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Direct carbonate electrolysis into pure syngas†

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Syngas, a mixture of carbon monoxide (CO) and hydrogen (H₂), is a feedstock for a wide variety of chemical processes and is currently produced from fossil fuels. The need to reduce carbon dioxide (CO₂) emissions motivates the production of syngas from atmospheric CO₂, powered by renewable electricity. Current CO₂ electrolyzers require costly separation processes to purify the CO₂ reactant stream and to remove unreacted CO₂ from the product stream. We demonstrate direct carbonate electrolysis (DCE) in a reactive capture system that avoids the initial CO₂ purification process and produces pure syngas with sufficient CO content for direct industrial use (H₂/CO ratios of 1–2). The DCE system incorporates a composite CO₂ diffusion layer (CDL) that attains high CO selectivity by achieving high alkalinity and available CO₂ concentration at the cathode. Applying this strategy, we produce pure syngas in the cathode outlet gas stream with a H₂/CO ratio of 1.16 at 200 mA cm⁻², corresponding to a CO faradaic efficiency (FE) of 46% and an energy intensity of 52 GJ tsyngas⁻¹. By eliminating intensive upstream and downstream processes, DCE achieves syngas production with 13% less energy than CO₂ electrolysis combined with water electrolysis, 39% less energy than past carbonate reduction work, and 75% fewer emissions than the conventional fossil fuel based route.

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Broader context

The electrochemical conversion of captured CO₂ into CO could reduce CO₂ emissions while producing the carbon content of a valuable feedstock, syngas, for upgrade into long-chain hydrocarbons. Most CO₂ electrolyzers require pure gaseous CO₂ streams, and, thus, incur substantial capital and operational costs for CO₂ capture liquid regeneration and CO₂ purification. Additionally, CO₂ reactant can be lost to carbonates in the electrolyzer which crossover to the anode, regenerate, and mix with the O₂ rich anode gas stream. Excess CO₂ in the cathode gas product stream also demands separation. These sequential purification steps are costly and limit the viability of electroproduced chemicals and fuels. We demonstrate a reactive capture system that produces pure syngas with an industrial H₂/CO ratio through direct electrolysis of a CO₂ post-capture solution. We designed a CO₂ diffusion layer that achieves high local CO₂ reactant concentration and high alkalinity favourable for CO₂ conversion. This report illustrates the potential for the renewable electroproduction of syngas in a net-zero emissions future.

Introduction

Syngas is a commodity feedstock used in the production of hydrocarbons and oxygenates *via* methanol routes and Fischer–Tropsch synthesis.^{1,2} Syngas is currently produced from fossil-fuels *via* coal gasification and/or methane reforming.^{3,4} These

pathways are energy intensive and have high CO₂ emission intensities (1.5 tCO₂e tsyngas⁻¹).⁵

Electrochemical syngas production methods use renewable electricity to produce syngas with a lower carbon footprint. These methods combine CO from CO₂ electrolysis and H₂ from water electrolysis.^{6–8} However, current CO₂ electrolyzers require high-purity gaseous CO₂ feeds,^{9–12} and sourcing this CO₂ from air – *via* direct air capture – or from industrial sources, is costly.^{13–16} Within current electrolyzers, utilization of reactant CO₂ is low, and CO₂ is lost to carbonates and crossover to the anode.^{17–19} As a result, both the anodic and the cathodic outlet streams require CO₂ separation.^{20,21}

Reactive capture is an electrolysis approach that shortens this process (Fig. 1A). This pathway avoids the thermally-driven

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Fig. 1 Direct carbonate electrolysis to produce syngas enabled by a CDL. The chemical balance of carbon capture from air, anodic OER, CO_2 regeneration, CO evolution reaction, H_2 evolution reaction, and *in situ* CO_2 recapture by OH^- are presented in eqn (1)–(6), respectively. (A) Schematic of DCE integrated with carbon capture in a reactive capture system. (B) Schematic of the conversion of carbonate into CO_2 facilitated by the CDL and protons, and subsequent conversion of CO_2 into CO via electroreduction on the Ag catalyst. Unreacted CO_2 is recaptured by OH^- generated as a by-product of CO and H_2 evolution. The carbonate electrolyte is recirculated.

capture liquid regeneration from the post-capture solution (carbonate electrolyte in the case of hydroxide-based direct air capture) and the subsequent CO_2 dehydration, compression, and transportation steps.^{22–26} In the electrolyzer, protons are generated by the anodic oxygen evolution reaction (OER) and transported to the cathode through a cation exchange membrane (CEM) (Fig. 1B). The protons then react with the carbonate ions to regenerate CO_2 *in situ*. Syngas is produced through co-synthesis of CO from the regenerated CO_2 and H_2 from the aqueous solution. Unreacted CO_2 is recaptured by hydroxide ions (OH^-), a by-product of CO and H_2 evolution, to form carbonate.

