Water Use in Coal-Fired Power Plants Before and After CO₂ Capture

Water is intricately linked to the production and conversion of energy and fuels. Conventional oil and gas production and coal seam gas production result in large amounts of produced water that needs to be dealt with [1–3]. Oil production from shales and tight formations require water for fracking purposes. The generation of electricity in thermal power plants requires water to run the steam cycle; water is also a very convenient cooling medium. Water production and electricity generation are often combined in cogeneration plants [4]. Flowing water can also directly generate electricity through turbines. Water unavailability can severely limit the processes of energy conversion and production. When water is available, it often needs to be brought to the right quality before it can be used, for instance, boiler feedwater must be pretreated to remove all the impurities such as solids, salts, oxygen, etc., to control deposition, carryover, and corrosion.

Water use in coal-fired power plants is largely determined by the cooling system design [5, 6]. This can be based on:

- Once through cooling system using seawater or river water. In this case, large amounts of water are withdrawn from the source and returned after use. The cooling duty is performed through the transfer of sensible heat to the cooling water (typical use: 85–170 m³/MWhₑ).
- Evaporative or wet cooling systems using water from surface or subsurface sources. Here, water is used in a recirculating system. The cooling duty is performed by the transfer of latent heat to ambient air through evaporation in a cooling tower (typical use: 1.5–3.0 m³/MWhₑ for make-up to replace losses and blowdown to limit build-up of impurities).
Dry cooling systems using ambient air. The cooling duty is performed through the transfer of heat to ambient air. Water can be used as the cooling medium in a closed system and only limited makeup is need (typical use: 0.05–0.07 m³/MWhₑ).

Environmental control systems also require water for their operation. The standard limestone-slurry-based flue gas desulphurization process will reduce the temperature of the flue gas through evaporative cooling (typical use: 0.1–0.4 m³/MWhₑ). Bottom ash generated from coal combustion is often transported to a storage location in the form of slurries for which cooling water blowdown can be used. Minor water requirements in thermal power plants are for, for example, boiler feed water makeup (typical use: 0.02–0.03 m³/MWhₑ).

Water use can be described through (1) its withdrawal from the source and (2) its consumption, which is the amount not returned to the source. In the latter case, most of this water is transferred to the atmosphere through evaporation. Power plants at inland locations are often limited by the achievable water withdrawal rates. In those circumstances, the use of a recirculating system with a cooling tower will result in a lower water withdrawal. Only replacement of the evaporative losses and blowdown (to avoid the build-up of salts and other impurities) is needed. If water availability is further limited, that is, in arid regions, a dry cooling system is needed. This leaves the water consumption for the wet flue gas desulphurization as the largest water use component in those circumstances. Water is then predominantly discharged to the atmosphere as vapor in the flue gas.

Retrofitting a coal-fired power station with a CO₂ capture system will impact on the overall performance of the power station. CO₂ emissions will be reduced significantly as capture systems based on reactive absorption are designed to typically capture 90% of the CO₂ present in the flue gas. These systems are mostly based on the thermally reversible binding between CO₂ and amines in an aqueous solution [7]. The energy required for the capture (heat for absorbent regeneration; electricity for rotating machinery) and compression (electricity or steam) is normally provided by the power station and will reduce the output of a coal-fired power station by 20–30% [8, 9]. Process simulation studies indicate that the specific cooling duty will increase in line with the reduction of the power plant efficiency [10–13]. In Figure 1, the specific (evaporative) cooling duty is shown as a function of the coal-fired power plant efficiency assuming that 10% thermal energy input is lost to the air or exhaust gases [14]. It can be seen that a reduction in the power station efficiency by 25% (40% to >30%) will result in an increase in the cooling duty of 60%.

