<td>12 ± 1</td>
</tr>
<tr>
<td>Primary effluent</td>
<td>19</td>
<td>45</td>
<td>235 ± 122</td>
<td>105 ± 71</td>
<td>20 ± 9</td>
<td>23 ± 8</td>
<td>18 ± 3</td>
</tr>
<tr>
<td>Primary effluent</td>
<td>15</td>
<td>45</td>
<td>199 ± 8</td>
<td>71 ± 3</td>
<td>22 ± 2</td>
<td>21 ± 7</td>
<td>20 ± 1</td>
</tr>
</tbody>
</table>

Rₑₑₑₑ : external resistance.

Rₑₑₑₑₑₑ : TCOD removal efficiency.
Figure 3. MBER performance with synthetic solution: (A) current generation, (B) transmembrane pressure (TMP), and (C) COD concentrations in the influent and the effluent. Arrow a: physical cleaning; arrow b: chemical cleaning.

Membrane fouling
Due to the lack of active anti-fouling methods, reducing membrane fouling in an MBER is important for successful operation. In this study, both high organic loading rates and high water flux accelerated membrane fouling. Physical cleaning with DI water to remove fouling layers had a limited effect on membranes when treating wastewater. Chemical cleaning employing a high pH solution (pH 12), sodium hypochlorite, and backwash generally restored the water flux (Table S3) and cleaned the membrane surface (Fig. S3). The EDS spectrum revealed that the fouled membrane contained more elements, such as nitrogen, calcium, magnesium, copper, aluminum, silicon, and chloride, than the new membrane and there was almost no difference between the interior and the exterior membranes, except that calcium was detected on the interior side of the membrane (Fig. S4). The significant presence of nitrogen (16–18%) in the fouled membrane was possibly due to protein and inorganic nitrogen contaminants in the wastewater.

The two main components of the PVDF membrane, carbon and fluoride, also changed in weight. Using the mass of fluoride as a reference, the C/F ratio increased from 1.01 of the new membrane to 1.16 of the fouled membrane for the exterior membrane, and from 0.83 to 0.88 for the interior membrane (Table S4). The ratio of oxygen-to-fluoride also increased from 0.49–0.53 in the new membrane to 0.68 in the fouled membrane (Table S4). The excessive carbon, nitrogen, and oxygen in the fouled membrane indicated biofouling,28 which was also confirmed by the FTIR spectra, which showed a new peak on the fouled membrane at 1642 cm⁻¹ (Fig. S5), related to the amide I (C=O and C–N stretching), suggesting the presence of proteins.29

DISCUSSION
This study has demonstrated the proof of concept of a membrane bioelectrochemical reactor and provided an alternative approach to anaerobic treatment of domestic wastewater. During membrane cleaning, the anode electrode was taken out from the MBER and exposed to the air. It was observed that current generation was restored quickly after the anode electrode was
Limited nitrogen removal is one of the major barriers to the application of anaerobic treatment (e.g. AnMBR\(^{11}\)). Nitrogen, mainly ammonia, cannot be removed under anaerobic conditions unless ammox is developed. In this study, the ammonium concentration slightly decreased in the anode effluent, because of microbial growth and/or ammonium diffusion through the CEM. However, after the anode effluent was used as catholyte to rinse the cathode, an obvious reduction of ammonium concentration to 2 ± 1 mg L\(^{-1}\) was observed in the catholyte; meanwhile, the nitrate concentration increased to 10 ± 6 mg L\(^{-1}\), indicating the presence of nitrification on the surface of the cathode electrode. It is likely that nitrifying bacteria took advantage of oxygen and formed biofilm on the cathode electrode. This ‘fixed-biofilm’ nitrification did not significantly increase the suspended solids in the catholyte, and the turbidity of the catholyte was ~1 NTU, slightly higher than the 0.5 NTU of the anode effluent. Nitrate removal is critical to lowering the concentration of total nitrogen, but bioelectrochemical denitrification is not expected on the cathode because of the presence of oxygen. Previous denitrifying biocathode in MFCs\(^{10}\) raises the possibility that the cathode of an MBER may be specially designed to provide a partially anoxic condition to accomplish both nitrification and denitrification; to achieve this, shortcut nitrification–denitrification\(^{31}\) will be more advantageous. Phosphorus, on the other hand, could not be removed in the MBER. Like other anaerobic treatments of low-strength wastewater that are being studied, the MBER would require post-processes to remove or recover nutrients.\(^{11}\)

It should be noted that a precise energy analysis for a bench-scale system is very difficult, and the present information attempts to promote more (future) work related to energy analysis, because of the lack of systematic analysis of energy balance in MFCs. The accuracy of an energy balance in the present system is largely affected by the energy required for the pumping system, and our theoretic calculation could underestimate the energy consumption. If an AnMBR is used as a reference, because of the similarity between AnMBR and MBER, our system (treating actual wastewater) produced less than 15% of the energy that the pumping system consumes (0.2 kWh m\(^{-3}\)) in a full-scale AnMBR.\(^{32}\) The present MBER generated a low (operating) power density < 2 W m\(^{-3}\) with the wastewater; it is possible to further improve its power output by increasing organic loading rate and optimizing reactor configuration and operation. If the power output can be improved by 10 times to 20 W m\(^{-3}\), an MBER will produce sufficient energy to provide an energy-neutral treatment system.

Although the HRTs of 15–36 h applied to the MBER are within the range of AnMBRs treating domestic wastewater,\(^{11}\) a shorter HRT (and higher organic loading rate) will benefit wastewater treatment with less land footprint of the facility. To maintain a short HRT, membrane fouling must be reduced; in future studies, one can consider reducing sludge retention time (SRT) and involving active anti-fouling methods. The SRT of the present MBER was more than 180 days, and the previous study found that a short SRT (e.g. 30–60 days) could greatly slow down membrane fouling.\(^{33}\) Using biogas to remove foulants is a common approach to control fouling in AnMBRs; however, the use of a carbon brush electrode in an MBER isolated hollow-fiber membranes from contact with gas bubbles and could also trap bubbles with carbon fibers. Fluidized granular activated carbon (GAC) that has been used to control fouling\(^{26}\) will be considered to maintain a low TMP in a future MBER.

**CONCLUSIONS**

This study has demonstrated a proof of concept of an MBER for wastewater treatment. Further MBER development must understand and address some key challenges. For instance, the low CE and CR with real wastewater could be due to other electron sinks such as sulfate reduction and methanogenesis, and constructing a carbon balance would help understand the electron flow. It will be of strong interest to examine the treatment of wastewater with high COD content and stability of the MBER system during long-term operation. Membrane fouling must be effectively reduced. The results have suggested that installing membranes within the anode compartment may not be an optimal approach to integrate membranes with MFCs, due to the difficulty of membrane cleaning. Future MBERs will consider fluidized GAC when membranes are inside the anode compartment, or a side-stream membrane system that is placed sequentially to MFCs.

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SUPPLEMENTARY DATA

The power consumption and energy balances are shown in Table S1 and S2, membrane cleaning methods are in Table S3, and the mass ratios of the main elements are in Table S4. Current generation, TMP variation, membrane condition, and FTIR spectra are shown in Fig. S1–S5.

Supporting Information

Supporting information may be found in the online version of this article.

REFERENCES