Availability analysis of post-combustion carbon capture systems: minimum work input

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Abstract: This paper describes the availability analysis of a generic, post-combustion carbon capture plant. The analysis first establishes the minimum work input required in an ideal plant with a flue gas inlet temperature equal to the sink temperature. The analysis shows that the ideal work input is surprisingly low and that, roughly equal amounts of work are required to first separate and then compress the CO₂ contained in a typical flue gas stream. The analysis is then extended to include the effects of variable inlet temperature and extraction efficiency. This extended analysis shows that there is a considerable quantity of available energy in the flue gas of a normal power station. Indeed, in principle, carbon capture is theoretically possible without any external work input for fuels of low carbon/hydrogen ratio such as heavy fuel oil and natural gas. When burning coal, the minimum work input would be significantly reduced if the flue gases' availability were utilized.

The final section of the paper compares the actual work input of a variety of carbon capture schemes found in the literature, with the minimum work input for an ideal process. This comparison shows that the techniques presently found in the literature have a low second-law efficiency.

Keywords: carbon capture, post-combustion, availability, minimum work input, second-law efficiency

1 INTRODUCTION

Carbon sequestration has become an accepted part of a global effort to reduce man's augmentation of the greenhouse effect to sustainable levels. Substantial research programmes, sponsored by both government and industry, have shown that the technique is both technically and economically feasible and has advantages when compared with many of the alternatives [1]. Current research on carbon sequestration technology concentrates on broadly two main areas: carbon capture and carbon storage. Traditionally the storage technology is not considered critical from a thermodynamic perspective [2]; the process of separating CO₂ from the flue gas and its subsequent liquefaction (or compression) requires larger energy input.

Many commentators state that since the removal of CO₂ from gas streams is an established practice in the chemical industry, the application to the power industry is simply a matter of transferring a mature technology [3, 4]. There is undoubtedly logic in this position, however, the removal of CO₂ from flue gas in a power station still represents a major technical challenge. This is partly due to the low partial pressure of CO₂ in the flue gas, although, there are many techniques to deal with this situation [5]. However, another important factor is the economics of the power industry. The power industry does not have the luxury of producing a high value product. Additionally, it is increasingly one of intense competition and often subject to price shocks in one fuel sector and not another. Conversely, the chemical industry has the advantage that its products are generally of much higher value than energy, so there is less of an imperative for raising efficiency and minimizing capital cost. Additionally, in the chemical industry there is often a requirement for process 'robustness'. This can lead to

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relatively thermally inefficient, complicated, or capital intensive processes being justified.

Carbon sequestration cannot be achieved without a heavily integrated industry with huge investment on fixed equipment. Hence, to make carbon sequestration an industrial reality, rather than an academic curiosity, one of two things is required: Government intervention to reduce the risk of investors or a carbon capture process which is robust, reliable, thermally efficient, and above all cheap (both to build and operate) and hence reducing the risk of investors.

This series of papers, will attempt to analyse from first principles the loss mechanisms inherent in various carbon capture technologies and suggest possible routes for optimization. This paper starts the analysis by attempting to outline a thermodynamically sound basis for comparing the efficiency of different carbon capture cycles. Later papers will extend this analysis to include terms specific to two types of post-combustion carbon capture plant: physical and chemical absorption systems.

2 ANALYSIS OF GENERIC CARBON CAPTURE PROCESS

In this section the minimum work input required to separate CO₂ from flue gas in an ideal post-combustion carbon capture plant will be evaluated. Figure 1 shows a diagram of an idealized scheme producing CO₂ as a compressed liquid. To simplify the analysis it has been assumed that this generic process takes place isothermally. The flue gas is assumed to consist of nitrogen mixed with CO₂ at standard temperature and pressure (25 °C and 1.01 bar). This makes possible the assumption of burning pure carbon with complete combustion of the oxygen content of the air, however, the following analysis will show that this is not a limitation.