Hence, the implementation of CO₂ capture on existing coal-fired power stations is often considered to result in large additional cooling water requirements. However, it has been shown that for retrofit situations, the increase is less dramatic as the cooling requirement for the capture plant is partially offset by the decreased cooling requirement for the power station [15]. The heat requirement for the amine-based capture plant is provided by steam extracted from the power plant. The steam will not expand through the turbines and subsequently discharge its heat of condensation in the condenser, but instead will condense in the reboiler of the capture plant with the heat of condensation being used to release the CO₂ from the amine solution. The amount of cooling required for the capture and compression process is also dependent on the energy efficiency of this process. Amines requiring less heat for release of CO₂ will require a lower amount of steam to be extracted. More efficient rotating machinery is also beneficial, although some of the losses might be carried away by ambient air, rather than by cooling water. Air cooling in general will be beneficial in reducing cooling water requirements, but is usually more capital intensive and not as efficient as water cooling. Capture processes that require electricity only could also be advantageous provided the equipment is air cooled. Also, integration of the water streams between a capture plant and the host power plant will assist in reducing the usage of freshwater [16].

In countries where flue gas desulphurization is not commonplace (e.g., Australia), the introduction of CO₂ capture retrofit will necessitate significant pretreatment of the flue gas, giving rise to additional cooling water demand. Even without this pretreatment requirement, the flue gas temperature will need to be reduced to enable effective CO₂ capture by a liquid absorbent process. Evaporative cooling of the flue gas is the easiest to implement but will require water of suitable quality to be available.
So, while there is a significant benefit in maximizing the efficiency of the capture system, it is likely that there will be significant cooling requirements. These might be met with an air-cooled system but at a higher cost compared to a water-cooled system. This study presents an option in which water is produced as part of the capture system, providing a contrary perspective on the topic of water requirement in coal-fired power plants retrofitted with CO₂ capture.

**Water Production in an Absorption-Based CO₂ Capture Process**

Liquid absorbent-based processes are currently the commercially dominant technology for both CO₂ capture at low pressure (flue gases) and high pressure (synthesis gas, natural gas). The liquid absorbents used are alkaline solutions often based on amine (monoethanolamine, methyl-diethanolamine, piperazine, etc.) or carbonates. In an absorption-based CO₂ capture process implemented on a coal-fired power station (Fig. 2), the flue gas is first cooled down and/or pretreated to get to a temperature level beneficial for the chemical absorption process and to remove components, for example, SO₂, that might impact negatively on the CO₂ capture process. In general, other acid components, particularly those more acidic than CO₂, will react with and accumulate in the basic solutions. Over time the effectiveness of these solutions to capture CO₂ will be reduced and the process performance compromised.

After pretreatment the flue gas is sent to the absorber where it is brought into contact with the absorption liquid. CO₂ will selectively absorb into the aqueous solution and react with the amine compounds contained in the absorption liquid. The resulting CO₂-rich solution is then fed to the desorber via the lean/rich heat exchanger. At the prevailing desorber temperatures, CO₂ will be released as a wet gas stream. Most of the water vapor is subsequently recovered in the condenser as water and returned to the desorber or another point in the absorption liquid circuit to maintain the plant water balance. The CO₂ product will go to the compression process in preparation for transportation and/or geological storage. The regenerated, CO₂ lean solution leaves the desorber at the bottom and

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**Figure 2.** Amine-based CO₂ capture process flow diagram.
is pumped to the absorber top via the lean/rich heat exchanger and the trim cooler.

The gas leaving the CO₂ absorption section will flow through one or more washing stages to recover amines present in droplets or as vapor in the treated gas. This will ensure emissions of amines and/or amine degradation products to the atmosphere are kept well within the regulatory and/or economic limits.

