Syngas (CO–H₂) production using high temperature micro-tubular solid oxide electrolyser

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CO₂ and/or H₂O were reduced to CO/H₂ in micro-tubular solid oxide electrolyser with yttria-stabilized zirconia (YSZ) electrolyte, Ni-YSZ cermet cathode and strontium(II)-doped lanthanum manganese (LSM) oxygen-evolving anode. At 822 °C, the kinetics of CO₂ reduction were slower (ca. −0.49 A cm⁻²) at 1.8 V) than H₂O reduction or co-reduction of CO₂ and H₂O, which were comparable (ca. −0.83 to −0.77 A cm⁻² at 1.8 V). Performances were improved (−0.85 and −1.1 A cm⁻² for CO₂ and H₂O electrolysis, respectively) by substituting the silver current collector with nickel and avoiding blockage of entrances to pores on the inner lumen of micro-tubes induced by silver paste applied previously to decrease contact losses. The change in current collector materials increased ohmic potential losses due to substituting the lower resistance Ag with Ni wire, but decreased electrode polarization losses by 80–93%. For co-electrolysis of CO₂ and H₂O, isotopically-labelled C¹⁸O₂ was used to try to distinguish between direct cathodic reduction of CO₂ and its Ni-catalysed chemical reaction with hydrogen from reduction of steam. Unfortunately, oxygen was exchanged between C¹⁸O₂ and H₂¹⁷O, enriching oxygen-18 in the steam and substituting oxygen-16 in the carbon dioxide, so the anode off-gas isotopic fractions were meaningless. This occurred even in alumina and YSZ tubes without the micro-tubular reactor, i.e. in the absence of Ni catalyst, though not in quartz tubes. Unfortunately, larger differences between the thermal expansion coefficients of quartz and YSZ precluded using a quartz tube to house the micro-tubular reactor. However, the kinetic results, CO/H₂ yields from off-gas analysis, diffusion considerations and model predictions of reactant and product gas adsorption on Ni suggested that syngas should be produced by electrochemical reduction of steam to H₂, followed by its Ni-catalysed chemical reaction with CO₂.

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1. Introduction

As contributions of renewable energy to electricity grids increase, large-scale energy storage becomes more important, because of the intermittency of renewables and the dynamics of power demand arising from social habits causing misalignments, precluding supply-demand balancing. Two different scales have to be addressed: (1) daily fluctuations and (2) seasonal fluctuations. Therefore, storing electrical energy in chemical bonds using electrolysis of H₂O and/or CO₂ to balance supply and demand could be a promising technology, as infrastructure to store chemicals (e.g. methane in gas grids) is already available. Furthermore, smoothing geographical variations could also be more facile.

Solid oxide electrolyser offer the advantage of fuel flexibility [1,2] in addition to thermodynamic and kinetic benefits of operating at temperatures >700 °C, meaning a larger proportion of electrical energy can be substituted by thermal energy [3] (Fig. 1). The proposed reaction scheme is outlined in reactions (1)–(4) for co-electrolysis of CO₂ and H₂O to produce syngas:

Fuel electrode (e.g. Ni): CO₂ + 2e⁻ → CO + O²⁻ (1)

Electrolyte (e.g. YSZ): O²⁻ (cathode) → O²⁻ (anode) (2)

Air electrode (e.g. LSM): \( \frac{1+v}{2}(2O₂⁻ → O²⁻ + 4e⁻) \) (3)

Overall: CO₂ + vH₂O → CO + vH + \( \frac{1+v}{2}O₂ \) (4)

Electrolysis of CO₂ on its own has the disadvantage that carbon formation via the Boudouard reaction (5) is possible, if the CO concentration is too high and temperatures too low [6]; carbon...
could also be formed by CO reduction. Hence, co-electrolysis is beneficial to produce syngas, because water has been reported to suppress carbon formation [3] and syngas can be used as the feedstock, e.g. in the Fischer-Tropsch process to produce synthetic liquid fuels [7] or direct methanation by the Sabatier reaction downstream of the electrolytically active zone [8]. However, the syngas composition achieved by the electrochemical processes can be altered by the (reverse) water gas shift reaction (6), which would limit the tunability of the syngas production.

\[2\text{CO} + \text{O}_2 \rightarrow \text{CO}_2 + \text{H}_2\]

\[2\text{CO} + \text{H}_2 \rightarrow \text{CO}_2 + \text{H}_2\text{O}\]

Co-electrolysis has been studied on various cathode materials including LSCM (lanthanum strontium chromium manganite)
into the role of the water gas shift reaction in co-electrolysis will be discussed based on the electrochemical performance, electrical impedance, gas diffusion considerations in conjunction with microstructure, off-gas composition, adsorption modelling and electrolysis of isotopically-labelled reactant gases.