The first step in analysing this generic process is to calculate the minimum work input required. This will enable the second law efficiency of actual capture plant to be assessed. Reversible work input for

any steady flow process is the difference in the steady flow availability function of the streams entering and leaving the process [6, 7]. Assuming kinetic and potential energy terms to be negligible and heat rejection to occur reversibly to an infinite reservoir at T_0 , the reversible work is given by

$$W_{\text{rev}} = \Delta H - Q_{0,\text{rev}} = (H_{\text{in}} - H_{\text{out}}) - T_0(S_{\text{in}} - S_{\text{out}}) \quad (1)$$

which on a molar basis reduces to

$$w_{\text{rev}} = (b_{\text{in}} - b_{\text{out}}) = \Delta b \quad (2)$$

For multiple streams, equation (2) becomes

$$w_{\text{rev}} = \sum_{j=1}^J m_{j,\text{in}} \cdot b_{j,\text{in}} - \sum_{k=1}^K m_{k,\text{out}} \cdot b_{k,\text{out}} \quad (3)$$

where j and k refer to individual inlet and outlet streams, respectively.

To calculate the steady flow availability function of a given stream, values for the enthalpy and entropy are required. Carbon capture plant typically has streams entering and leaving that are mixtures of either gases or liquids. Ignoring liquid mixtures at this stage and assuming that ideal gas mixtures are formed, the enthalpy of a mixture of ideal gases on a molar basis is simply the mole fraction weighted average of the constituent's enthalpies, thus [6]

$$h_{\text{mix}}(T) = \sum y_i \cdot h_i(T) + \sum y_i \cdot h_i^0 \quad (4)$$

where h_i^0 is the standard state enthalpy of component i , $h_i(T)$ the molar enthalpy of component i , and y_i the mole fraction of component i .

The molar mixture entropy for a mixture of ideal gases is similar, but includes an extra term representing the ideal entropy of mixing, thus [6]

$$s_{\text{mix}}(T) = \sum y_i \cdot s_i(T) + \sum y_i \cdot s_i^0 + R \left(\sum y_i \cdot \ln \frac{1}{y_i} - \ln p_r \right) \quad (5)$$

where $p_r = p/p^0$, p the system pressure, p^0 the standard state pressure = 101.3 kPa, s_i^0 the standard state entropy of component i , and $s_i(T)$ the molar entropy at standard pressure of component i .

Combining equations (4) and (5) yields an expression for the molar steady flow availability function of

Fig. 1 Model of generic carbon capture process

an ideal mixture of ideal gases

$$b_{\text{mix}} = \sum y_i \cdot h_i(T) - T_0 \left[\sum y_i \cdot s_i(T) + \Re \left(\sum y_i \cdot \ln \frac{1}{y_i} - \ln p_r \right) \right] \quad (6)$$

The terms for standard state enthalpy and entropy in equations (4) and (5) have been ignored as these cancel in steady flow problems with no chemical reaction. Thus, for the generic carbon capture process shown in Fig. 1, the steady flow availability function of the incoming flue gas, per mole of CO₂ entering the capture process is given by (dropping the (T) on the entropies and enthalpies)

$$b_{\text{in}} = h_{\text{CO}_{2,\text{(g)}}} + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} h_{\text{N}_{2,\text{(g)}}} - T_0 \left[s_{\text{CO}_{2,\text{(g)}}} + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} s_{\text{N}_{2,\text{(g)}}} + \Re \left\{ \ln \left(\frac{1}{y_{\text{CO}_2}} \right) + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} \ln \left(\frac{1}{(1 - y_{\text{CO}_2})} \right) - \frac{\ln p_r}{y_{\text{CO}_2}} \right\} \right] \quad (7)$$

where y_{CO_2} is the initial mole fraction of CO₂ in the flue gas stream.

The sum of the steady flow availability function of the two outlet streams is simpler to evaluate since, due to the complete separation of CO₂ from the flue gas, there is no entropy of mixing term. In the following equation (8), the effect of pressure on the enthalpy and entropy of liquid CO₂ has been assumed to be negligible. However, for completeness the enthalpy and entropy of liquid CO₂ can be evaluated at the saturation pressure of CO₂ if desired.

$$b_{\text{out}} = h_{\text{CO}_{2,\text{(l)}}} + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} h_{\text{N}_{2,\text{(g)}}} - T_0 \left(s_{\text{CO}_{2,\text{(l)}}} + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} (s_{\text{N}_{2,\text{(g)}}} - \Re \ln p_r) \right) \quad (8)$$