The proposed concept for water production involves the replacement of the trim cooler by a forward osmosis (FO) operation. In a FO operation, water is selectively transferred through a permeable membrane from a solution of low osmotic pressure to a solution of high osmotic pressure. Similar to the trim cooler, the FO operation might utilize a saline water source such as seawater or from an aquifer that is unsuitable for direct use. As the solutions used for CO₂ capture are usually aqueous solutions with a higher concentration than, for example, seawater, there will be a driving force for transfer of water from its source to the absorption liquid. The water can be recovered from the solution at no additional heat input to the CO₂ capture process in the top of the desorber as shown in Figure 3. The wet CO₂ product leaving the desorber is supplied to a condenser or other heat exchanger with the condensate removed from the absorption liquid loop.

Although requiring more equipment, water taken up by the forward osmosis operation can also be recovered in the absorber, making use of the thermal energy released in the absorber as a result of the exothermic reaction between CO₂ and the amines present. Part of the reaction enthalpy is removed through increased evaporation of water to the flue gas, but a temperature increase in the absorber is inevitable. This will reduce the driving force for CO₂ transfer. Therefore, the thermal energy is usually removed in an intercooler in which part or all of the absorption liquid is cooled. In the arrangement shown in Figure 4, the absorption liquid is evaporatively cooled under vacuum with subsequent condensation of the evaporated water which can then be removed from the absorption liquid loop. In this case, the water recovery process is proposed to involve the use of a membrane-based condenser in which the water evaporates through a permeable membrane which keeps the absorption liquid separated from the vapor and condensate.

The next section provides further detail on the process performance with respect to the amount of water produced from the system.

**Process Modeling of Integrated CO₂ Capture–Desalination Process**

A CO₂ capture process model was developed using the ProTreat® software (Optimized Gas Treating, Inc., Houston, Texas, USA) building on an existing model successfully utilized in support of CSIRO’s CO₂ capture pilot plant program [17]. The CO₂ capture/water recovery process performance was analyzed using the feed gas conditions in Table 1. These are representative of a flue gas in a coal-fired power plant with the flow rate typical of a postcombustion CO₂ capture pilot plant.

The model of the standard absorption/desorption process using a 5-mol/L monoethanolamine solution (as represented by the process flow diagram in Fig. 2) was run to determine the base case technical performance. Monoethanolamine was initially chosen as the capture agent as it is often used as the reference case in CO₂ capture processes [17]. Performance is customarily expressed by the specific reboiler duty, that is, the thermal energy required to release a mass unit of CO₂ in the desorber. A wash section was employed after the absorber, with excess water recycled back to the lean absorbent. The temperature of the wash water was adjusted until the water balance was achieved for the plant (i.e., water in incoming flue gas matched that in the exiting gas streams to within ±1 kg/h). A condenser was employed on the desorber column gas exit. The desorber exit temperature was set to 40°C with 100% of the condensate recycled back to the desorber column. The amount of CO₂ captured was 96.8 kg/h which represents 90% removal from the incoming flue gas. The absorption liquid to flue gas ratios (L/G expressed as mass flow rates for the absorption liquid, L, and flue gas, G) was varied in the base case process to determine the point where the reboiler duty was at its minimum. The results shown in Figure 5 indicate that a minimum reboiler duty of 3.4 MJ/kg CO₂ is obtained for an L/G around 2.6–2.7. At this point, the reflux ratio in the desorber is around 0.41 kg H₂O/kg CO₂.

In the next series of process simulations, water was added to the process upstream of the lean solvent trim cooler mimicking the effect of the forward osmosis operation. The process simulator ProTreat® will add/remove water or amine at this point to ensure mass balance for the plant. The desorber overhead condenser was replaced by a cooler (set to 40°C) and a separator was used to remove the condensate from the system. The condensate leaving the separator was split, with water matching the amount added before the trim cooler removed and the remainder being recycled to the stripping column.

The performance of the CO₂ capture process was modeled for a number of different values for the water addition at the optimum L/G ratio of 2.6–2.7 kg/kg at the same levels of CO₂ capture (90%). In Figure 6, the specific reboiler duty is shown at different levels of water addition by the FO unit.