2. Experimental

The micro-tubular reactors were fabricated by a dual-layer co-extrusion phase inversion technique to produce electrode | electrolyte hollow fibres, which after co-sintering were brush-coated with the outer anode layers [17]. The individual layers (Fig. 2) were a NiO – (ZrO2)0.92(Y2O3)0.08 | YSZ cathode | YSZ electrolyte | La0.8Sr0.2MnO3 | YSZ anode | LSM current collection layer; NiO was activated by reduction to elemental nickel in a hydrogen atmosphere. The thicknesses of the individual layers were: ca. 230 μm Ni-YSZ | 26 μm YSZ | 25 μm LSM-YSZ | 25 μm LSM, with an active length of the outer anode of 15 mm.

The cells were previously analysed for gas tightness, mechanical strength, material homogeneity (using energy dispersive spectroscopy), porosity and pore size distribution. Details of the measurement techniques and results can be found elsewhere [17].

2.1. Scanning electron microscopy and energy dispersive spectroscopy

Microstructures of the micro-tubular solid oxide reactors were imaged using scanning electron microscopy (SEM). The exact dimensions of each fibre post-testing were confirmed using Hitachi table-top scanning electron microscopes (TM-1000 and TM3030) at 15 kV accelerating voltage. Energy dispersive spectroscopy (EDS) was performed at 20 kV on a JEOL instrument (JSM 6400).

2.2. Electrochemical Characterisation

2.2.1. Current collection, sealing and reactor assembly

Current collection on the Ni-YSZ inner electrode was achieved via two designs. Design one used twisted 0.25 mm diameter silver wires (99.9% purity, annealed, VWR International LLC, UK), wrapped with silver wool (99.9% purity, VWR International LLC, UK) fed through the inner lumen of the hollow fibre. The contact between wires/wool and the inner surface was enhanced by conductive silver paste (VWR International LLC, UK). The second design option was a tight coiled 0.25 mm diameter nickel wire (99.9% purity, Sigma Aldrich, UK) pulled through the inner lumen. The nickel wire was connected to silver wires (0.25 mm diameter) such that the latter could be connected outside of the reducing atmosphere without risk of oxidation. For the outer LSM electrodes, silver wires of 0.2 mm diameter (99.99%, annealed, Advent Research Materials, UK) were wrapped around the hollow fibre and fixed with silver paste. The length of individual wires from fibre to electrical connections was ca. 400 mm. The hollow fibre reactor was then cemented into an alumina tube (to feed the cathode gases) and the entire assembly sealed into a quartz tube with stainless steel end-caps as described previously [17].

2.2.2. Electrolyser Performance Tests

To generate steam, a controlled evaporation unit was used (CEM-unit, Bronkhorst, UK), in which water was mixed and heated with the feed gases to produce vapour at a controlled rate. However, the design of the system required a carrier gas flowing at ≥50 ml min⁻¹, so the active gas fraction (H₂O, H₂, CO₂, CO) was only 44% during electrolysis operation. All gases or liquid (water) flow rates were controlled by automated mass-flow controllers (Bronkhorst, UK); the individual feed gas scenarios are outlined in Table 1. For the oxygen-18 isotopically labelled co-electrolysis, 95.5% enriched C¹⁸O₂ was used (99.9% chemical purity, Isotec, Sigma Aldrich, UK) and individual cathodic feed gas compositions are outlined in the respective results section; the total flow of 90 ml min⁻¹ with 50 ml min⁻¹ carrier gas was kept constant and the anodic feed gas composition corresponded to the ‘standard’ (C¹⁸O₂) electrolysis scenarios.

The reactor was heated in a tubular furnace (Elite Thermal Systems Ltd., UK) to 650 °C and the nickel oxide reduced to nickel with hydrogen for 2 hours. After reduction, the temperature was increased to the operating temperatures (768, 822 and 872 °C), monitored by a N-type thermocouple. The silver wires were connected to a potentiostat/galvanostat equipped with a frequency

| CO(\text{g}) | CO₂(\text{g}) | \text{H}_2(\text{g}) | \text{H}_2\text{O}(\text{g})(\text{ml hr}^{-1}\text{ liquid feed before evaporation})/\text{ml min}^{-1}\text{gas volume} | \text{H}_2(\text{g}) | \text{Total} | Anode (LSM-YSZ|LSM) 10% O₂ in Argon/ ml min⁻¹ |
|----------------|----------------|----------------|---------------------------------------------------------------------------------|----------------|--------|----------------------------------|
| Heat-up Reduction | 8  | 32  | –  | –  | –  | 50  | 50  | 62  |
| Scenario 1: ‘CO₂ electrolysis’ | 8  | 32  | –  | 16 | –  | 50  | 56  | 62  |
| Scenario 2: ‘H₂O electrolysis’ | –  | –  | 8  | 32 (1.392) | –  | 50  | 50  | 90  |
| Scenario 3: ‘co-electrolysis’ | 4  | 16  | 4  | 16 (0.696) | –  | 50  | 90  | 90  |

¹ CO or H₂ were co-fed to the electrolyser to prevent re-oxidation of the nickel at open-circuit potential difference.
response analyser (FRA) 4.9 module for electrochemical impedance measurements (PGSTAT302N, Metrohm Autolab B.V., Netherlands). Electrochemical test procedures for operation were created in Nova 1.10.2 software (Metrohm Autolab B.V., Netherlands).