Previous studies of reactive capture using bicarbonate electrolyte have demonstrated high CO selectivity (CO faradaic efficiency (FE) > 50%). However, the limited CO_2 recapture capacity of bicarbonate electrolyte results in a product gas stream diluted with CO_2 .^{26–28} Direct electrolysis of carbonate (rather than bicarbonate) electrolyte allows for the collection of high-purity gaseous products, evidenced by the lack of CO_2 (<400 ppm) detected in the gas stream.²⁹ Carbonate electrocatalytic conversion into syngas has previously been achieved

with a H_2/CO ratio of 3 (CO FE of 25%) and energy intensity of $86 \text{ GJ tsyngas}^{-1}$. However, this mixture does not have sufficient CO content to meet industrial syngas standards.^{30–34}

Here, we present an adlayer strategy that modulates the cathode pH and maximizes CO_2 conversion to produce syngas with a H_2/CO ratio in the industrially relevant range (1–2, and corresponding to a CO FE between 33% to 50%). We develop a composite CO_2 diffusion layer (CDL) that enables cathode alkalinity to favour CO_2 electrolysis and increases CO selectivity by limiting the diffusion of protons to the cathode. We achieve a H_2/CO ratio of 1.16 (CO FE of 46%) at a current density of 200 mA cm^{-2} . An energy intensity of $52 \text{ GJ tsyngas}^{-1}$ was achieved, resulting in a 39% energy saving compared to the previous carbonate electrolysis report.

Results & discussion

Increasing CO_2 conversion

Previous (bi)carbonate electrolyzers have employed a bipolar membrane (BPM) to dissociate water and provide protons and