It was found that addition of water did not affect the specific reboiler duty, which was at the same minimum.
value of 3.4 MJ/kg CO$_2$ up to a water addition of 0.41 kg H$_2$O/kg CO$_2$. However, as the amount of the water addition/removal increased, the reflux flow from the desorber column condenser decreased to zero. If more water was then added to the liquid absorbent, its removal in the desorber resulted in a higher reboiler duty, which is not desired. The maximum water recovery without incurring additional energy consumption is at the point where the reflux flow approaches zero. The increase in absorption liquid flow rate is around 2.5%, which is small enough not to require major changes in the capacities of heat exchange equipment used. This flow rate increase will have a negligible impact on the absorber and desorber column sizes.

An important parameter is the quality of the water recovered in the desorber overhead. At the maximum water recovery, the process modeling indicated that the levels of monoethanolamine were negligible ($\ll$1 ppm), with dissolved CO$_2$ being the main impurity (1750 ppm). Under realistic process conditions the water might contain

\begin{figure}
\centering
\includegraphics[width=\textwidth]{fig3}
\caption{CO$_2$ capture integrated with forward osmosis with water recovery in the desorber overhead.}
\end{figure}

\begin{figure}
\centering
\includegraphics[width=\textwidth]{fig4}
\caption{CO$_2$ capture integrated with forward osmosis with water recovery in the absorber.}
\end{figure}
volatile degradation products such as ammonia and products that are carried over as droplets. This can only be evaluated in practical trials, for example, carried out in pilot plants.

The second route for recovery of water taken up by the liquid absorbent through the forward osmosis process is in the absorber via an intercooling process. In this case, the absorption liquid is withdrawn from the absorber and allowed to cool via evaporation of water under conditions of partial vacuum (the so-called flash operation), followed by recovery of the water in a condensing heat exchanger. The amount of condensate will be dependent on the temperature of the absorption liquid, which is determined by the position in the absorber at which the liquid is extracted and the pressure set-point in the flash operation. This pressure should be lower than the vapor pressure of the absorption liquid for evaporation to occur. In the modified process model, the absorption liquid was withdrawn at the point where a maximum temperature (75°C for a 5 mol/L monoethanolamine solution) was obtained in the base case and pumped to the flash vessel where water was allowed to evaporate at the chosen pressure level. In Figure 7, the amount of water recovered in this manner from the absorber is given for different pressure levels of the flash operation, assuming the flashed vapor is cooled down to 20°C. The amount of water recovered increases with decreasing pressure in the flash vessel. The water recovery at the lowest pressure level of 10 kPa is equivalent to 0.41 kg H₂O/kg CO₂ produced, that is, quite similar to the levels of recovery determined for the desorber.

However, the water recovered in this manner from the absorber has as major impurities monoethanolamine (7700 ppm) and CO₂ (6600 ppm). Depending on the intended use of the water, additional treatment may therefore be necessary to remove these impurities. The presence of monoethanolamine as an impurity is determined by the amount of unreacted monoethanolamine in the absorption liquid and the temperature at the point of intercooling. An optimization study would indicate the best position for the intercooling.

The concerns around the presence of amine impurities in the product water can be reduced if amino acid salt solutions are utilized for CO₂ capture [19]. These capture agents have negligible vapor pressure and process modeling is possible as they are part of the suite of capture agents within ProTreat® [20].

### Table 1. Feed gas conditions for process modeling.

| Component     | Concentration (%) |
|---------------|-------------------|
| H₂O           | 6.9%              |
| CO₂           | 12.1%             |
| N₂            | 74.9%             |
| O₂            | 6.1%              |
| Flow rate     | 600 kg/h          |

| Component     | Concentration (%) |
|---------------|-------------------|
| Temperature   | 40°C              |
| Pressure      | 108 kPa            |

**Figure 5.** Specific reboiler duty as a function of L/G ratio for a 5 mol/L monoethanolamine solution at 90% CO₂ capture.