The electrochemical experiments included measuring the potential difference response of the cell to linearly scanned applied current and steady state measurements over varying time scales at constant potential difference or constant current. The intrinsic electrical properties of materials and interfaces were characterized using electrochemical impedance spectroscopic (EIS) measurements at open circuit potential difference ($U_{OCV}$) and operating potential differences in the frequency range of $10^3$–0.1 Hz (with an RMS amplitude of 0.01 V for constant potential difference operation.)

On-line analysis of the cathode (Ni-YSZ) off-gases was performed using mass spectrometry (Genesys 200D, ESS Ltd, UK) monitoring the reactant consumption/product formation, as well as detecting radial leakage across imperfections in the hollow fibres. A second optional mass spectrometer was used to detect the anodic (LSM-YSZ/YSZ) off-gases during experiments with $^{18}$O$\text{O}_2$ fed to the cathode. The water within the reactor was condensed in a quartz U-bend immersed in ice-water to avoid flooding of the mass spectrometer. A schematic of the equipment is outlined in Fig. 3.

3. Results and Discussion

3.1. Characterization of hollow fibre reactors

Finger-like voids extended into the cathode layer to a depth corresponding to 44% of its thickness, with the remaining part of the cathode being of a sponge-type nature with a mean porosity >30% after reduction. The mechanical strength was 156 (±26) MPa and the gas tightness $2.56 \times 10^{-10}$ mol m$^{-2}$ s$^{-1}$ Pa$^{-1}$, both values being comparable to values reported in the literature [18,19].

3.2. Reactor performance

3.2.1. Electrochemical performance of the MT-SOE

The theoretical open circuit potential differences for pure H$_2$O or CO$_2$ electrolysis are plotted in Fig. 4, using the Nernst equation (7a) and (7b). The measured values were generally in good agreement with their predicted values, implying gas-tight electrolytes.

$$\Delta E = \Delta E^\ast_{\text{CO}_2/\text{CO}} + \frac{RT}{2F} \ln \left[ \frac{P_{\text{CO}}}{P_{\text{CO}}^\ast} \right]$$  \hspace{1cm} \text{(7a)}$$

or

$$\Delta E = \Delta E^\ast_{\text{H}_2/\text{H}_2} + \frac{RT}{2F} \ln \left[ \frac{P_{\text{H}_2}}{P_{\text{H}_2}^\ast} \right]$$  \hspace{1cm} \text{(7b)}$$

Fig. 5(A) shows the effect of current density, cathode feed gas composition and inner electrode current collector design on cell potential difference at a fixed electrolyte thickness. Corresponding electrochemical impedance spectra are shown in Fig. 5(B) and summarized in Table 2. For all cases, H$_2$O electrolysis outperformed pure CO$_2$ electrolysis, whereas the co-electrolysis
The variation in ohmic resistance for different current collector designs, i.e., lower ohmic area specific resistance for silver wire plus silver paste \(^{15}\) compared to nickel wire only, was due to the higher electronic conductivity of silver [20] \((1.4 \times 10^5 \text{ S m}^{-1} \text{ at } 1073 \text{ K})\) compared to that of nickel [21] \((2.1 \times 10^6 \text{ S m}^{-1} \text{ at } 1073 \text{ K})\). For a detailed discussion about different current collector design options investigated for CO\(_2\) electrolysis alone, the reader is referred to [22], which also discusses the negative effect of silver paste on electrode polarization for the case of electrolysis of CO\(_2\) alone.

Notably, the difference in the apparent kinetics between H\(_2\)O and CO\(_2\) electrolysis also depended on the cathode current collection design. For the silver wire plus paste option, electrode polarization was 380% higher for CO\(_2\) electrolysis compared to that for H\(_2\)O electrolysis. Contrarily, for the nickel wire alone electrode polarization was only 61% higher for CO\(_2\) compared to H\(_2\)O splitting (at U\(_{\text{OCR}}\)), so CO\(_2\) electrolysis was affected more by the silver paste than was H\(_2\)O electrolysis.

Fig. 5(A) and (B) shows that this difference between the behaviour of CO\(_2\) and H\(_2\)O electrolysis varied with applied cell potential difference. However, even at an applied cell potential difference of e.g. 1.5V, 16% and 64% difference in overall performance between CO\(_2\) and H\(_2\)O electrolysis were measured for nickel wire only and silver wire plus silver paste current collector, respectively.