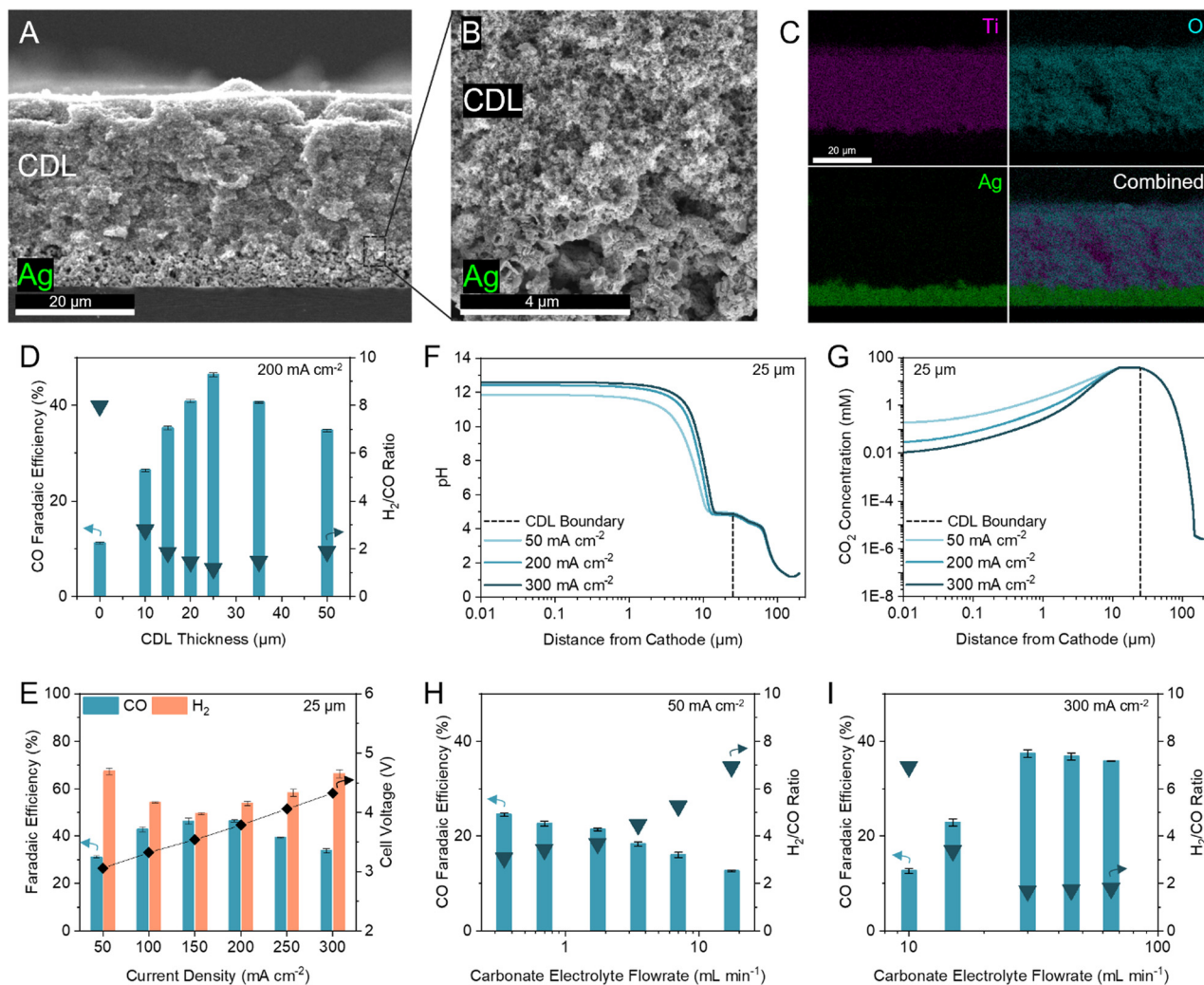


Fig. 3 Optimization of the CDL for industrial H₂/CO ratio. (A–C) Cross-sectional scanning electron microscopy (SEM) image of the CDL evenly air-brushed onto the Ag catalyst atop a silicon wafer (A), with seamless interfacial contact between the CDL and Ag catalyst (B), and corresponding energy dispersive X-ray (EDX) spectroscopy elemental mapping of Ti, O, and Ag (C). (D) FE towards CO and H₂/CO ratio at 200 mA cm⁻² with thicknesses of the CDL between 0 and 50 μm. (E) FE towards CO and H₂ at current densities between 50 to 300 mA cm⁻² with 25 μm CDL. Corresponding full cell voltages are noted on the secondary y-axis. (F) One-dimensional multi-physics modelling of pH at distances from the cathode and current densities of 50, 200, and 300 mA cm⁻² for 25 μm CDL. (G) One-dimensional multi-physics modelling of CO₂ concentration at distances from the cathode and current densities of 50, 200, and 300 mA cm⁻² for 25 μm CDL. Accompanying models of HCO₃⁻ and CO₃²⁻ concentrations are provided in the ESI† (Fig. S5). (H) FE towards CO and H₂/CO ratio at 50 mA cm⁻² with carbonate electrolyte flowrates between 0.35 to 17.5 mL min⁻¹ showing an increase in CO FE with decreasing flowrate. (I) FE towards CO and H₂/CO ratio at 300 mA cm⁻² with carbonate electrolyte flowrates between 10 to 65 mL min⁻¹ showing an increase in CO FE with increasing flowrate. Typical carbonate electrolyte flowrates are provided in the ESI† (Table S5). Error bars represent the standard deviation of at least three samples measured under identical conditions.

CO₂ availability. Thinner CDLs have a smaller gap between the catalyst and the CEM which shortens the proton diffusion distance and results in a lower cathode pH. Despite having the highest CO₂ concentrations in our simulations, the selectivity of the system with thinner CDLs was not optimal and approached the performance of the zero-gap configuration due to insufficient cathode alkalinity (Fig. 2B). As the CDL thickness is increased, the pH at the cathode increased but the local CO₂ concentration decreased (Fig. 2C). With the cathode pH plateauing at thicknesses greater than 25 μm, this CDL thickness provided sufficient CO₂ availability while maintaining an alkaline cathode environment to suppress HER.

We compared the difference in *iR*-compensated voltage of a zero-gap configuration to an otherwise identical system with a 25 μm CDL, with both experiments using an identical hydrogen-evolving catalyst (Supplementary note 3, ESI†). At 200 mA cm⁻², a voltage increase of 117 mV was observed for the CDL system which corresponded to an increase in pH of 2, consistent with the multi-physics model which predicted a pH increase of 1.7 (Fig. 2B and Fig. S8, ESI†). These findings suggest that the CDL increases the pH of the cathode environment to favour CO₂ conversion and thereby yielded a higher CO FE.

