

Supplementary Information
for
Economic and Energy Analysis of Capturing CO₂ from the Air

Contents

Table 1: Published air capture analyses	2
Figure 1: The Sherwood Plot, and our addition for Gas Separation Systems.....	3
Figure 2: The 2nd Law Efficiency and the Concentration Factor	5
Calculating Minimum Work	5
Specific Separation Processes in Figure 2.....	6
CO ₂ from flue gas (amine).....	7
N ₂ /O ₂ Separation.....	7
Ethanol Distillation.....	8
CO ₂ from NGCC (amine).....	9
Desalination, reverse osmosis.....	9
CO ₂ from syngas (PSA), CO ₂ from syngas (Selexol), H ₂ clean-up.....	9
References.....	12

Table 1: Published air capture analyses

Here we reproduce Table 1 from the manuscript, and include an additional column that lists assumptions made in the references. These assumptions are used to generate the numbers listed in the “Energy Required” column.

Reference	Steps in Proposed Process	Cost (\$/tonne CO ₂)	Energy Required ^a (kJ/mole CO ₂)	2 nd -Law Efficiency ^b	Assumptions Made in the Reference
Keith, 2006 (1)	NaOH scrubbing, causticization with lime, calcination, amine capture, compression.	\$136	648 thermal & 31 work (total: 736 primary ^c)	7.9%	\$4/GJ natural gas and \$.07/kWh carbon-neutral electricity. 50% capture. The steam for the amine capture system is supplied by the calciner. 50% capture.
Baciocchi, 2006 (2)	NaOH scrubbing, causticization with lime, thermal calcination in a proposed oxy-blown kiln, compression. Option A uses standard technology for the precipitation and dewatering of the CaCO ₃ sludge, Option B uses a prospective pellet reactor.	Not reported.	Option A: 515 thermal & 60 work (total: 686 primary ^c) Option B: 352 thermal & 53 work (total: 503 primary ^c)	A: 10.0% B: 13.5%	75% efficiency of direct utilization of thermal energy, 35% efficiency for electricity generation, all other energy transfer operations assumed to be 100% efficient. Used an oxy-fuel calciner (calcination in pure oxygen rather than air), noting that the technology for this does not yet exist. 500 ppm CO ₂ in air, 50% capture.
Nikulshina, 2006 (3)	Aerosol-type carbonator using Ca(OH) ₂ , solar calciner, conventional slaker. Note that the alternative configuration with H ₂ production is not considered here.	Solar calciner only: \$176-220	2,485 thermal	2.4%	Thermodynamic analysis. With heat exchangers (12,435 kJ/mole without heat exchangers). Heat exchange processes are ideal, no pumping work, all materials are pure, all reactions achieve chemical equilibrium. 500 ppm CO ₂ in air, 100% capture.
Zeman, 2007 (4)	NaOH scrubbing, causticization with lime, thermal calcination in a proposed oxy-blown kiln, compression.	Not reported.	225 thermal & 104 work (total: 522 primary ^c)	11.1%	Assumed 105 kJ/mole thermal energy recovery from the heat of lime hydration with steam. Used a pilot steam drying process for the precipitate. Used an oxy-fuel calciner (calcination in pure oxygen rather than air). Kiln efficiency of 70%. 380 ppm CO ₂ in air, 50% capture.
Stolaroff, 2008 (5)	NaOH spray tower, proposal and prototype. Does not include other aspects of an air capture process.	Proposed spray tower only: \$53-127 ^d	Proposed spray tower only: 53-94 work ^d (150-270 primary ^{c,d}) Prototype spray tower only: 190-390 work (540-1,100 primary ^c)	Insufficient information.	The proposed system is a 20 m high spray tower that captures 30% of the 400 ppm CO ₂ in air. Air velocity of 2 m/s. Costs assume \$.07/kWh carbon-neutral electricity, 15% industrial capital charge rate plus 4% annual rate for maintenance and capital investment. Pump and fan efficiencies are assumed to be 85%.
Mahmoudk hani, 2009 (6)	Recover NaOH from Na ₂ CO ₃ through novel two-stage crystallization/precipitation followed by titanate process. Does not include other aspects of an air capture process.	Not reported.	Recovery cycle only: 150 enthalpy change	Insufficient information.	Energy given is ΔH. Uses a novel process to precipitate sodium carbonate from its aqueous solution. Uses a titanate direct causticization process that has been tested at pilot scale but not been applied commercially.
Lackner, 2009 (7)	Filter with CO ₂ selective sorbent (resin). Air exchange, steam flush, compression.	\$220	38 work ^e (110 primary ^{c,e})	50%	Design based on laboratory results of a strong-base ion-exchange resin. Condensation of water and heat from compression provides all thermal needs. 400 ppm CO ₂ in air, 30% capture.