**Figure 6.** Specific reboiler duty at different values of the water added by the forward osmosis operation.

**Figure 7.** Water recovery in the absorber as a function of the pressure in the flash operation (5 mol/L monoethanolamine).
The base process flow sheet presented in Figure 1 was utilized to determine the process performance with a 6-mol/L aqueous sodium glycinate (Na-Gly) solution as the absorption liquid for CO₂. An optimum reboiler duty of 3.4 MJ/kg CO₂ was determined for L/G equal to 2.5, which was equivalent to the 5 mol/L monoethanolamine solution. Next, the modified process flow sheet with water recovery in the desorber presented in Figure 3 was utilized and the impact of water addition on the reboiler duty was assessed up to the point where the reflux ratio in the desorber approached zero at the optimum L/G ratio of 2.5. For the Na-Gly solution, the maximum amount of water that can be recovered without increasing the reboiler duty was equivalent to 0.31 kg H₂O/kg CO₂. This is around 25% lower than for monoethanolamine and reflects the higher ionic strength of the Na-Gly solution and as a result a reduced water vapor pressure at a given temperature.

The modified process flow sheet with water recovery in the absorber presented in Figure 4 was utilized to determine the performance with a 6-mol/L aqueous sodium glycinate (Na-Gly) solution as the absorption liquid for CO₂. Figure 8 gives the amount of water recovered as determined for different pressure levels in the flash vessel. The amount of water recovered increased with decreasing pressure in the flash vessel. The water recovery at the lowest pressure level of 10 kPa is equivalent to 0.39 kg H₂O/kg CO₂ produced. The product did not contain any Na-Gly and contained gaseous CO₂ as the main impurity (1470 ppm). This illustrates the advantage of utilizing amino acid salt solutions as capture agents in terms of recovered water quality. More water can be recovered from the absorber with intercooling steps used at other positions in the column.

The next section will provide further information on the experimental performance of forward osmosis utilizing aqueous solutions of amines and amino acid salts.

**Forward Osmosis Experimentation with Absorption Liquids**

The experimental section using forward osmosis evaluates the hypothesis mentioned above on the transfer of water from the aqueous saline water source (e.g., sea water) to the absorption liquids at their typical concentrations used for CO₂ capture. Monoethanolamine (Chemsupply, Australia), glycine (Bulk Nutrients, Australia), and sodium glycinate solution were used as absorption liquids. Sodium glycinate was prepared using equimolar amounts of glycine and sodium hydroxide (Chemsupply, Australia). Aqueous salt solution was made with sodium chloride (Chemsupply, Australia) at typical sea water concentrations. All reagents are analar grade.

The concept of FO is pictorially depicted in Figure 9. For an osmotically driven FO process, membranes with higher water flux (water permeating per unit membrane area per unit time) and lower reverse solute flux (permeation of draw solute from draw solution to feed solution) are desired [21–23]. In this respect, FO membrane from Porifera, Inc., USA was obtained. FO membrane area of 0.014 m² was used for this study. The membrane specification has a pure water flux of 28.8 ± 10 L/m² h and reverse solute flux of 0.4 g/L (pure water vs. 1 mol/L NaCl at 25°C) and the maximum operating temperature of 80°C.

Osmotic pressures of common chemical solutions are obtained from commercial software, for example, OLI Stream AnalyzerTM [24]. But for certain substances, especially for complex mixed salt solutions, the data are not available in the software to generate the thermodynamic properties theoretically. In this study, osmotic pressures of absorption liquids monoethanolamine, glycine, and sodium glycinate were determined experimentally. Glycine was included in the experiments because it will be present in a sodium glycinate solution as a product from the reaction with CO₂, even after regeneration of these solutions, albeit at low concentration. The absorbent solution with unknown osmotic pressure was introduced on one side of the FO membrane and sodium chloride with known osmotic pressure as the draw solution on the other side of the membrane. The concentration of the sodium chloride was gradually reduced, while the concentration of the absorption solution was kept constant or vice versa. Eventually the system reached an equilibrium and at this point the osmotic pressure of the absorption solution equals the osmotic pressure of the sodium chloride solution resulting in no water transfer across the membrane. Thus, the osmotic pressures of the two absorption liquids were determined. Figure 10 shows the water fluxes with the three absorption solutions in order to determine the equilibrium point concentration.
negative water flux indicates the reversal of water permeation direction between feed and draw solutions due to the change in osmotic pressure of the solutions.