Substituting equations (7) and (8) into equation (2) gives an expression for the reversible work of our generic process

$$w_{\text{rev}} = (h_{\text{CO}_{2,\text{(g)}}} - T_0 s_{\text{CO}_{2,\text{(g)}}}) p_{\text{in}} - (h_{\text{CO}_{2,\text{(l)}}} - T_0 s_{\text{CO}_{2,\text{(l)}}}) p_{\text{sat,out}} - T_0 \Re \left\{ \ln \left(\frac{1}{y_{\text{CO}_2}} \right) + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} \times \ln \left(\frac{1}{(1 - y_{\text{CO}_2})} \right) - \ln p_r \right\} \quad (9)$$

This result is notable as it only includes terms relating to CO₂ and thus shows that the reversible work for 100 per cent CO₂ separation is dependent only on

the initial molar concentration of CO₂ (assuming ideal gases and mixtures). Values of the flue gases' enthalpy and entropy are therefore not required, nor is the exact composition if the fuel's composition, combustion efficiency, and molar flowrates of the fuel and air entering the power station are known, since these can be used to evaluate the CO₂ concentration of the flue gas. A complication is the presence of other condensable gases such as water, but this will be considered in section 3.2.

Equation (9) can be split into two terms that represent the work required in two separate operations: the work to overcome the entropy of mixing and the work required to liquefy the CO₂, respectively

$$w_{\text{rev}} = w_{\text{rev},1} + w_{\text{rev},2} \quad (10)$$

where, on a mass basis, the work to overcome the entropy of mixing between CO₂ and the flue gas is given by

$$W_{\text{rev},1} = \frac{-T_0 \Re}{M_{\text{CO}_2}} \left\{ \ln \left(\frac{1}{y_{\text{CO}_2}} \right) + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} \times \ln \left(\frac{1}{(1 - y_{\text{CO}_2})} \right) - \ln p_r \right\} \quad (11)$$

This work input rises from zero at 100 per cent CO₂ concentration, to infinity at 0 per cent CO₂ concentration.

The second term for the work to liquefy the CO₂ is given by (on a mass basis)

$$W_{\text{rev},2} = (H_{\text{CO}_{2,\text{(g)}}} - T_0 S_{\text{CO}_{2,\text{(g)}}}) p_{\text{in}} - (H_{\text{CO}_{2,\text{(l)}}} - T_0 S_{\text{CO}_{2,\text{(l)}}}) p_{\text{sat,out}} \quad (12)$$

This work input remains constant and for the conditions specified has the value

$$W_{\text{rev},2} = -211.5 \text{ kJ/kg CO}_2$$

Figure 2 shows the combined minimum work inputs, per kg of captured CO₂, of the first two operations, as functions of the initial CO₂ mole fraction in the flue gas. The graph illustrates the relatively small amount of work required to achieve carbon capture in an ideal plant. Indeed, taking a typical coal burning power station, burning a coal with a gross heating value of 28 MJ/kg and a flue gas CO₂ concentration of 11 per cent, the resulting reduction in station output is a manageable 1.34 percentage points.

A further simplification to the analysis can be achieved by splitting the liquefaction operation into two further operations (J. H. Horlock, 2005, personal communication), as shown in Fig. 3. Overall minimum

Fig. 2 Minimum work input in ideal post-combustion carbon capture cycle with 100 per cent capture efficiency

Fig. 3 Three-operation model of generic carbon capture process

work input for our generic process is then given by

$$W_{\text{rev}} = W_{\text{rev},1} + W_{\text{rev},2,1} + W_{\text{rev},2,2} \quad (13)$$

where operation 2,1 is the isothermal compression of the separated CO_2 to its saturation pressure at T_0

$$W_{\text{rev},2,1} = (H_{\text{CO}_2(\text{g})} - T_0 S_{\text{CO}_2(\text{g})})_{P_{\text{in}}} - (H_{\text{CO}_2(\text{g})} - T_0 S_{\text{CO}_2(\text{g})})_{P_{\text{sat,out}}} \quad (14)$$

and operation 2,2 is the isothermal condensation of the CO_2 at its saturation pressure at T_0

$$W_{\text{rev},2,2} = (H_{\text{CO}_2(\text{g})} - T_0 S_{\text{CO}_2(\text{g})})_{P_{\text{sat,in}}} - (H_{\text{CO}_2(\text{l})} - T_0 S_{\text{CO}_2(\text{l})})_{P_{\text{sat,out}}} \quad (15)$$

Equation (15) must sum to zero since a steady flow, isobaric condensation process involves no shaft work, while the work potential of a heat rejection at T_0 is also zero. As a result, only equation (14) needs to be considered in the calculation of minimum work input. The advantage of these formulae is that there is no need to establish liquid enthalpies and entropies.