Table S1. Published air capture analyses.

a: does not include energy for CO₂ compression after capture for transport or storage purposes.

b: calculated by us based on the data provided in each publication.

c: assuming a 35% efficiency of converting primary energy to work.

d: depends on the size of the tower and thus is the result of an optimization calculation.

e: energy for compression of CO₂ to atmospheric pressure included for consistency, approximate energy for further compression removed (12 kJ/mol).

Figure 1: The Sherwood Plot, and our addition for Gas Separation Systems

As mentioned in the text; most points on a Sherwood plot can be connected using a function of the form $P = A/C$ where P is the price in \$/kg of a material of interest that occurs at concentration C in an initial mixture which costs about A \$/kg_i to process. Although this approach is adequate when fitting large groups of data such as the blocks on the plot by Grübler et al. (8), it is not the best possible fit to the data when fewer, but more precise, points are available. To illustrate this point, in the figure below we plot P vs. C for a number of processes that have been modeled extensively with IECM, including SO_x, NO_x, and Hg removal from flue gases (9).

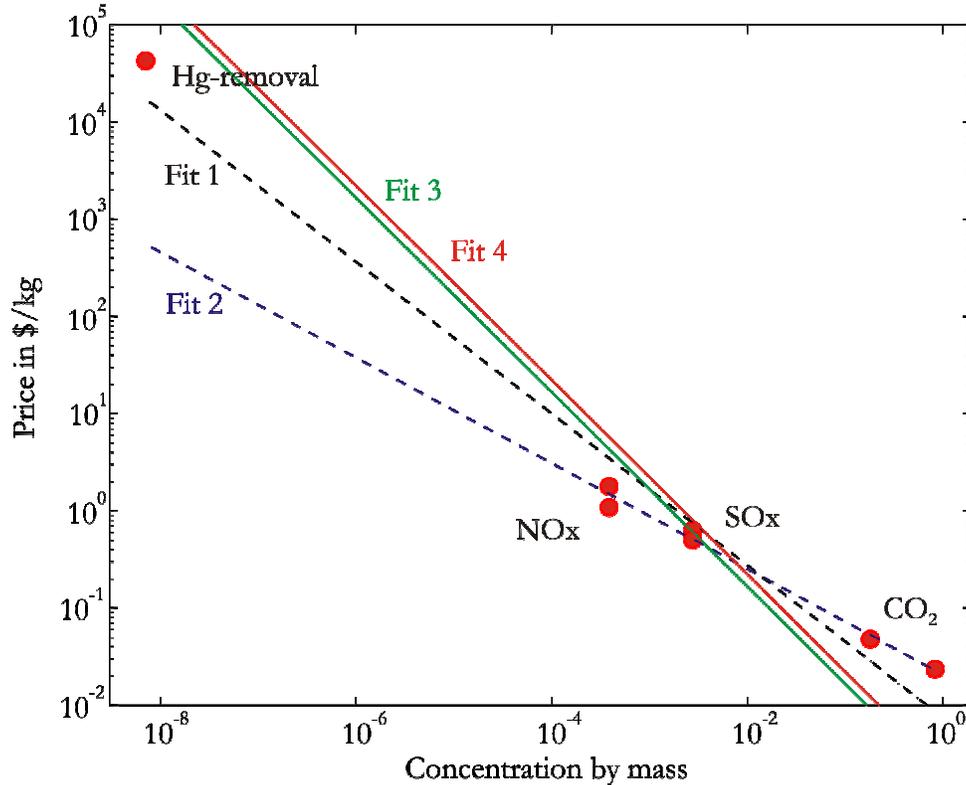


Figure S1: Gas separation Sherwood plot showing cost to separate Hg, NO_x, SO_x, and CO₂ from flue gas at their respective concentrations. Fits 1 and 2 are power law fits; fit 2 does not include the mercury removal. Fits 3 and 4 are of the form $P = A/C$, where A is the only fitting parameter; fit 4 does not include mercury removal. See Table SM.2 for more details.