Table 2 gives the experimentally determined osmotic pressure of monoethanolamine, glycine, and sodium glycinate absorption liquids. Osmotic pressure of sodium chloride is also provided as a reference. The osmotic pressure of sodium glycinate amino acid absorption solution was higher than monoethanolamine. The osmotic pressure of glycine was lower than the other two absorption liquids.

The absorption liquids were tested for water flux permeation with representative sea water using 0.59 mol/L NaCl solution (Table 3). It was found that at typical concentrations of absorption liquids used in the CO2 capture process, water was able to be transferred from the sea water solution to the absorption liquids. The experimental results indicate that the proposed concept of water production and sea water desalination with absorption liquids is possible with the use of forward osmosis in the CO2 capture system. As the osmotic pressure of sodium glycinate amino acid solution is higher than monoethanolamine and glycine, its water flux was found to be the highest.

In forward osmosis, membrane orientation also has an effect on the flux. Orientation of active layer of the FO membrane toward the feed or draw solution had a significant effect on the flux profile. As shown in Figure 11, the FO operation with sea water as draw solution (membrane active layer facing the absorption liquid) produced a higher flux compared to having sea water as feed. This was valid for all the three absorption solutions studied.

It is to be noted that the flux profile was obtained with a starting NaCl concentration of about twice that of sea water.

In order to study the effect of absorption liquid temperature on water flux, sodium glycinate solution was tested at elevated temperature (45 ± 2°C) while keeping the sea water at ambient temperature. Figure 12 shows that the water flux increased with the increase in absorption liquid temperature. This means that more water can be produced (enhancement in sea water desalination) at higher absorbent temperature. The sea water is being concentrated due to the selective transfer water across the membrane from sea water to the absorption solution. The concentration profile of sea water is also shown in Figure 12 for the two different absorption liquid temperatures under the operating conditions adopted in this study. The effect of temperature on the reverse solute diffusion requires further evaluation.

Monoethanolamine was also tested at elevated temperature (45 ± 2°C), with sodium chloride (starting concentration about twice the concentration of sea water) draw solution at ambient temperature. The flux and the temperature patterns of the monoethanol amine through the FO is shown in Figure 13. More than 15% improvement in water flux was achieved with monoethanolamine operation at inlet temperature of 45°C, compared to at 25°C. As the absorption liquid flows through the FO semipermeable membrane, with draw solution (sea water or other concentrations) on the other side, the absorption liquid is being cooled and the exit (return from FO) temperature was found to be about 26.5°C (see

**Figure 9.** Diagrammatic representation of forward osmosis membrane operation characteristics.
Fig. 13). This supports the concept of replacing the trim cooler in CO$_2$ capture system with FO, with additional benefit of water desalination.

Table 2. Osmotic pressure of absorption liquids (3 wt % at 25°C).

| Test solution      | Osmotic pressure (atm) |
|--------------------|------------------------|
| Monoethanolamine   | 13.2                   |
| Glycine            | 11.1                   |
| Sodium glycinate   | 21.7                   |
| Sodium chloride    | 26.5                   |

Table 3. Sea water desalination using three absorption liquids in forward osmosis (0.59 mol/L NaCl as sea water, 25°C).