The magnitude of the two work input terms, given by equations (11) and (14), are roughly equivalent for CO_2 concentrations found in a typical power station's flue gas. However, separating CO_2 from the flue gas is of greater concern to a station designer than the subsequent compression process. This does not mean that there are not considerable losses during liquefying (or compressing) of many tons of CO_2 vapour per minute, but these losses are independent of the initial concentration of CO_2 in the flue gas and can be considered as a 'fixed cost' in thermodynamic terms. In addition, there is little scope for new ideas in the field of gas compression and condensation. Conversely, many of the methods traditionally used by chemical engineers to perform gas separation, exhibit significant entropy production [8] due to practical considerations. As an example, if we consider a carbon capture plant using a physical absorption process, the 'driving force' for mass transfer of CO_2 from flue gas to absorbent is a difference in the fugacity of CO_2 in the two streams. In principle this fugacity difference (which can be likened to a pressure drop) only needs to be sufficient for the entropy of mixing to be overcome; in practice, to avoid impossibly large contacting equipment, a compromise is made and a larger difference of fugacity is specified. This 'driving' difference in fugacity is in turn indicative of the avoidable part of the entropy production and results in physical absorption systems exhibiting much lower efficiencies in practice than theoretically obtainable. Similar arguments can be used to show that overcoming the entropy of mixing is likely to lead to avoidable irreversibilities in other types of capture plant.

3 GENERIC CAPTURE PROCESS WITH REALISTIC INLET AND OUTLET CONDITIONS

Availability analysis in section 2 assumed that various streams enter or leave the carbon capture plant at sink temperature and that 100 per cent carbon capture is required. It is likely that carbon capture plant will be specified to capture less than 100 per cent of the CO₂ contained in the flue gas. Further, it is usual for flue gas to be passed to any post-combustion scrubbing system at temperature far above ambient. To accommodate these variations in inlet and outlet conditions, further analyses are required.

3.1 Capture efficiency

CO₂ capture efficiency can be defined thus

$$\eta_c = \frac{\text{Moles}_{\text{captured}}}{\text{Moles}_{\text{in}}} \quad (16)$$

Allowing for capture efficiency complicates the availability analysis of our generic process since there will now be an entropy of mixing term for both the flue gas entering and leaving the capture plant. Nonetheless, a similar analysis to that of section 2 can be conducted. The availability of the inlet stream is still given by equation (7), whereas that of the outlet streams is given by

$$\begin{aligned} b_{\text{out}} = & (1 - \eta_c)(h_{\text{CO}_{2,\text{(g)}}} - T_0 s_{\text{CO}_{2,\text{(g)}}}) + \eta_c(h_{\text{CO}_{2,\text{(l)}}} \\ & - T_0 s_{\text{CO}_{2,\text{(l)}}}) + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}}(h_{\text{N}_{2,\text{(g)}}} - T_0 s_{\text{N}_{2,\text{(g)}}}) \\ & - \Re T_0 \left\{ (1 - \eta_c) \cdot \ln \frac{(1 - \eta_c y_{\text{CO}_2})}{(1 - \eta_c) y_{\text{CO}_2}} + \frac{(1 - y_{\text{CO}_2})}{y_{\text{CO}_2}} \right. \\ & \left. \cdot \ln \frac{(1 - \eta_c y_{\text{CO}_2})}{(1 - y_{\text{CO}_2})} - \frac{(1 - \eta_c y_{\text{CO}_2}) \ln p_r}{y_{\text{CO}_2}} \right\} \quad (17) \end{aligned}$$

Combining equations (7) and (17) and collecting terms gives the minimum work input per kg of captured CO₂

$$\begin{aligned} W_{\text{rev,c}} = & \frac{(b_{\text{in}} - b_{\text{out}})}{\eta_c M_{\text{CO}_2}} = (H_{\text{CO}_{2,\text{(g)}}} - H_{\text{CO}_{2,\text{(l)}}}) \\ & - T_0(S_{\text{CO}_{2,\text{(g)}}} - S_{\text{CO}_{2,\text{(l)}}}) - \frac{\Re T_0}{\eta_c M_{\text{CO}_2}} \\ & \times \left\{ \ln \left(\frac{1}{y_{\text{CO}_2}} \right) + \frac{(1 - \eta_c y_{\text{CO}_2})}{y_{\text{CO}_2}} \ln \frac{1}{(1 - \eta_c y_{\text{CO}_2})} \right. \\ & \left. + (1 - \eta_c) \ln(1 - \eta_c) y_{\text{CO}_2} - \eta_c \ln p_r \right\} \quad (18) \end{aligned}$$

Note, that with a capture efficiency of 100 per cent, equation (18) is equivalent to equation (9).