Table S2 indicates the different possible fits, their respective fitting coefficients, and what such fits mean for the extrapolated cost of air capture. Most fits produce excessive projected costs for air capture, while fit nr. 2 appears to be the most sensible fit to use to get a reasonable estimate of probable costs of air capture. This fit suggests costs of around \$1150 / tCO₂.

		Cost of air capture according to this fit (\$/tCO₂)
Fit 1 (with Hg-removal)	$P = 0.0075 (1/C)^{0.7812}$	2444
Fit 2 (without Hg-removal)	$P = 0.0208 (1/C)^{0.5434}$	1161
Fit 3 (with Hg-removal)	$P = 0.0017 / C$	2741
Fit 4 (without Hg-removal)	$P = 0.0022 / C$	3644

Table S2: Fitting functions for Figure SM.1. For each fit, we include a projection of what air capture would cost according to the fitting function, assuming $C_{\text{air}} = 400$ ppm.

Figure 2: The 2nd Law Efficiency and the Concentration Factor

The vertical and horizontal axes in Figure 2 are, respectively, the 2nd-Law Efficiency (η) and the concentration factor (CF). The 2nd-Law Efficiency is defined as the ratio of the theoretical minimum work required to carry out a process (W_{min}) to the actual work required (W_{actual}) to perform the process:

$$\eta = W_{min} / W_{actual} \quad (1)$$

In separation processes, one or more feed mixtures are resolved into two or more product streams (see Figure SM.2). In separation systems that target CO₂, one product has a significantly higher CO₂ concentration than the feed.

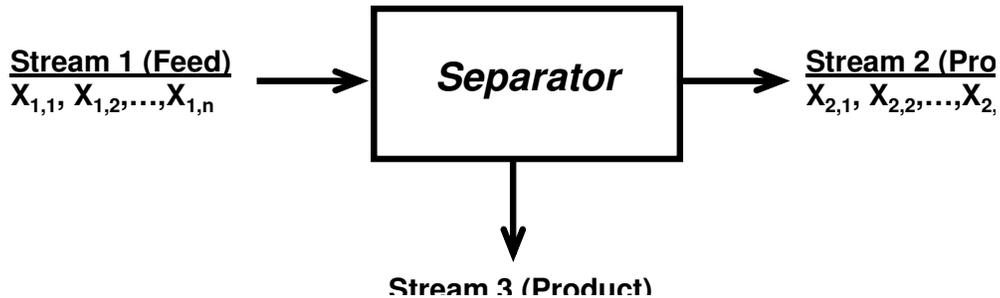


Figure S2: A generic n-component separation process with one feed and two product streams; $X_{j,k}$ denotes the mole fraction of component k in stream j

The concentration factor, CF , is defined here as the ratio of the targeted component's mole fractions in the concentrated product stream and in the feed mixture.¹ For example, for a two-component system in which targeted component 1 is concentrated in stream 3, it follows that:

$$CF \equiv \frac{X_{3,1}}{X_{1,1}} \quad (2)$$

Calculating Minimum Work

The theoretical minimum work required to achieve a change in thermodynamic states is the net change in work potential (i.e., thermodynamic availability or exergy) of the system (10). The change in work potential is minimized when a flowing system undergoes an isothermal, isobaric change, and in that limit, the work potential reduces to the Gibbs Free Energy (11). Therefore, the absolute minimum work required for a given separation processes is equal to the difference between the Gibbs Free Energies (G) of the product and feed streams:

$$W_m = \Delta G = \Delta H - T \Delta S \quad (3)$$

¹ More than one component can be targeted; there is one targeted species in the processes considered here.

In the simple case of a separation process of the type represented in Figure SM.1, in which all streams consist of *ideal mixtures*, this reduces to:

$$W_m = -RT (N_1 \sum_{k=1..n} X_{1,k} \ln X_{1,k} - N_2 \sum_{k=1..n} X_{2,k} \ln X_{2,k} - N_3 \sum_{k=1..n} X_{3,k} \ln X_{3,k}), \quad (4)$$

Where N_j denotes the molar flow rate of stream j .

