Supplementary Information

Table SI1: List of literature cost estimates used in our study. Values are scaled to a standardised capacity using the six-tenths rules for ease of comparison.

Process	Capacity [units]	Investment Cost (M GBP_{2017})	Scaled capacity [units]	Scaled cost (M GBP_{2017})	Ref
Electrolysis	36 [MW]	28	500 [MW_{LHV}]	136	1
	10 [MW]	6		67	2
	10 [MW]	12		123	2
	44 [MW]	39		167	3
	1 [MW]	1.2		51	3
	200 [MW]	76.3		132	3
	1270 [MW]	1828		1045	4
DAC*	1 [Mt]	860	1 [Mt]	UB: 860	5
	1 [Mt]	590		LB: 590	5
Methanation	311 [$m^3\ h^{-1}{}_{CH4}$]	0.61	10^5 [$m^3\ h^{-1}{}_{CH4}$]	20	1
	216000 [$m^3\ h^{-1}{}_{CH4}$]	3560		570	6
	70 [$m^3\ h^{-1}{}_{CH4}$]	0.48		38	7
H_2 Storage**	2×10^6 [Nm^3]	50	10^5 [Nm^3]	82	8
	1×10^5 [Nm^3]	8		77	1
	5×10^4 [Nm^3]	5		68	1
	4×10^7 [Nm^3]	380		108	9
	5×10^7 [Nm^3]	390		91	9
	4×10^7 [Nm^3]	390		107	9
	4×10^8 [Nm^3]	340		23	4
CH_4 Storage	$1x10^8$ [Nm^3]	48	10^5 [Nm^3]	8	10
	$1x10^8$ [Nm^3]	69		11	10
	$1x10^8$ [Nm^3]	21		3	10
	$5x10^7$ [Nm^3]	42		10	10
	$3x10^7$ [Nm^3]	8		3	11
	$3x10^8$ [Nm^3]	69		6	11
HGCC	16 [MW]	53	500 [MW_e]	418	12
CCGT	702 [MW]	530	500 [MW_e]	430	13
	1 [MW]	0.6		320	14
	507 [MW]	305		300	15
	560 [MW]	261		240	15
	550 [MW]	277		260	15
	543 [MW]	330		310	15
	431 [MW]	180		200	16
	630 [MW]	326		280	17
CCGT-CCS	1000 [MW]	1460	500 [MW_e]	960	14
	432 [MW]	420		460	15

	482 [MW]	470		490	15
	467.5 [MW]	400		420	15
	482[MW]	510		520	15
	559 [MW]	650		600	17
SMR	1.2×10^5 [kg_{H2} d^{-1}]	86	1×10^5 [kg_{H2} d^{-1}]	75	18
	6.3×10^5 [kg_{H2} d^{-1}]	190		62	19
	1.15×10^2 [kg_{H2} d^{-1}]	0.3		16	20
	1.2×10^6 [kg_{H2} d^{-1}]	470		105	21
	2.4×10^4 [kg_{H2} d^{-1}]	20		48	21
	4.8×10^2 [kg_{H2} d^{-1}]	0.9		22	21
SMR-CCS	1.2×10^5 [kg_{H2} d^{-1}]	111***	1×10^5 [kg_{H2} d^{-1}]	97	18
	6.3×10^5 [kg_{H2} d^{-1}]	220		73	19
	5.16×10^5 [kg_{H2} d^{-1}]	320		120	22
	2.32×10^5 [kg_{H2} d^{-1}]	260		160	23
	1.2×10^6 [kg_{H2} d^{-1}]	590		130	21
	2.4×10^4 [kg_{H2} d^{-1}]	28		65	21
	9.13×10^5 [kg_{H2} d^{-1}]	1082		290	4
	6.1×10^5 [kg_{H2} d^{-1}]	460		190	24
ATR	2.14×10^5 [kg_{H2} d^{-1}]	23	1×10^5 [kg_{H2} d^{-1}]	14	25
	1.5×10^5 [kg_{H2} d^{-1}]	1.4		1	26
	1.15×10^2 [kg_{H2} d^{-1}]	0.2		14	20
	1×10^2 [kg_{H2} d^{-1}]	0.3		18	27
	1×10^6 [kg_{H2} d^{-1}]	71		18	27
ATR-CCS	2.32×10^5 [kg_{H2} d^{-1}]	200	1×10^5 [kg_{H2} d^{-1}]	120	23
	9.13×10^5 [kg_{H2} d^{-1}]	950		250	4
	6.1×10^2 [kg_{H2} d^{-1}]	580		150	24