As before, equation (18) can be split into two terms representing, respectively: the minimum work to liquefy the captured CO₂ and the minimum work to

overcome the entropy of mixing. The latter work input is given below by equation (19), which is plotted in Fig. 4 for a number of different capture efficiencies

$$\begin{aligned} W_{\text{mix}} = & -\frac{\Re T_0}{\eta_c M_{\text{CO}_2}} \left\{ \ln \left(\frac{1}{y_{\text{CO}_2}} \right) + \frac{(1 - \eta_c y_{\text{CO}_2})}{y_{\text{CO}_2}} \right. \\ & \times \ln \frac{1}{(1 - \eta_c y_{\text{CO}_2})} + (1 - \eta_c) \\ & \left. \times \ln(1 - \eta_c) y_{\text{CO}_2} - \eta_c \ln p_r \right\} \quad (19) \end{aligned}$$

Figure 4 shows that there is a marked reduction in the required work input as capture efficiency is reduced, though it seems unlikely that this fact can be utilized in any practical way. This effect can be explained by considering the change in the Gibbs function of the various streams and their components. At first sight, a separation process should increase the Gibbs function of both the flue gas and the CO₂ as they are both purified – hence the need for work input. However, although the Gibbs function of any CO₂ extracted is clearly increased, the CO₂ that remains in the flue gas experiences a further dilution process and there is a fall in its Gibbs function. This dilution process need not, however, result in any *lost work*, since it occurs simultaneously with the extraction process – the two processes are coupled and the fall in Gibbs function of the CO₂ left in the flue gas reduces the work input required to extract the remainder. Another way of viewing this, is that the partial pressure of the CO₂ that remains in the flue gas, provides a driving force to help extract the remainder.

Fig. 4 Minimum work input to overcome the entropy of mixing between CO₂ and flue gas versus initial CO₂ mole fraction at different capture efficiencies

3.2 Temperature of flue gas

Flue gases are passed to the chimney of typical power stations at temperatures varying from circa 150 °C in a heavy-oil fired station [9], to circa 80 °C in a natural gas installation [10]. Extra useful work can, in principle, be produced as a result of reducing this discharge temperature prior to admission to the capture plant using some form of thermodynamic cycle. The quantity of available energy in the flue gas is not inconsiderable due to the significant amount of latent heat of the steam it contains, as well as the flue gases' sensible heat. This available energy is normally rejected from power plants with the chimney gases. Nonetheless, it is a requirement of many carbon capture schemes that the flue gas temperature is reduced from its inlet level to a point at which absorbent degradation does not occur [11]. Further, it is not improbable that some form of capture process will be conceived in the future that makes use of this available energy to some extent. Therefore, it was decided that a further analysis was required to take the potential work output into account.

In the analysis in sections 2 and 3.1, the terms for the flue gases' enthalpy cancelled and did not appear in equations (9) or (18). In addition, the only term associated with the flue gases' entropy in these equations, was that of the entropy of mixing and this was expressed purely in terms of the initial mole fraction of CO₂ in the flue gas. However, if the flue gases' inlet temperature is different from the temperature at which the capture process is assumed to take place, these simplifications cannot be made; a more complete analysis is required. To simplify the resulting analysis, the capture plant can be broken into two separate sections: the first, cooling the flue gas to sink temperature, while the second is identical to the generic carbon capture process already analysed. Figure 5 shows the relevant diagram of the separate cooling plant.