For non-ideal mixtures (i.e., gases and solutions), however, we must account for the excess properties that depend on interactions between molecules. Since the ideal substance presumes the absence of molecular interactions, the excess properties reflect deviations from ideality, and these deviations are accounted for by including the activity coefficient such that the partial molar Gibbs Free Energy is adjusted (10):

$$G_i - G_i^{ideal} = RT \ln(\gamma_i) \quad (5)$$

Where G_i is the partial molar Gibbs Free Energy of component i in the real mixture, G_i^{ideal} is the partial molar Gibbs Free Energy of component i in the ideal mixture, and γ_i is the activity coefficient of component i . The resulting minimum work is thus:

$$W_m = -RT (N_1 \sum_{k=1..n} X_{1,k} \ln \gamma_{1,k} X_{1,k} - N_2 \sum_{k=1..n} X_{2,k} \ln \gamma_{2,k} X_{2,k} - N_3 \sum_{k=1..n} X_{3,k} \ln \gamma_{3,k} X_{3,k}), \quad (6)$$

In most of the cases considered here, the activity coefficients are close to 1, and thus the mixtures are near ideal. The only case for which an activity factor is included are that of ethanol distillation and desalination.

Specific Separation Processes in Figure 2

Below we describe in detail the data that was used to construct Figure 2 of the manuscript. Table S2 lists the data points in the figure. For each data point in the figure, a set of parameter values were needed:

- (1) The mole fractions of the various components in the input stream,
- (2) The mole fractions of the various components in the output streams,
- (3) The actual work required by the system to perform the separation, and
- (4) the activity coefficients—if necessary—of the mixture's components.

Once those values are known, CF and η , as defined above, can be calculated. In some cases, the separation is accomplished by the direct application of primary energy by heating. In those cases, we apply 35% conversion efficiency between the primary energy content of the fuel used and the potential work that could be produced with that primary energy. The potential work calculated with that efficiency and the primary energy content of the fuel is the work that the applied quantity of primary energy could have been harnessed to perform externally (i.e., work as an 'opportunity cost').

One of the criteria used to screen separation processes for our analysis was that there be a well-defined minimum work requirement. Many separation systems—such as the removal of SO_2 from power-plant flue gas by reaction with $CaCO_3$ —involve spontaneous reactions. In principle, the minimum work for such processes would be negative, meaning that one could harness the

separation process to do external work. In practice, these systems require work, and as such, η is not well defined.

CO₂ from flue gas (amine).

Description. Post-combustion CO₂ capture by scrubbing with amines was examined. An amine solution is sprayed through the flue gas to absorb the CO₂, then heated to desorb the gas and regenerate the amine.

Calculations. The flue gas was assumed to contain 11% CO₂, 90% of which was captured as a pure CO₂ stream (following Herzog, 2009 (12)). For the minimum work calculation, a temperature of 298 K was assumed (again following Herzog, 2009 (12)). The MIT coal report (13) gives a loss of power plant efficiency in both subcritical and ultra supercritical Pulverized Coal power plants from the CO₂ separation process of 5.7% of the primary energy of the coal. Using the higher heating value (HHV) given in the report (25,350 kJ/kg) and the weight fraction of carbon in the coal (0.61), which was all assumed to be converted to CO₂, the energy use per mol of CO₂ captured was calculated. Estimates from Rao (2006) (14) for the thermal and electrical energy consumption of a post-combustion, amine capture with a 90% capture efficiency were taken from their Figure 2 (0.17 MWh/tCO₂ thermal and 0.15 MWh/tCO₂ electrical), with the compression energy subtracted from that calculated from their Equation 3 (0.12 MWh/tCO₂). Additionally, we calculated the energy requirements of a post-combustion carbon capture system using amine scrubbing from Aspen-based IECM software developed by Rubin et al (9).

An example of how the above literature values were used to calculate *CF* and η :

Reference	Stream 1 CO ₂ Concentration	Stream 3 CO ₂ Concentration	CF	Stream 2 CO ₂ Concentration	ln(γ)	W_m (kJ/(mol CO ₂))	W_{actual} (kJ/(mol CO ₂))	η
MIT (2007) (13)	X _{co2} = 11%	X _{co2} = 100%	9.1	X _{co2} = 1.2%	0	7.2	28	25%

N₂/O₂ Separation.