**CCS and DAC estimates include compression costs*

***Capacity is calculated working capacity assuming 30% of physical cavern size is for cushion gas*

****This value assumes 30% increase over regular SMR as prescribed by the corresponding reference*

Table SI2: Breakdown of operating cost assumptions based on Coulson and Richardson.[28]

Variable Costs	
Raw materials	Calculated from Aspen flow sheet
Miscellaneous materials	10% of Maintenance cost
Utilities	Calculated from Aspen flow sheet using data from Table SI3
Fixed Cost	
Maintenance	5% of fixed capital
Operating labour	Assumes pay of 30 GBP_{2017} per hour pay for 5 shift crews with 3 shifts a day and 50% extra for allowances and overheads.
Laboratory costs	20% of operating labour
Supervision	20% of operating labour
Plant overheads	50% of operating labour
Capital charges	10% of fixed capital
Insurance	1% of fixed capital
Local Taxes	2% of fixed capital
Royalties	1% of fixed capital

Table SI3: Price of the main utilities

Utility	Cost (GBP_{2017})	Units	Ref
Electricity	0.0939	/kWh	[29]
Process water	0.93	/t	[30]
Cooling water	0.014	/t	[28]

Table SI4: Simplified stream table for Case 2 (DAC + Electrolysis to produce SNG for storage and subsequent combustion for power).

		Electrolyser Feed	H2 leaving electrolyser	DAC feed	CO2 captured by DAC	Sabatier feed	Sabatier exit	Salt cavern feed	SNG into CCGT	Air in	Final Flue gas
Temperature	K	298.00	333.00	298.00	318.00	330.08	573.00	323.15	323.15	298.00	298.00
Pressure	bar	1.01	30.00	1.01	1.01	1.01	20.00	100.00	100.00	1.01	1.01
Mole Flows	kmol/sec	4.28	4.28	3588.44	1.07	5.35	3.24	1.13	1.13	12.31	15.98
CO	kmol/sec	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2	kmol/sec	0.00	0.00	1.44	1.07	1.07	0.02	0.02	0.02	0.00	1.07
H2	kmol/sec	0.00	4.28	0.00	0.00	4.28	0.06	0.06	0.06	0.00	0.85
CH4	kmol/sec	0.00	0.00	0.00	0.00	0.00	1.05	1.05	1.05	0.00	0.00
O2	kmol/sec	0.00	0.00	751.60	0.00	0.00	0.00	0.00	0.00	2.58	3.00
H2O	kmol/sec	4.28	0.00	0.00	0.00	0.00	2.11	0.00	0.00	0.00	1.32
N2	kmol/sec	0.00	0.00	2801.89	0.00	0.00	0.00	0.00	0.00	9.62	9.62
AR	kmol/sec	0.00	0.00	33.52	0.00	0.00	0.00	0.00	0.00	0.12	0.12
Mass Flows	kg/sec	77.06	8.62	103943.02	47.06	55.68	55.68	17.73	17.73	356.71	442.87
CO	kg/sec	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO2	kg/sec	0.00	0.00	63.17	47.06	47.06	0.70	0.70	0.70	0.22	47.28
H2	kg/sec	0.00	8.62	0.00	0.00	8.62	0.13	0.13	0.13	0.00	1.71
CH4	kg/sec	0.00	0.00	0.00	0.00	0.00	16.90	16.90	16.90	0.00	0.00
O2	kg/sec	0.00	0.00	24050.26	0.00	0.00	0.00	0.00	0.00	82.53	96.11
H2O	kg/sec	77.06	0.00	0.00	0.00	0.00	37.95	0.00	0.00	0.00	23.82
N2	kg/sec	0.00	0.00	78490.69	0.00	0.00	0.00	0.00	0.00	269.36	269.36
AR	kg/sec	0.00	0.00	1338.90	0.00	0.00	0.00	0.00	0.00	4.59	4.59

Operating & Maintenance (O&M) cost calculations

There are two components of the O&M costs: variable and fixed. To calculate fixed costs such as capital charges, insurance, local taxes and royalties, the investment (capital) costs are multiplied by the numbers reported in the "Fixed Costs" sub-section in Table SI2. The Laboratory, supervision and plant overheads are calculated as a fraction of the operating labour.

The variable costs, on the other hand, were calculated depending on the technology.

Electrolysis: We assume the only operating costs are water and electricity. The cost of water is determined by the product of the process water price (see Table SI3) and the equivalent amount of water required to produce the H_2. The electricity cost is a product of the electricity price and power requirements for the electrolyser.

DAC: We assume that the operating costs are mainly associated with the adsorbent make up, electricity to run equipment and caustic solution for adsorbent regeneration. We assume an adsorbent purge and make up rate of 3% and a solvent make up rate of 1%. The solids purge rate is similar to that of other sorbent looping processes[31, 32] while the 1% solvent loss is used considering the fact that sodium hydroxide has low volatility. The energy requirements for the equipment (absorber fans, liquid pump, precipitator and slaker, ASU and calciner) were all taken from Keith et al.[5] and scaled according to the total CO_2 captured.