The current densities achieved of \(-0.5 \text{ A cm}^{-2}\) at 1.3V with 35% humidity at 822 \(^{\circ}\)C (steam electrolysis – feed gas scenario 2) were comparable with those of (micro)-tubular reactors reported elsewhere [11,23] ranging between ca. \(-0.4 \text{ A cm}^{-2}\) (850 \(^{\circ}\)C, 25%
steam) to ca. 0.7 A cm\(^{-2}\) (900°C, 30% steam), though the latter electrolyte thicknesses were thinner at only 20–15 μm.

3.2.2. Why was CO\(_2\) electrolysis more affected by current collector design?

Using analysis in differences of impedance spectra (8), as shown in Fig. 7, the variations between CO\(_2\) and H\(_2\)O/co-electrolysis were determined to be all due to low frequency resistances, which can be attributed to differences in gas diffusion rates [3,24]. The magnitude of changes was determined by the cathode lumen current collector design, with a large variation detected for the Ag paste-containing cells.

\[
\Delta Z(f_n) = \frac{\left| Z_{\text{gast}}(f_{n-1}) - Z_{\text{gast}}(f_{a-1}) \right| - \left| Z_{\text{gast}}(f_{a-1}) - Z_{\text{gast}}(f_{n-1}) \right|}{\ln(f_{n-1}/f_{a-1})}
\]

(8)

The application of conductive silver paste on the lumen surface was the only difference between the reactors, so most probably it increased gas diffusion transport resistances into the cathode microstructure (Fig. 8) by partial blockage of the entrances to surface pores. Energy dispersive spectroscopic analysis (Fig. 9) confirmed that silver was confined to the inner surface of the hollow fibre reactor, rather than within the Ni-YSZ scaffold.

Being a lighter and smaller [25] (265 versus 330 pm) molecule than CO\(_2\), water could diffuse more easily into the structure. Table 3 lists the binary diffusion coefficients, calculated according to the procedure by Fuller et al. [26], for the different feed gases relative to the helium carrier gas, with molecular diffusion rates of steam being predicted to be ca. 1.5 greater than for CO\(_2\).

However, silver paste probably caused partial blockage of the entrances to pores in the cathode, effectively decreasing the effective pore size. For smaller pores, molecule-wall interactions are more likely; hence, Knudsen diffusion would be more appropriate (9). The mean free path for the five gas species are ca. 510, 310, 240, 490, 410 nm for He, CO\(_2\), CO, H\(_2\)O and H\(_2\), respectively, calculated according to Eq. (10) using kinetic molecular diameters [25,27]. The mean peak pore sizes in ‘unblocked’ reduced hollow fibres were measured to be ca. 17, 400 and 700 nm using mercury porosimetry. Thus, the mean free paths of the reactants and products were of similar size to the pore sizes, so Knudsen diffusion was relevant.

Table 4 lists the diffusion coefficients for 400 and 700 nm pore sizes (approximating \(K_0\) to 0.25 \(d_{\text{pore}}\)); the ratio of the diffusion coefficients compared to water would be: 3, 0.64, 0.80 for H\(_2\), CO\(_2\) and CO, respectively, again CO\(_2\) being the slowest to diffuse.

\[
D_{i,k} = \frac{4}{3}K_0\sqrt{\frac{8RT}{\pi M_i}} \sim \frac{1}{M_i}
\]

(9)

\[
\lambda_m = \frac{RT}{\sqrt{2\pi d_{\text{pore}}N_iP}}
\]

(10)

3.2.3. Implication of performance differences of CO\(_2\) and H\(_2\) for their co-electrolysis

The electrochemical performance of co-electrolysis was much closer to that of H\(_2\)O than of CO\(_2\) electrolysis, implying that CO was produced mainly by the chemical reaction between CO\(_2\) and electro-generated H\(_2\) reaction (6).