Description. The oxygen and nitrogen in air are separated into pure streams. Most air separation is done through cryogenic distillation, in which the air is cooled then distilled, while other methods include pressure swing adsorption, membrane separation, and chemical separation.

Calculations. From the MIT coal report (2007), the efficiency loss from the air separation unit in an oxy-fuel coal plant is 6.4% of the primary energy of the coal. Assuming feed air that is 21% oxygen is purified to 95% oxygen (as given) with 100% recovery (assumed) yields a concentration factor of 4.5. The same HHV and carbon mass fraction of the coal is used as in the CO₂ from flue gas calculation above to determine the actual energy used per mole CO₂. For the minimum work calculation, a temperature of 298 K was assumed. The Office of Technology Assessment (1993) (15) gives three numbers for air separation technologies, “Current,” “State-of-the-Art (2010),” and “Advanced (2010).” Taking the energy use for the state-of-the-art system and normalizing it to the amount of oxygen produced (42% of that ton) yields the actual energy

required per mole O₂. Singh (2003) (16) reports an air separation unit that produces oxygen of 99.5% at 300 ton/h using 86 MWe.

Ethanol Distillation.

Description. To derive ethanol from corn, the corn is first processed and fermented, resulting in a mixture that is about 10 wt% ethanol (Balat, 2008 (17) and Tiffany, 2003 (18)), with the balance being solids and water. The concentration of ethanol at this step is constrained by the maximum concentration that the yeast can handle (about 10%) and the more efficient use of facilities from a greater concentration of ethanol. The mixture is separated to 99.6% ethanol or more (17), which requires two steps: distillation takes the ethanol to its azeotrope with water (95%), and then a molecular sieve finishes the dehydration process.

In the case of ethanol-water separation, it is important to note that the solutions diverge from ideality sufficiently that the activity of the ethanol/water mixture must be taken into account (19).

Calculations. For the minimum work calculations a binary solution of ethanol and water was assumed, with the initial and final wt% of ethanol at 10% and 100% and a 100% recovery factor. These assumptions yield an initial mole and final mole fraction of 4.1% and 100%, respectively, and therefore a concentration factor of 24. For the initial stream, an activity coefficient of ethanol in water of 3.75 was used (19), while the activity coefficient of the solvent water was assumed to be 1. A temperature of 297 K was assumed. Numbers for the energy intensity of the distillation process were found in Leeper (1992) (20), Hettinga (2007) (21), and Kim (2005) (22). Hettinga (21) and Leeper (20) report the energy required as 3.9 GJ/m³ ethanol and 6.7 GJ/m³ ethanol, respectively, which are assumed to be heat requirements and thus are converted to work requirements assuming a 35% primary energy-to-work efficiency per the discussion above. Kim’s Table 2 (22) reports the energy use of dry and wet milling processes. For both dry and wet milling, the total primary energy use of the four cases reported were averaged, with any electrical contribution assumed to have been produced with the same 35% conversion efficiency. The fraction of energy used for distillation given in Table 3 was then applied to both dry and wet milling energy totals (56.5% and 46%, respectively), and the primary energy converted to work with a 35% efficiency.

The following exemplifies how the above literature values were used to calculate *CF* and η :

Reference	Stream 1 EtOH Mole Fraction	Stream 3 EtOH Mole Fraction	CF	Stream 2 EtOH Mole Fraction	ln(γ) for EtOH	W_m (kJ/(mol EtOH))	W_{actual} (kJ/(mol EtOH))	η
Hettinga (2007) (21)	0.04	1	~24	0	~1.3	~7	~80	~9%
Leeper (1992) (20)	0.04	1	~24	0	~1.3	~7	~140	~5%

CO₂ from NGCC (amine).

Description. Bolland (2003) (23) compared CO₂ capture options for natural gas-fired combined cycle plants. We examined the post-combustion capture of CO₂ with chemical absorption using amine solvents.