PCC: For the Case scenarios which include post-combustion CO_2 capture (PCC), we assume the operating costs come from the CCS solvent make up, the steam used to strip the CO_2 and the energy required to regenerate the solvent. The reboiler duty[33] was taken to be 3395 kJ/kg_{CO2} while the solvent purge rate was assumed to be 2%.[34]

CCGT/HTCC: For the turbines, we assume the only operating cost to be water for cooling and air for combustion. Here the average water consumption[16] is multiplied by the water cost in SI3 while the air is costed by multiplying the price of compressed air with the air requirements derived from the mass balances in Table SI4.

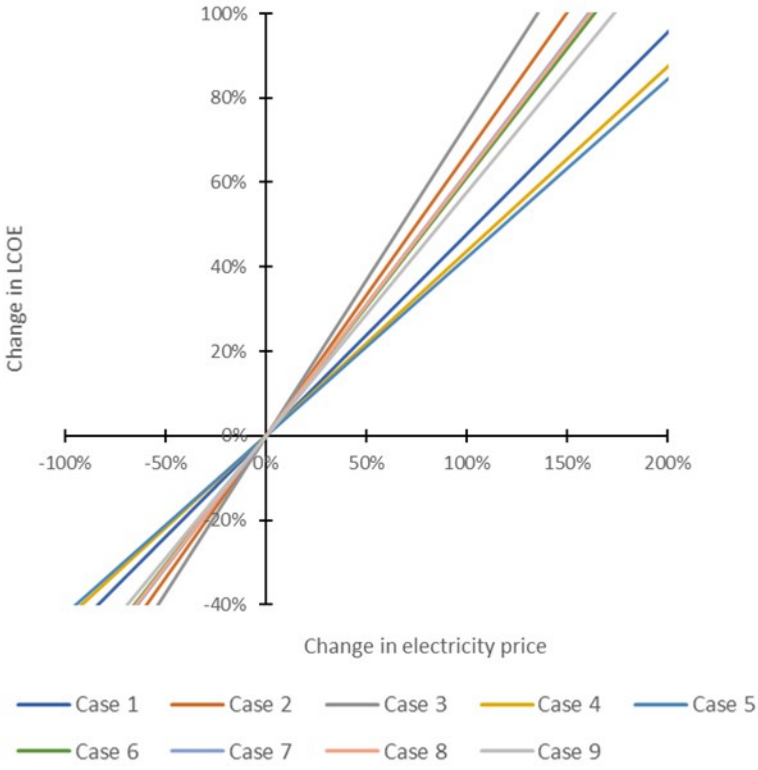

Figure SI1: Sensitivity analysis on the cost of electricity.

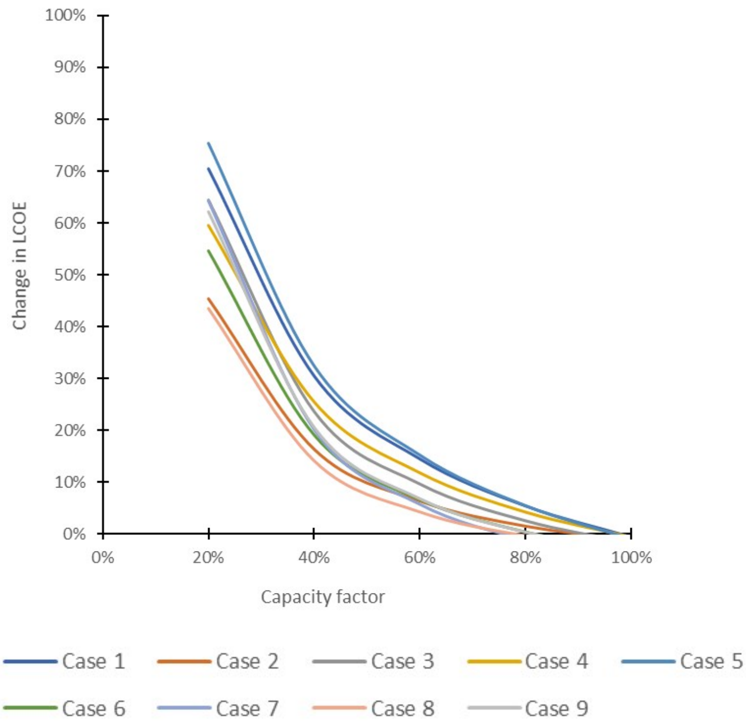

Figure SI2: Sensitivity analysis on the capacity factor

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