3.2.3.1. Evidence I: Gas diffusion and the effect of microstructure. Firstly, the diffusion coefficients of CO\(_2\) were smaller than those for H\(_2\)O, with a minor enhancement for CO\(_2\) diffusion by increased binary diffusion coefficients when H\(_2\) was present. However, this positive effect for co-electrolysis would have been valid only for the molecular diffusion regime in larger finger-like voids. Within the sponge-type microstructure close to the electrode | electrolyte interface, Knudsen diffusion would dominate, independently of the gas mixture.
The anode functional layer thickness for hydrogen oxidation on a Ni-YSZ cermet electrode in a SOFC has been reported to be ca. 20–50 μm in the temperature range 700–1000 °C, depending on parameters such as the precursor particle size of YSZ and NiO [28,29]. As hydrogen oxidation kinetics are faster than steam reduction kinetics and H2O (or CO2) diffusion rates are slower than those of H2 (as outlined in Tables 3 and 4), the effective cathode active functional layer thickness in solid oxide electrolyser mode will have been greater than that in SOFC mode, electronic and ionic conductivities being the same for both operational modes. However, in SOE mode, a significant fraction of triple phase boundaries in the 220 μm thick NiO-YSZ cermet cathode would not have been electrochemically active. A quantitative estimate of that fraction would require modelling [30] beyond the scope of this paper. Furthermore, most of these active triple phase boundaries would be located in the sponge-type porosity at the electrode | electrolyte interface (Knudsen diffusion regime), where CO2 and CO Knudsen diffusion coefficients were significantly smaller than those of H2 and H2O.

As shown schematically in Fig. 10, this electrode functional layer results from the difference between the ionic conductivity of the YSZ electrolyte phase and the electronic conductivity of the nickel electrode phase, causing potentials, overpotentials and hence

Table 3

<table>
<thead>
<tr>
<th>Binary diffusion coefficients for the five feed gas species at 822 °C.</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
</tr>
<tr>
<td>-------------------</td>
</tr>
<tr>
<td>He</td>
</tr>
<tr>
<td>CO₂</td>
</tr>
<tr>
<td>CO</td>
</tr>
<tr>
<td>H₂O</td>
</tr>
<tr>
<td>H₂</td>
</tr>
</tbody>
</table>

Note: the silver paste was required for establishing contact between the silver wire and Ni-YSZ electrode, as previous experiments without the paste resulted in very poor contact and noisy electrochemical measurements. The nickel wire has a higher mechanical strength (Young’s modulus of 160 GPa compared to 46.6 GPa, for nickel and silver respectively [39,38]), so enabled better contact without the addition of conductive paste. Note: similar trends in terms of poor electrode performance were detected for a nickel wire with silver paste, confirming the silver paste as the distinguishing factor.

Table 4

<table>
<thead>
<tr>
<th>Knudsen diffusion coefficients for the five gas species at 822 °C.</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
</tr>
<tr>
<td>-------------------</td>
</tr>
<tr>
<td>400 nm pore</td>
</tr>
<tr>
<td>700 nm pore</td>
</tr>
</tbody>
</table>

The yield CO to H₂ ratio did not vary within the accuracy of the mass spectrometry. Obviously, the ratio of H₂/CO to H₂O/CO₂ increased with applied cell potential difference as the overall conversion increased.
Fig. 10. Schematic of zones in the Ni-YSZ cathode of activity for electrochemical reaction and of heterogeneous catalytic chemical reaction \((6)\), the (reverse) water gas shift reaction.

Fig. 11. Effect of temperature on product yield composition for cathode current collectors containing silver paste and nickel wire only and theoretical yields if the (reverse) water gas shift (WGS) equilibrium was achieved.
current densities to be spatially distributed in the direction(s) of current flow, with faradaic current densities being concentrated close to the electrode | planar electrolyte interface.

Therefore, the chemical reaction between CO₂ and electro-generated H₂ was probably confined within the finger-type voids (outer 44% of Ni-YSZ), which according to Fig. 10 would have been far less electrochemically active. The reverse water gas shift reaction would have been catalysed by the nickel; i.e. no triple phase boundaries are required, though interactions with the ‘ceramic support’ have been reported to enhance locally rates of the (reverse) water gas shift reaction [31,32]. Thus, a much larger number of active sites for the chemical reaction was provided within the bulk of the Ni-YSZ electrode (Fig. 11), enabling thermodynamic equilibrium for reaction (6) to be achieved.

3.2.3.2. Evidence 2: Gas compositional split of product gases. Fig. 11 shows the product compositional split for (electro-) chemically produced H₂ and CO from on-line mass spectrometric analysis of the cathode-off-gases for the two options of current collector design and the theoretical compositional split, if the (reverse) water gas shift equilibrium compositions were achieved. Note that the experiments to quantify outlet gas compositions were run under constant current at ca. 1.3, 1.5 and 1.7 V, but the differences in gas composition of the yield² were marginal, so that data in Fig. 11 represent averaged values. Charge yields of ca. 100% were confirmed using Faraday’s law.

For the nickel wire cathode current collector, the compositional yield split was ca. 50:50 (temperature dependent), whereas the electrochemical performance was closer to that of H₂O electrolysis than CO₂ electrolysis (80:20 H₂O:CO₂ electrolysis current density). However, the relatively small difference between steam electrolysis and (CO₂ + H₂O) co-electrolysis could also have been due to the lower partial pressure of H₂O in co-electrolysis 

\[ p_{\text{H}_2\text{O}} = 0.18 \times 10^5 \text{ Pa} \quad \text{and} \quad p_{\text{CO}_2} = 0.18 \times 10^5 \text{ Pa} \] 

compared to H₂O electrolysis alone \( p_{\text{H}_2\text{O}} = 0.36 \times 10^5 \text{ Pa} \). Thus, the H₂O reactant conversion would have also been twice in co-electrolysis mode, which could have increased reactant polarization.