The analysis of this process is similar to that of the process shown in Fig. 1. For simplicity, all the water is assumed to exit as condensate and a mixture of N₂, O₂, SO₂, and CO₂ will then pass to the carbon capture plant. Applying equation (3) and assuming ideal gases,

Fig. 5 Model of pre-cooling process

the minimum work per kg of CO₂ entering the plant, is given by

$$W_{\text{rev}} = \sum_i \frac{y_i}{y_{\text{CO}_2}} [(H_{i,(g)} - T_0 S_{i,(g)})_{\text{in}} - (H_{i,(g)} - T_0 S_{i,(g)})_{\text{out}}] + \frac{y_{\text{H}_2\text{O}}}{y_{\text{CO}_2}} [(H_{\text{H}_2\text{O},(g)} - T_0 S_{\text{H}_2\text{O},(g)})_{\text{in}} - (H_{\text{H}_2\text{O},(l)} - T_0 S_{\text{H}_2\text{O},(l)})_{\text{out}}] - \frac{T_0 R}{M_{\text{CO}_2}} \left\{ \left(\frac{1 - y_{\text{H}_2\text{O}}}{y_{\text{CO}_2}} \right) \ln \left(\frac{1}{1 - y_{\text{H}_2\text{O}}} \right) + \frac{y_{\text{H}_2\text{O}}}{y_{\text{CO}_2}} \ln \left(\frac{1}{y_{\text{H}_2\text{O}}} \right) - (1 - y_{\text{H}_2\text{O}}) \ln p_r \right\} \quad (20)$$

Summation in equation (20) includes terms that represent the work extractable due to the sensible heat of the N₂, O₂, SO₂, and CO₂, respectively. The next term represents the work extractable due to the sensible and latent heats of the steam. The last term represents the work required to overcome the entropy of mixing between the steam and the remainder of the flue gas during the condensation process.

To illustrate the use of this equation and also show the influence of a fuel's hydrogen content on plant performance, the minimum work input for a combined flue gas cooling and carbon capture plant has been calculated for three idealized fuels. The fuels are assumed to have the following empirical formulae [12]

Coal: C₁₀₀H₇₆O₁₀N₃
Heavy fuel oil: C₄₄H₆₉S_{0.75}
Natural gas: CH₄

The products of combustion of these fuels burning in dry air are represented by the reaction formulae given below. In each case Z represents the equivalence ratio

Figure 6 shows a plot of the minimum work input calculated for these reference fuels as a function of

Fig. 6 Minimum work input in ideal post-combustion carbon capture cycle with flue gas inlet temperature of 125 °C for three different fuels

the equivalence ratio, with an initial flue gas temperature of 125 °C. The work input is per kg of CO₂ captured. These plots show the aggregate of equations (18) and (20) for different fuels, with a capture efficiency of 90 per cent. For all three fuels, at low equivalence ratio there is the potential for a significant net power output due to the available energy in the flue gas stream. This effect is partly due to the sensible heat of the excess of nitrogen and oxygen that exist in lean combustion systems – a phenomenon well known to the boiler-making industry [13]. However, even under stoichiometric conditions, the work input required to perform carbon capture is considerably less than it would be if the available energy, carried by the flue gas, had not been considered. A significant part of this available energy is due to the latent heat of the steam in the flue gas. As a result, for fuels with low carbon/hydrogen ratio, a net work output is theoretically obtainable at any equivalence ratio.

There are clearly immense practical difficulties in converting the available energy of the flue gas to shaft

work. More likely is that the available energy might be used to provide part of the heat requirement in, say, solvent regeneration. Nonetheless, the analysis indicates the large amount of availability wasted in power stations which discharge hot flue gas straight into the atmosphere and particularly those operating with low overall fuel/air ratio such as combined cycle gas turbine plant.

4 EFFICIENCY OF POST-COMBUSTION CARBON CAPTURE PROCESSES

Availability analysis conducted in preceding sections has established the minimum work input required in a generic post-combustion carbon capture plant. This section will compare a number of different carbon capture processes which have been proposed in the literature, with the generic plant considered as a benchmark.

The literature does not usually contain sufficient data to establish the exact flows of heat and work in a given plant. Therefore, a simpler method of comparison has been adopted here which looks at the reduction in power station output between a conventional station and one with carbon capture, since these values usually are available in the literature. This reduction in output is then compared with the work input for the generic carbon capture plant, as given by equation (18), running at an equivalent flue gas CO₂ concentration. The resulting *carbon capture cycle efficiency* has been thus defined as

$$\eta_{ccc} = \frac{W_{rev}}{\Delta W} \quad (24)$$

ΔW is the difference in power output between two power stations, with and without carbon capture plant, which are otherwise identical.