Calculations. The initial molar concentration of CO₂ in the exhaust gas was assumed to be 3.6%, the final concentration of the CO₂ stream was assumed to be 100%, and the fraction of CO₂ captured was given (90%). For the minimum work calculation, a temperature of 298 K was assumed. Efficiency losses for the natural gas plant were given, totaling 6.3% of the primary energy of the fuel (not including compression). That primary energy loss was converted to J/mol CO₂ using the natural gas LHV given by Bolland & Undrum (45.5 MJ/kg) and assuming a ratio of mol carbon: mol natural gas of 1.16, which is consistent with the volume % of CH₄ (80%), the molar mass of the fuel (20.6 g/mol), and the LHV given (the composition of the stream chosen to fit those parameters is, by mole: 80% methane; 7% ethane; 3.2% propane; 1.5% butane; 6.8% CO₂; and 1.5% N₂).

Desalination, reverse osmosis.

Description. The reverse osmosis process, accounting for approximately 40% of desalination capacity worldwide (Miller, 2003, (24)), uses a membrane with pressurized feed to separate water from brine. We highlighted reverse osmosis because it is a major commercial desalination process with lower energy consumption than the other widely used distillation processes and it uses electrical power, thus providing a clear comparison between energy consumption and minimum work.

Calculations. Abou Rayan (2002) (25) presents a case study of a desalination plant in Sinai, Egypt. The salt content (wt%) in the feed water, product water, and brine water were 44,000 ppm, 355 ppm, and 52,115 ppm, respectively, which were converted to mol percentages assuming that the salt ions had a molar mass of 29.25 g/mol, the average of the molar masses of Na and Cl. The concentration factors for salt and water were both near 1, at 1.19 and 1.03, respectively, with the concentration factor for water used because it is the desired substance. The activity coefficient for sodium hydroxide in water was taken to be .94 (Hamer, 1972, (26)). For the minimum work calculation, a temperature of 300 K was assumed. Based on the energy consumption given (8.81 kWh/m³ product), a thermodynamic efficiency of 34% was obtained. This agrees with the more efficient side of the range given by Miller (24), who observed that reverse osmosis desalination uses 3-10 times more electrical energy than the thermodynamic minimum.

CO₂ from syngas (PSA), CO₂ from syngas (Selexol), H₂ clean-up.

Description. Hydrogen produced by steam reforming of syngas must be purified to remove CO₂ and other contaminants. Hydrogen purification is commonly performed with a pressure swing absorption (PSA) system, which uses a bed of porous material to selectively adsorb a particular component. If a pure stream of CO₂ is desired from this process, it can be separated through an additional PSA bed that only targets CO₂.

Hydrogen produced by coal gasification, e.g. in an IGCC cycle, may be purified through a Selexol solvent system that uses a physical absorption process.

Calculations.

CO₂ from syngas (PSA): Ogden (1999) (27) reports that the gas produced by the shift reactors contains by volume 77% H₂, 19% CO₂, and 4% other gases (CH₄, CO, trace N₂). Ogden cites Katofsky's (1993) (28) process of CO₂ separation added to hydrogen purification. In that process a separate PSA remove CO₂ prior to the syngas passing through the PSA that purifies hydrogen. The purity of the output CO₂ was assumed to be 100% with a given 70% recovery factor, and the initial flow stream composition assumed to be that given by Ogden. For the minimum work calculation, a temperature of 300 K was assumed. The minimum thermodynamic work is calculated using the same procedure as the other separation systems, however, the actual work must take into account the varying pressures of the input and output streams. Katosky reports that the optimal pressure of the PSA feed is 1.4-2.8 MPa, whereas the CO₂ output stream is at .13 Mpa. The other output stream is assumed to be at the same pressure as the feed. To account for the pressure difference, the separation system is modeled with a constant pressure at the inlet and outlet, with either a compressor and expander to raise and lower the high pressure streams to low pressure, or a compressor to raise the low pressure stream to high pressure. Compression or expansion energy is estimated as the theoretical energy for the isothermal compression of an ideal gas ($R T \ln(P_f/P_i)$), with a 90% assumed efficiency. This compression and expansion energy added to the reported energy use of 100 kWh/tonne CO₂ allows for a calculation of thermodynamic efficiency.