Furthermore, when silver paste was applied to the current collector of the reactor’s lumen, thermodynamic equilibrium compositions were not achieved, as partial blockage of surface pores hindered CO₂ diffusing into the electrode structure. As the heterogeneously-catalysed water gas shift reaction was fast [33,34], if CO₂ was present within the Ni-YSZ cathode and penetrated to the electrode functional layer to react electrochemically, thermodynamic equilibrium would have been achieved for the water gas shift reaction [6]. Hence, if partial surface blockage could limit the influx of CO₂, diffusional hindrance to reach the electrode functional layer would have been important for all reaction; i.e. the electrochemical reaction was most likely to have been H₂O reduction alone by reaction (1b).

With CO₂ being blocked partially from entering pores, the water gas shift reaction would then have been confined to a decreased number of nickel surface sites, decreasing heterogeneous reaction rates. Homogenous reverse water gas shift reaction within the gas phase has been reported to be slow compared to heterogeneous reaction rates on nickel sites [33,34]; thus, the overall chemical reaction rate would have been smaller with partially blocked surface pore entrances, precluding equilibrium compositions of the reverse water gas shift reaction from being achieved.

Table 5
Parameters used to calculate the adsorption/desorption and surface coverage.

<table>
<thead>
<tr>
<th>Reaction</th>
<th>A (or a)</th>
<th>β (or b)</th>
<th>E_a (or d_a)</th>
<th>kJ mol⁻¹</th>
</tr>
</thead>
<tbody>
<tr>
<td>H₂ + Ni(s) + Ni(0) → H₂O + Ni(0)</td>
<td>5.593 × 10⁻¹⁵</td>
<td>0.0</td>
<td>88.12</td>
<td></td>
</tr>
<tr>
<td>H₂O + Ni(s) → H₂O(s)</td>
<td>4.579 × 10⁻¹²</td>
<td>0.0</td>
<td>62.68</td>
<td></td>
</tr>
<tr>
<td>CO₂ + Ni(0) → CO(2g)</td>
<td>9.334 × 10⁻⁷</td>
<td>0.0</td>
<td>28.80</td>
<td></td>
</tr>
<tr>
<td>CO + Ni(0) → CO₂</td>
<td>4.041 × 10⁻¹⁵</td>
<td>0.0</td>
<td>112.85</td>
<td></td>
</tr>
<tr>
<td>Additional surface coverage dependence of E_a (gCO):</td>
<td></td>
<td></td>
<td>50.0</td>
<td></td>
</tr>
<tr>
<td>H₂O + H₂(g) → Ni(0) + H₂O</td>
<td>1 × 10⁻¹⁷</td>
<td>0.0</td>
<td>0.0</td>
<td></td>
</tr>
<tr>
<td>H₂O(g) + H₂ + H₂O</td>
<td>1 × 10⁻¹⁴</td>
<td>0.0</td>
<td>0.0</td>
<td></td>
</tr>
<tr>
<td>CO(2g) + Ni(0) + CO₂</td>
<td>1 × 10⁻⁵</td>
<td>0.0</td>
<td>0.0</td>
<td></td>
</tr>
<tr>
<td>CO(3g) → Ni(0) + CO</td>
<td>5 × 10⁻¹</td>
<td>0.0</td>
<td>0.0</td>
<td></td>
</tr>
</tbody>
</table>

Site occupancy number (σ_s): 1 [–]
Surface site density (Γ): 2.60 × 10⁻⁵ mol m⁻²

Fig. 12. (A) Equilibrium adsorption rates and (B) surface coverage of H₂O, H₂, CO₂ and CO on Ni for feed gas scenario 1 (–), 2 (– –) and 3 (– –).
Fig. 13. Schematic of experiment using C\textsuperscript{18}O\textsubscript{2} to distinguish between direct electrochemical reduction of CO\textsubscript{2} and its (indirect) chemical with electrochemically-generated H\textsubscript{2} by the reverse water gas shift reaction (6).