We screened the selectivity and full cell voltage of the optimized 25 μm CDL at current densities between 50 to





Fig. 4 Long-term operation of DCE over 23 hours with CO FE, H_2/CO ratio, full cell voltage, and capture solution pH noted. Experiment conducted at a constant current density of 100 mA cm^{-2} with a 1 cm^2 active area. Schematic and picture of the experimental set-up are provided in the ESI† (Fig. S10A and B).

300 mA cm^{-2} and found that the CDL resulted in minimal voltage penalties while significantly increasing CO FE compared to the zero-gap configuration (+0.23 V and +35.3% CO

FE at 200 mA cm^{-2}) (Fig. 3E). Above 200 mA cm^{-2} , the CO selectivity decreased, and the CO partial current plateaued, which indicated a CO_2 mass transfer limit at the cathode (Fig. S9, ESI†). We hypothesized that the decrease in CO FE at the lower and higher current densities were due to an imbalance of cathode alkalinity and CO_2 concentration at these extremes. The multi-physics model showed that compared to operating at 200 mA cm^{-2} , the pH is lower at 50 mA cm^{-2} , while the CO_2 concentration is lower at 300 mA cm^{-2} (Fig. 3F and G). To investigate further, we decreased the carbonate flow rate while operating at 50 mA cm^{-2} and found that the CO FE increased (Fig. 3H). A slower flowrate increases the local pH due to the accumulation of OH^- . At 300 mA cm^{-2} , the CO FE increased as the carbonate electrolyte flowrate increased up to 30 mL min^{-1} ; however, further increases in flowrate resulted in similar, or slightly decreased, CO FE (Fig. 3I). This result suggests that at high current densities, the local environment is excessively alkaline, and CO_2 availability is low. Increasing the carbonate electrolyte flowrate lowered the local pH and thereby increased the availability of CO_2 for reaction.

To assess the long-term stability of the engineered CDL, we operated the DCE system with continuous CO_2 capture and