Defining efficiency in this way fails to account for the available energy in the flue gas according to equation (20). However, if the flue gases' available energy was considered as a credit to the capture plant, this would lead to anomalies; Fig. 6 shows that the ideal work input may cross through zero, which would lead to

Table 1 A comparison of different post-combustion carbon capture processes

	Units	Frosting cryogenic cycle (refrigeration system COP = 0.55)	Frosting cryogenic cycle (refrigeration system COP = 0.41)	Sub-critical PF plant with amine capture	Sub-critical PF plant with amine capture	Sub-critical PF plant with seawater capture
Cost of capture	MJ/kg CO ₂	1.224	1.708	1.219	1.18	2.526
Estimated flue gas CO ₂	Mole fraction	13.70%	13.70%	10.70%	10.70%	11.20%
Minimum work input: Equation (18) with $\eta_c = 90\%$	MJ/kg CO ₂	0.361	0.361	0.376	0.376	0.373
η_{ccc}		29.48%	21.13%	30.82%	31.84%	14.77%
Reference		[14]		[15]	[16]	[17]

COP, coefficient of performance; PF, pulverized fuel.

meaningless values of efficiency. In addition, since it is unlikely that the flue gases' available energy would be converted to work without the capture plant, it seems reasonable to ignore the contribution that the flue gases' available energy could make when assessing the viability of a particular scheme.

Table 1 includes the actual work input and carbon capture cycle efficiency for a number of different capture processes. Each process was analysed by the respective authors with a capture efficiency of 90 per cent. The processes selected all involve the post-combustion removal of CO₂ from flue gas and use chemical absorption, physical absorption or freezing methods.

As can be seen from Table 1, all of the carbon capture techniques presented have a low second-law efficiency. Of note, however, is the especially poor performance of seawater absorption, despite the fact that such systems are not required to liquefy the CO₂, since the seawater/CO₂ solution can be disposed of directly into the oceans. Nonetheless, the figures clearly indicate that there must be significant entropy production taking place in the absorption section of this type of plant. This may be due to the pressure losses or other forms of lost work generation.

5 CONCLUSION

In the current paper an analysis has been conducted on the minimum work requirement of a generic post-combustion carbon capture plant. The analysis showed that surprisingly little work input is required in the separation and compression stages of such systems. The model was extended to consider the effect of reduced capture efficiency and also the considerable amount of available energy contained in typical flue gas as it is passed to the capture process. It was shown that, rather than work input, it is theoretically possible to achieve a work output from post-combustion capture plant. This is due to the large amount of sensible and latent heat contained in the flue gas.

In the final part of the paper a comparison was made of different types of carbon capture plant. This section showed that all of the post-combustion systems, presently being researched, have poor performance. Of note however, is the especially poor performance of seawater absorption systems.

The next paper in this series, [18], will examine the lost work generation inherent to carbon capture systems in general. In particular, an extension to the analysis conducted above will be undertaken, which considers the unavoidable lost work inherent to mixing processes, both with and without chemical reaction. The analysis of the unavoidable irreversibility will be complemented by consideration of the likely external irreversibility. The conclusion of the paper

will be suggestions of possible means of reducing the lost work generation to a minimum in carbon capture processes.

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APPENDIX

Notation

b	molar steady flow availability function
B	steady flow availability function – mass basis
FG	flue gas
h	molar enthalpy
h^0	standard state molar enthalpy
H	enthalpy – mass basis
m_i	moles of species i
M_i	molecular weight of species i
p	pressure
$\mathfrak{R}$	universal gas constant = 8.3141 kJ/kmol K
s	molar entropy
s^0	standard state molar entropy
S	entropy – mass basis

T	thermodynamic temperature
w_{rev}	minimum work – mole basis
W_{rev}	minimum work – mass basis
y_i	vapour phase mole fraction – species i
Z	equivalence ratio = theoretical/actual air flow
η_c	capture efficiency
η_{ccc}	carbon capture cycle efficiency

Subscripts

c	of capture process (as in efficiency)
cr	control region (volume)
(g)	gaseous state
i	generic species reference
in	of stream flowing into process or operation
j	streams into process or operation
k	streams out of process or operation
(l)	liquid state
mix	mixture property
out	of stream flowing out of process or operation
rev	of reversible process or operation
sat	saturation property

Superscripts

0	standard state: gas at 1.01 bar, 298 K
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