CO₂ from syngas (Selexol): Chen (2005) (29) describes the Selexol process for CO₂ removal from syngas in an IGCC power plant. Chen's Table D-3 and D-16 give the initial and final component molar flow rates, respectively, of the CO₂ stream, while the composition of a third, hydrogen stream was assumed to be the difference of the initial and final CO₂ streams. Table D-17 gives the total energy use of the process (9.98 kW). For the minimum work calculation, a temperature of 298 K was assumed. To account for the pressure change (input stream at 550 psi; CO₂ output stream at 56.5 psi; hydrogen output stream assumed to be at the same pressure as the input stream), the same methodology as above for CO₂ separation with PSA was used.

H₂ clean-up: Larson (2005) (30) describes the production of hydrogen and/or energy from biomass gasification. In the H-MAX-VENT case, hydrogen at a concentration of 90.4% is purified to 99.999% purity in a PSA, at 95% recovery. The energy of the PSA purge gas compressor, given in the reference's Table 3, is given as ~.45 kWh/kg H₂. For the minimum work calculation, a temperature of 298 K was assumed. To account for the pressure change (input stream at 20.3 bar; hydrogen output stream at 19.8 bar; impurities output stream at 1.5 bar), the same methodology as above for CO₂ separation with PSA was used.

Separation Process	Reference	W_{min} (kJ/mol)	W_{actual} (kJ/mol)	η	Initial Conc. (mol %)	Final Conc. (mol %)	Conc. Factor (species)
CO ₂ from flue gas (amine)	IECM Calculations	6.9	35	20%	11%	99%	8.6 (CO ₂)
	MIT (2007) (13)	7.2	28	25%	11%	100%	9.1 (CO ₂)
	Rao (2006) (14)	7.2	32	22%	11%	100%	9.1 (CO ₂)
N ₂ /O ₂ Separation	MIT (2007) (13)	5.3	25	21%	21%	95%	4.5 (O ₂)
	Singh (2003) (16)	5.7	33	17%	21%	100%	4.7 (O ₂)
	Office of Technology Assessment (1993) (15)	6.1	Current: 29 State-of-the-art: 44	Current: 21% State-of-the-art: 13%	21%	100%	4.8 (O ₂)
Ethanol Distillation	Hettinga (2007) (21)	7.0	80	8.8%	4.2%	100%	24 (EtOH)
	Kim (2005) (22)	7.0	Dry: 180 Wet: 140	Dry: 4.0% Wet: 5.0%	4.2%	100%	24 (EtOH)
	Leeper (1992) (20)	7.0	140	5.1%	4.2%	100%	24 (EtOH)
CO ₂ from NGCC (amine).	Bolland (2003) (23)	10	56	18%	3.6%	100%	28 (CO ₂)
Desalination, reverse osmosis.	Abou Rayan (2002) (25)	.19	.57	34%	97%	100%	1.0 (H ₂ O)
CO ₂ from syngas (PSA), CO ₂ from syngas (Selexol), H ₂ clean-up	Chen (2005) (29)	4.1	23-26 ^a	16-17%	33%	97%	2.9 (CO ₂)
	Larson (2005) (30)	.66	4.4-5.8 ^a	11-15%	90%	100%	1.1 (H ₂)
	Ogden (1999) (27)	5.2	24-35 ^{a,b}	15-21%	19%	100%	5.3 (CO ₂)

Table S3: Data Points for Figure 2. All values rounded to two significant figures.

a: range depends on methodology of accounting for the pressure difference..

b: range further depends on ambiguity in the inlet pressure in the reference.