Fig. 14. Mass spectrometric intensities measured at the anode for isotopic oxygen fractions (36–18O\textsubscript{2}, 34–16O\textsubscript{18}O and 32–16O\textsubscript{2}) at different cathodic feed gas compositions and electrolysis current densities at 822° C operating temperature.
Table 6

| Oxygen 16 and 18 fractions for different feed gas compositions and current densities at 822 °C. |
|------------------------------------------------|-------------|----------------|----------------|
| Current density/A cm⁻² | O–16 | O–18 |
| C¹⁶O₂/C¹⁸O | −0.52 | 100% | 0% |
| C¹⁶O₂ alone | −0.26 | 100% | 0% |
| C¹⁸O₂ alone | −0.52 | 100% | 0% |
| C¹⁸O₂ÈH₂¹⁸O | −0.52 | 14% | 86% |
| C¹⁸O₂ alone | −0.52 | 53% | 47% |
| C¹⁸O₂ alone | −0.52 | 17% | 83% |

3.2.3.3. Evidence 3: Gas adsorption. Fig. 12(A) shows the equilibrium adsorption/desorption rates of the four gas feed species in their respective gas feed scenarios¹, and calculated by solving simultaneously the adsorption and desorption Eq. (11)–(17) using data in Table 5 [35,36,37].

\[ \text{rate}_{\text{ads}} = k_{a,i} |c_i| |c_N|^{\text{sites}} \]  
\[ k_{a,i} = \frac{S^0_i}{(T_{\text{total}}/T)} \sqrt{\frac{RT}{2\pi M_i}} \]  
\[ S^0_i = a_i T^b \exp \left( \frac{d_i}{RT} \right) \]  
\[ c_N^{\text{sites}} = \frac{(1 - \sum \theta_k) \Gamma}{\alpha_k} \]  
\[ \text{rate}_{\text{des},i} = k_{d,i} |c_{surf,i}|^{\text{active}} \]  
\[ k_{d,i} = A_i T^b \exp \left( \frac{E_{a,i}}{RT} \right) \exp \left( \frac{\theta_k \Gamma}{RT} \right) \]  
\[ c_{surf,i} = \frac{\theta_k \Gamma}{\alpha_i} \]  

In co-electrolysis, H₂O adsorption was predicted to be four orders of magnitude faster than CO₂ adsorption. Thus, it seems plausible that steam electrolysis was the dominant electrochemical process, whereby carbon monoxide was produced mainly via the reverse water gas shift reaction (6), especially in view of the additional evidence from gas diffusion and gas compositional changes. Furthermore, with CO responsible for >30% surface site coverage, blocking active reaction sites (>41% total coverage in co-electrolysis at 822 °C, compared with 22% total coverage for electrolysis of H₂O alone), the preferred strategy could be to use steam electrolysis followed by indirect CO₂ reduction for syngas production. This would also enable the syngas composition to be tuned better to the requirements of downstream processes.

3.2.4. Labelled C¹⁸O₂ÈH₂¹⁸O electrolysis

In order to distinguish further between the relative rates and product distributions of electrochemical versus chemical reactions within the hollow fibre reactors (Fig. 13), the experiments described above were repeated, but with C¹⁸O₂ replaced by 95.5% oxygen-18 enriched isotopically-labelled C¹⁸O₂. Both anode and cathode off-gas streams were monitored separately using the same model of mass spectrometer. In order to limit the possibility of the water gas shift reaction exchanging labelled oxygen prior to the reactor, H₂O and CO₂ alone were fed to the reactor at 50:50 molar ratio with a total active gas flow of 40 ml min⁻¹, comparable to gas flow scenario 3 in Table 1, while a cathodic current of −0.026 A cm⁻² was applied to the cell to protect the nickel from re-oxidation.

Fig. 14 shows an exemplar result of mass spectrometric analyses of the anode off-gases at different operating conditions, highlighting that the labelled oxygen could indeed be distinguished. Assuming that the labelled oxygen signal (32, 34 and 36) intensities would scale similarly regardless of the isotope, the isotopic oxygen fractions from the anode could be calculated (Table 6). Note that using the mass spectrometric analysis of the cathode off-gas, the CO₂ fractions were 83% (C¹⁶O₂), 15% (C¹⁸O¹⁸O) and 2% (C¹⁸O₂) %, implying a 90.5% isotopic enrichment in contrast to the 95.5% certified. Thus, the signal intensities were not exactly the same for the different isotopes, but within 5% error. Hence, the data for C¹⁸O₂ alone in Table 6 were reasonable in comparison to

¹ Note: the feed gas composition for co-electrolysis (scenario 3), was allowed to reach the (reverse) water gas shift equilibrium composition at the respective temperature. These equilibrium compositions were calculated for reaction (6) using thermodynamic data from [4,5].
the ‘measured’ enrichment, allowing for some back-diffusion of oxygen-16 into the YSZ membrane, as the anode sweep gas contained 10% 16O2, and/or residual oxygen-16 within the YSZ membrane. The results for co-electrolysis imply that both CO2 and H2O participated equally in the electrochemical process within the error of the isotopic mass spectrometry analysis. However, it was evident from analyses of the cathode off-gases (not shown here), that, unfortunately, the isotopic oxygen labels were being exchanged between CO2 and H2O i.e. oxygen-18 enrichment in the steam and oxygen-16 replacement in the carbon dioxide, so the anode off-gas isotopic fractions were meaningless.