| Absorption liquid         | Average water flux (L/m$^2$ h) |
|---------------------------|---------------------------------|
| 5 mol/L Monoethanolamine  | 16.83                           |
| 6 mol/L Glycine           | 12.44                           |
| 6 mol/L Sodium glycinate  | 35.1                            |

Considerations on Simultaneous Heat and Mass Transfer in the FO Unit

The FO unit will need to perform the dual tasks of cooling down the lean absorption liquid while transferring water from the salty cooling water. Ideally, these tasks will be integrated in a synergetic way where the heat transfer and transfer of water are matched. The process modeling indicates that if too much water is transferred, this will incur an additional energy demand to remove the water in the desorber. If the heat exchange of the FO unit is insufficient, an additional heat exchanger might be needed. Conversely, if the heat exchange function of the FO unit is very effective, then a limited amount of water might be transferred to the absorption liquid from the salty cooling water. While this requires a more in-depth design study on the heat transfer characteristics of the membrane module, one can make an early assessment of the likely degree of matching between heat and mass transfer in the FO module. This has been carried out for
Figure 11. Water flux profile for absorption liquids versus NaCl (A) 5 mol/L monoethanolamine versus 1.13 mol/L NaCl at 25°C, (B) 6 mol/L glycine amino acid solution versus 1.13 mol/L NaCl at 25°C, and (C) 6 mol/L sodium glycinate amino acid solution versus 1.13 mol/L NaCl at 25°C.

Figure 12. Effect of temperature of absorption solution on water flux (0.59 mol/L NaCl sea water draw solution at 25°C, starting concentration of sodium glycinate absorption liquid feed solution 6 mol/L, forward osmosis membrane active layer facing feed solution).

Figure 13. Water flux and temperature profile at elevated monoethanolamine temperature (5 mol/L monoethanolamine absorption feed solution).
a practical CO₂ capture plant with a capacity of 1 Mtonne per year with 7500 h of operating time using the results from the ProTreat® modeling. The absorption flow rate will increase by 2.5% in the FO unit through the transfer of water. In Table 4, the area requirement for the two separate functions of heat and water transfer are given. For the heat transfer the range of typical coefficients for shell and tube heat exchangers are used, whereas for the FO unit typical values for the water permeability are used. The area requirements are quite similar, indicating that with the current level of knowledge, neither the heat transfer nor the water transfer will be dominant in the design. It is clear that further insights need to be obtained on the combined heat and mass transfer process. This can be achieved by a more detailed model for the heat and mass transfer phenomena in the FO unit with experimental validation.

### Utilization of Recovered Water in Power Plant

The previous sections have shown that water can be recovered from available saline water streams in combination with a liquid absorbent-based CO₂ capture process integrated with a coal-fired power plant. This can be done at negligible additional energy consumption and provides an opportunity for those locations where water has limited availability. Table 5 provides an overview of the water production rates through the recovery from the absorber and the desorber. Water production rates are determined by the amount of CO₂ captured, and the process modeling results indicate that 0.7–0.8 (kg H₂O/kg CO₂) can be recovered. With coal-fired power stations having typical CO₂ emissions intensities of 700–900 kg CO₂/MWhₑ and with 90% CO₂ capture, this translates to 0.4–0.6 m³/MWhₑ produced by the integrated CO₂ capture–desalination process. Here the water production refers to the original

| Table 4. Comparison heat and water transfer for a typical CO₂ capture application. |
|-----------------------------------|--------|--------|--------|
| Plant capture capacity            | 133.3 ton CO₂/h |
| Absorption flow rate              | 2080 ton/h before FO |
| Absorption liquid temperature     | 57 to 40°C |
| Salty cooling water temperature   | 25–35°C |
| Osmotic pressure gradient         | 10 MPa |
| Heat transfer area (heat transfer coefficient = 0.5–2.0 kW/m²K) | 950–3750 m² |
| Membrane area (water permeability = 0.5–2.0.10⁻³ kg/m²sec MPa) | 750–3000 m² |

| Table 5. Overview of water production rates from saline water through the CO₂ capture process. |
|--------------------------------------------------|--------|--------|--------|
| Absorption liquid                  | Water recovery from desorber (kg H₂O/kg CO₂) | Water recovery from absorber (kg H₂O/kg CO₂) | Total water recovery (kg H₂O/kg CO₂) |
| 5 mol/l Monoethanolamine            | 0.41   | 0.41   | 0.82   |
| 6 mol/l Sodium glycinate            | 0.31   | 0.39   | 0.70   |