References

1. Keith D, Ha-Duong M, & Stolaroff J (2006) Climate strategy with CO₂ capture from the air. *Climatic Change* 74(1):17-45.
2. Baciocchi R, Storti G, & Mazzotti M (2006) Process design and energy requirements for the capture of carbon dioxide from air. *Chemical Engineering & Processing* 45(12):1047-1058.
3. Nikulshina V, Hirsch D, Mazzotti M, & Steinfeld A (2006) CO₂ capture from air and co-production of H₂ via the Ca (OH)₂-CaCO₃ cycle using concentrated solar power—thermodynamic analysis. *Energy* 31(12):1379-1389.
4. Zeman F (2007) Energy and material balance of CO₂ capture from ambient air. *Environ. Sci. Technol* 41(21):7558-7563.
5. Stolaroff J, Lowry G, & Keith D (2008) Carbon dioxide capture from atmospheric air using sodium hydroxide spray. *Environmental Science & Technology* 42(8):2728-2735.
6. Mahmoudkhani M & Keith D (2009) Low-energy sodium hydroxide recovery for CO₂ capture from atmospheric air—Thermodynamic analysis. *International Journal of Greenhouse Gas Control* 3(4):376-384.
7. Lackner K (2009) Capture of carbon dioxide from ambient air. *The European Physical Journal-Special Topics* 176(1):93-106.
8. Grübler A (1998) *Technology and Global Change* (Cambridge University Press).
9. Wilcox J (Calculated from Integrated Environmental Control Software (IECM). Assuming a 500-MW power plant burning Appalachian Medium Sulfur coal. NO_x reduction involves a Selective Catalytic Reduction unit with a 73.48% removal efficiency; work = 488.31 kJ/mol NO_x. See <http://www.iecm-online.com> for further details and current public version of the IECM (2007) software.
10. Perry R & Green D (2007) *Perry's chemical engineers' handbook* (McGraw-Hill Professional).
11. Fermi E (1936) *Thermodynamics* (Dover Publications, Inc, New York, NY).
12. Herzog H, Meldon J, & Hatton A (2009) Advanced Post-Combustion CO₂ Capture. *Clean Air Task Force*.
13. MIT (2007) The Future of Coal: Options for a Carbon-Constrained World.
14. Rao A & Rubin E (2006) Identifying cost-effective CO₂ control levels for amine-based CO₂ capture systems. *Ind. Eng. Chem. Res* 45(8):2421-2429.
15. OTA (1993) Office of Technology Assessment, Industrial Energy Efficiency. (OTA-E-560, Washington, DC, US Government Printing Office).
16. Singh D, Croiset E, Douglas P, & Douglas M (2003) Techno-economic study of CO₂ capture from an existing coal-fired power plant: MEA scrubbing vs. O₂/CO₂ recycle combustion. *Energy conversion and management* 44(19):3073-3091.
17. Balat M & Balat H (2008) C. z, Progress in bioethanol processing. *Progress in Energy and Combustion Science* 34:551-573.
18. Tiffany D & Eidman V (2003) Factors associated with success of fuel ethanol producers.
19. Sherman S, *et al.* (1996) Compilation and correlation of limiting activity coefficients of nonelectrolytes in water. *Ind. Eng. Chem. Res* 35(4):1044-1058.
20. Leeper S (1992) Membrane separations in the recovery of biofuels and biochemicals: An update review. *Separation and Purification Technology*:99-194.

21. Hettinga W, *et al.* (2007) Technological learning in US ethanol production. *Quantifying Reductions in Production Costs and Energy Use. A Master Thesis, Utrecht University, The Netherlands.*
22. Kim S & Dale B (2005) Environmental aspects of ethanol derived from no-tilled corn grain: nonrenewable energy consumption and greenhouse gas emissions. *Biomass and Bioenergy* 28(5):475-489.
23. Bolland O & Undrum H (2003) A novel methodology for comparing CO₂ capture options for natural gas-fired combined cycle plants. *Advances in Environmental Research* 7(4):901-911.
24. Miller J (2003) Review of water resources and desalination technologies. *Sandia National Laboratories Report, SAND2003-0800.*
25. Abou Rayan M & Khaled I (2003) Seawater desalination by reverse osmosis (case study). *Desalination* 153(1-3):245-252.
26. Hamer W & Wu Y (1972) Osmotic coefficients and mean activity coefficients of uni-univalent electrolytes in water at 25 C. *J. Phys. Chem. Ref. Data* 1(4):1047-1099.
27. Ogden J, Dennis E, Steinbugler M, & Strohhahn J (1999) Hydrogen energy systems studies.
28. Katofsky R (1993) The production of fluid fuels from biomass. (Princeton University).
29. Chen C (2005) A technical and economic assessment of CO₂ capture technology for IGCC power plants. (Carnegie Mellon University).
30. Larson E, Jin H, & Celik F (2005) Gasification-based fuels and electricity production from biomass, without and with carbon capture and storage. *Princeton Environmental Institute, Princeton University.*