The first cause of the isotopic exchange would be associated with the water gas shift reaction. However, the H2/CO yields were similar to what has been reported in Fig. 11; i.e. the (reverse) water gas shift (WGS) equilibrium composition was achieved for cells with cathode current collectors of Ni wire alone, whereas cells with the inner lumen of the hollow fibre partially covered in silver paste produced more H2. Hence, it would appear bizarre if the (reverse) WGS reaction was responsible for exchanging labelled oxygen, but did not reach equilibrium, especially as the ratio of H2O/CO2 compared to products (H2/CO) was in excess of 19:1.

To elucidate the reasons for these observations, the hollow fibre reactor was replaced by an alumina tube and C18O2/H2 (scenario A), H2/H18O/C18O2 (scenario B) and H2/H18O/C18O2 (scenario C) at different compositions were fed through the alumina tube at the operating temperature of 822°C. Neither scenario A nor scenario B (Supplementary data: Fig. S1 A–C), representing the left and right hand-side of the reverse water gas shift reaction (reaction 6), respectively, showed any marked variation in the gas composition. Hence, in the absence of nickel catalyst, the homogeneous gas phase (reverse) water gas shift reaction rates were negligible. However, the C18O2 and H2/18O (scenario C) exchanged oxygen-16 and oxygen-18 immediately, though the total amount of carbon dioxide and steam remained the same; i.e. any (intermediate) product was unstable (Fig. 15 A).

On repeating this experiment with a quartz tube instead of an alumina tube, no exchange reaction occurred in any of the three scenarios. Hence, it can be concluded that the exchange of labelled oxygen occurred via adsorbed carbon dioxide and water on the alumina. Unfortunately, using a quartz tube in the reactor system was precluded, as the thermal expansion coefficients [38] of alumina and zirconia (ca. 9 × 10−6 compared to 10 × 10−6 °C−1) were better matched than zirconia with quartz (ca. 0.7 × 10−6 °C−1), so sealing was more facile in the former case. Extending this investigation to a thick YSZ tube, resulted in even more 18O-18 swapped labels, as the oxygen could self-diffuse via vacancies into the lattice as well, replenishing the surface O–16 for further exchange; hence, only a small initial decrease resulted following the introduction of water, as there was another fast process allowing isotope exchange. The results for YSZ indicated a fast surface-exchange of CO2; a detailed study of the self-diffusion of 16O2 ions from C18O2 into YSZ has been reported elsewhere [39].

4. Conclusions

CO2 and/or H2O were reduced to CO/H2 in micro-tubular solid oxide electrolyser systems with 26 μm thick yttria-stabilized zirconia (YSZ) electrolyte, Ni-YSZ cermet cathode and strontium-doped lanthanum manganite (LSM) oxygen-evolving anode. With a silver wire plus silver paste cathode current collector and a potential difference of 1.5 V, current densities for splitting H2O alone and CO2 alone were ~0.46 and ~0.24 A cm−2, respectively. By contrast, with a nickel wire cathode current collector without any paste, current densities increased by 63% to ~0.76 A cm−2 for H2O electrolysis at pH2O = 0.36 × 105 Pa, but the difference between the performances of H2O and CO2 electrolyses decreased to only 16%. SEM images and impedance spectra strongly suggested that silver paste as a component of the cathode current collector caused partial blocking of surface pores in the cathode lumen. This inhibited gas diffusion rates and exacerbated differences in current densities between CO2 and H2O electrolyses at fixed potential differences, due to the smaller diffusion coefficient of CO2 compared to that of H2O.

Co-electrolysis was found to proceed mainly via the reverse water gas shift reaction between CO2 and the electro-generated H2. The supporting evidence included similar electrochemical performances of (CO2 + H2O) co-electrolysis and H2O splitting. Faster gas diffusion for H2O, the reactor microstructure, compositional off-gas analysis and adsorption/desorption modelling of reactant and product gases onto cathode materials.

Unfortunately, when isotopically-labelled C18O2 was used in an attempt to further confirm this hypothesis, isotopic oxygen labels were exchanged between C18O2 and H2/18O, enriching oxygen-18 in the steam and substituting oxygen-16 into the carbon dioxide, so the anode off-gas isotopic fractions were meaningless. This occurred even in alumina or YSZ tubes without the micro-tubular reactor; i.e. even without a Ni catalyst, though not in a quartz tube. However, mismatched thermal expansion coefficients of quartz with the micro-tubular reactors precluded its use as a feed-tube material. However, valuable insight into the behaviour of CO2 on ceramic materials was obtained, including fast surface-exchange of CO2 in YSZ.

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Appendix A. Supplementary data

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References