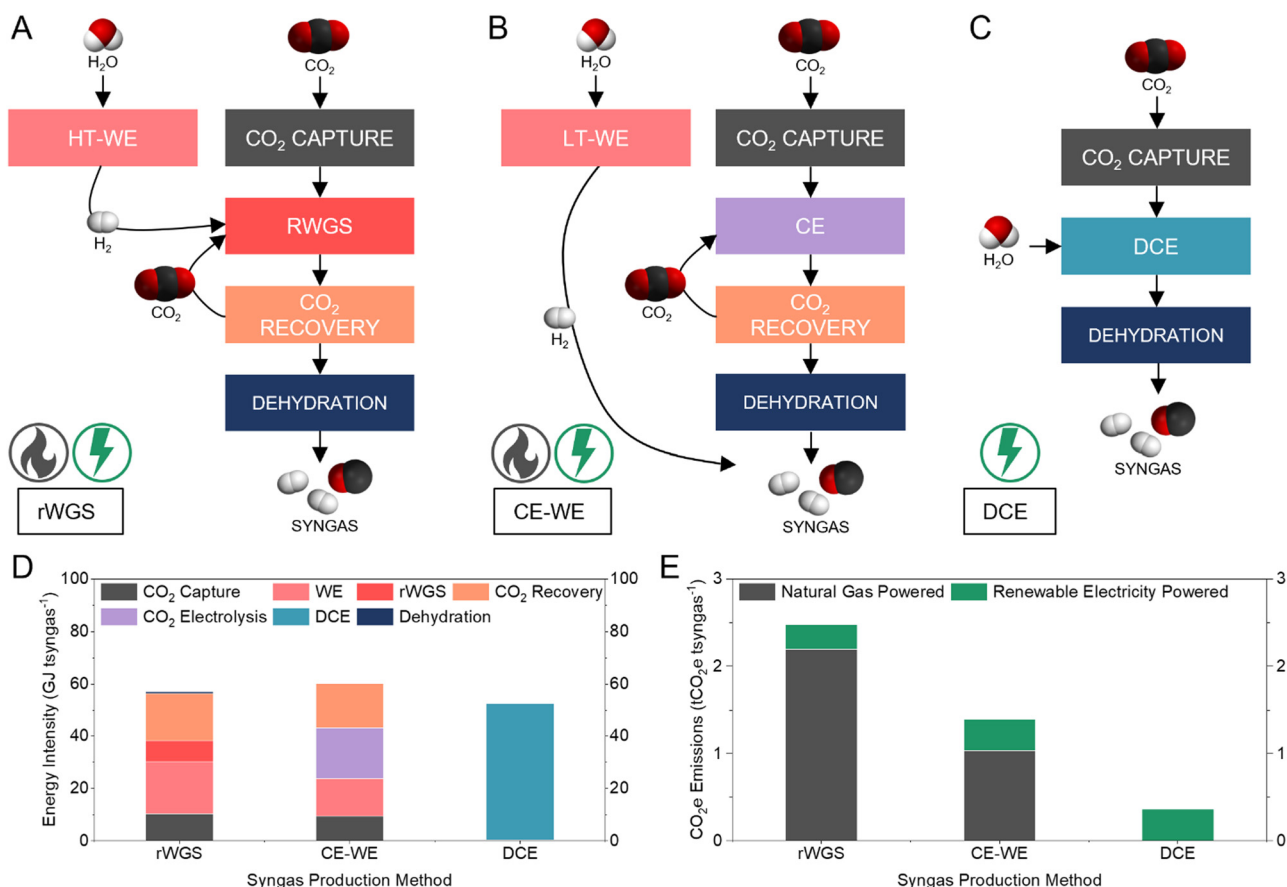


Fig. 5 Comparison of three syngas production methods. (A–C) Schematic showing process pathways, major chemical inputs and outputs, and energy source of rWGS (A), CE-WE (B), and DCE (C). (D) Energy intensity comparison to produce one tonne syngas using rWGS, CE-WE, and DCE. Syngas dehydration energies for CE-WE and DCE are too small to be seen in this figure. Detailed breakdown available in the ESI† (Table S3). (E) CO_2e emissions to produce one tonne syngas using rWGS, CE-WE, and DCE. The CO_2e associated with the energy input in each process is considered. Detailed breakdown available in the ESI† (Table S4).



Two hydrophilic microporous membrane filters were used: 125 μm PVDF (Filter 1, 0.45 μm pore size) was purchased from Sigma Aldrich and 100 μm nylon (Filter 2, 5 μm pore size) was purchased from Sterlitech. Both filters were used as received.

Operation of the electrochemical cell

The carbonate electrolysis experiments were performed in a 1 cm^2 electrolyzer with serpentine flow channels ingrained in both the stainless-steel cathode and the titanium anode current collectors. The electrolyzer was assembled by placing a CEM (Nafion 117) over the cathode, then placing the anode on the membrane. In all experiments, unless otherwise specified, the cathode feedstock was a carbonate electrolyte prepared by purging CO_2 at 80 sccm into 85 mL of 2 M KOH for 40 minutes, similar to a previous report.²⁹ The anode was fed with 0.05 M H_2SO_4 . After starting the experiment, the carbonate electrolyte was continuously purged with Ar gas flowing at 20 mL min^{-1} . The first gas sample is typically collected 20 minutes after the start of the experiment to ensure complete purging of excess CO_2 from the electrolyte preparation process and even mixing of gaseous products. The electrochemical measurements were performed with a potentiostat (Autolab PGSTAT204) and the full cell voltages reported are not iR compensated unless otherwise specified.

Product analysis

The cathode gas outlet stream was analyzed in 1 mL sample volumes by a gas chromatograph (PerkinElmer Clarus 590) coupled with a thermos conductivity detector (TCD) and flame ionization detector (FID). The gas chromatograph used Ar gas as the carrier (99.999%, Linde) and was equipped with a Molecular Sieve 5A Capillary Column and a packed Carboxen-1000 Column. Liquid product detection was performed using proton nuclear magnetic resonance spectroscopy ($^1\text{H NMR}$) on an Agilent DD2 600 spectrometer in D_2O using water suppression mode, with dimethyl sulfoxide (DMSO) as the internal standard. Calculation of FE and energy efficiency are included in Supplementary note 1 in the ESI.† pH measurements were conducted using Apera Instruments AI311 Premium Series PH60.

Author contributions

D. S. and E. H. S. supervised the project. Y. C. X. and C. M. G. conceived the study. Y. C. X. designed and performed all the electrochemical experiments and generated the figures. G. L., C. P. O., R. K. M., and M. F. assisted with electrochemical experiments. C. M. G., G. L., C. P. O., and J. Z. assisted with CDL design. T. A. assisted with pH investigation experiment design. Y. C. X. and Y. X. performed gaseous product analysis. J. L., P. P., and J. E. H. performed liquid product analysis. J. P. E. and R. K. M. assisted with energy analysis. Y. Z. performed SEM imaging. S. L. performed system modelling. A. S. R. assisted with system modelling. Y. C. X. and D. S. wrote the manuscript. All authors discussed the results and assisted during manuscript preparation.

Conflicts of interest

There are no conflicts to declare.

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