**Figure 14.** Overview of water use ranges in coal-fired power plants (left) and potential water production (right) in both cases per unit of electricity generated.
power plant output, that is, before CO₂ capture. This reference is chosen to compare the water production to water use rates and water production from flue gas.

An alternative means to produce water in a power plant is through recovery of water present in the flue gas. This can be realized using, for example, gas separation membrane technology [25] or through an absorption process [26]. The amount of water present in the flue gas of a coal-fired power station is dependent on the coal type, in particular its water content, the power plant efficiency, and will typically be in the range of 0.25–0.45 m³/MWhₑ, assuming complete recovery of the water vapor. It has to be borne in mind that recovery of water in the flue gas will inevitably lead to an additional cooling duty equivalent to the amount of water recovered if evaporative cooling is used. The combined CO₂ capture–desalination will only utilize existing cooling duties to recover the water in a liquid form and can produce more water of adequate quality. In Figure 14, an overview is provided of the various water uses in a coal-fired power plants (left hand side) and the production of water from the power plant from the flue gas or a CO₂ capture process (right hand side). The whole range of power plant water uses and type of cooling sources are considered. It can be seen that the integrated CO₂ capture–desalination process is able to produce sufficient water to fulfill the needs of the flue gas desulphurization process and might even provide a significant part of the water requirement in an evaporative cooling water system. In arid regions, coal-fired power plants are often air cooled. Implementing CO₂ capture would result in a cooling requirement that would be easily met with the integrated CO₂ capture–desalination process where there is a saline water source available of sufficiently low temperature to provide the cooling duty.

### Evaluation and Further Work

Our study indicates that 0.4–0.6 m³/MWhₑ can be produced by the integrated CO₂ capture–desalination process at negligible additional energy use. At these amounts and anticipated water qualities the process is of direct interest to coal-fired power plants where CO₂ capture is retrofitted with a liquid absorbent-based process. Further work will need to focus on developing a process design that will provide the basis for the costing of the process. The process design would benefit from pilot plant work in actual service that would confirm the process performance and its variation under different process conditions. Maintenance of the overall water balance in the capture process will be an important point of attention. Duration experiments would confirm the process robustness and provide insights into, for example, the forward osmosis membrane performance over time with the alkaline absorption liquids, as well as the requirements for liquid pretreatment prior to the forward osmosis unit. Recovery of water in the desorber is quite straightforward through an overhead condenser. Recovery of water from the absorber is more complex as it involves evaporation and condensation under vacuum. Membrane evaporation integrated with a condenser requires further technology development.

The proposed method of water production needs to be competitive compared to other methods of water production via desalination. Reverse osmosis processes are able to produce freshwater from seawater at an electrical energy requirement of ~5 kWhₑ/m³ [27]. So existing desalination processes might be able to produce water of good quality with only limited additional electricity consumption. As a result of the high concentrations of amines or amino acid salts in solution the integrated CO₂ capture–desalination process is able to recover water from higher concentration solutions than seawater, that is, it is applicable to a wider range of salt concentrations and could be used to extract freshwater from the reject water from existing seawater desalination processes. Significant synergistic potential is also expected in the desalination of formation water associated with CO₂ storage where high salt concentrations brines are encountered [28]. The process concept would enable the production of freshwater from these brines where the CO₂ capture operation is close to the CO₂ storage complex.

### Conflict of Interest

None declared.

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