Carbon Dioxide Capture for Storage in Deep Geologic Formations – Results from the CO_2 Capture Project

Capture and Separation of Carbon Dioxide from Combustion Sources

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Chapter 7

KPS MEMBRANE CONTACTOR MODULE COMBINED WITH KANSAI/MHI ADVANCED SOLVENT, KS-1 FOR CO_2 SEPARATION FROM COMBUSTION FLUE GASES

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ABSTRACT

The Kansai Electric Power Co., Inc. (Kansai) and Mitsubishi Heavy Industries, Ltd (MHI) have developed solvents for a CO_2 capture process. One of the solvents, KS-1 was selected for this combined process with the gas/liquid membrane contactor, developed by Kvaerner Process Systems a.s. (KPS).

The KPS membrane contactor and Kansai/MHI, KS-1 solvent both pose technical advantages to the current convention of CO_2 capture processes, respectively; however, the combined effect has never been determined.

This test was undertaken in order to determine the extent of advantages the combined process holds over the current standard of CO_2 capture. Data was recovered for the construction of a mathematical model regarding the performance of the combined process. This data were in turn, used for the scale-up calculations for a CO_2 capture plant at a 350 MW power plant.

In the second phase of the CCP project, the principle of a membrane water wash unit was tested in a small-scale laboratory unit. This was done to verify the upscale calculations done for a membrane water wash unit in the first phase of the   project.

INTRODUCTION

Background
The membrane contactor research work was initiated with a study made for the Norwegian State Pollution Control Authorities (SFT) in 1992. Since then AkerKvaerner has continued to develop and improve this "new" gas-treating process. In the effort to reduce space and weight requirements in connection with removal of CO_2 on offshore installations, the technology of replacing the absorber column with a more compact membrane unit was considered to be the most promising.

This development work opens a number of international industry applications for the membrane gas/liquid contactor technology also for use onshore (Figure 1).

Through their cooperation, Kansai/MHI has created a new type of amine solvent for the capture of CO_2 from flue gases. Pilot operation, laboratory tests, and commercial experience have shown that the KS-1 process technology is superior to other amine-type processes in terms of capture performance and energy consumption for conventional CO_2 capture process. The KS-1 process is known to be one of the best CO_2 capture methods available in the market.

The Combined Process
The Kansai–MHI process/solvent
The KS-1 process is the product of a rigorous joint-research project between Kansai and MHI. This process utilises an amine-type solvent for the capture of CO_2 from flue gases. Pilot and laboratory research studies,

as well as commercial experience have shown that the KS-1 process technology is superior to other amine-type processes in terms of capture performance and energy consumption for conventional CO_2 capture processes.
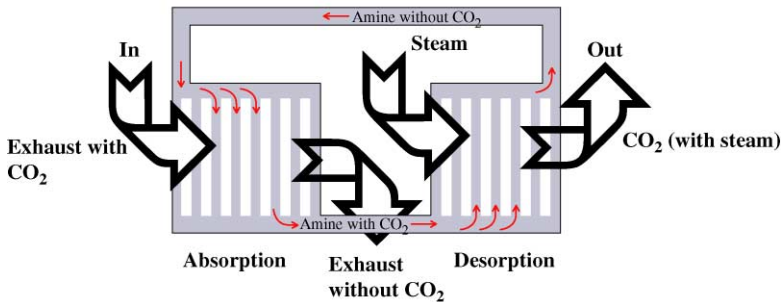
Figure 1: Membrane contactor CO_2 removal process.

In the comparative study of conventional process, significant reduction (over 25%) in regeneration energy was observed for KS-1 process under MHI's in-house experiments. One of the reasons for KS-1 solvent having lower energy consumption, in contrast to MEA, is due to the difference in CO_2 solubility. As can be seen from Figure 2, the range of CO_2 loading for KS-1 is much wider than the range for MEA. This chemical property of KS-1 allows for a higher CO_2 solubility per unit volume of solvent in comparison to MEA; signifying the fact that less solvent volume is required for KS-1 solvent to capture the same quantity of CO_2 as MEA. In solvent regeneration, less stripping steam is required to regenerate the significantly lowered circulation volume of KS-1 solvent. Hence, the consumption of regeneration energy at the regenerator reboiler is reduced.
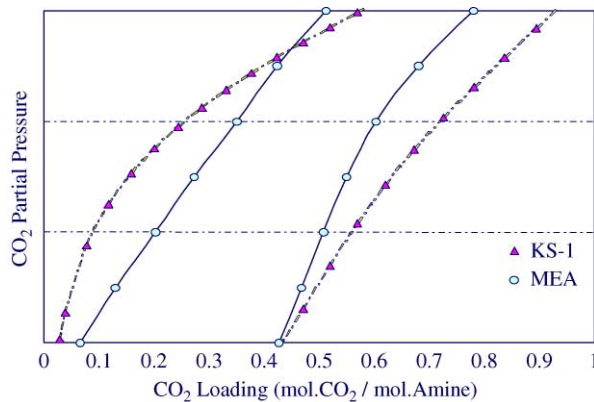
Figure 2: Solubility of CO_2 in KS-1 and MEA solutions.

KS-1 solvent is a sterically hindered amine, in contrast to MEA, which is a primary amine (unhindered). When MEA undergoes reaction with CO_2, its main reaction forms carbamate as its product, a stable compound (Figure 3). Higher heat of dissociation will be required during solvent regeneration, in order to break its bond with CO_2. For a sterically hindered amine, such as the KS-1 solvent, the primary reaction pathway does not involve the formation of carbamate, hence, less stable products are formed; therefore, less energy is required for solvent regeneration.

Mono-ethanol Amine (MEA)

{Main Reaction} $2\ R\text{-}NH_2 + CO_2 \longleftrightarrow R\text{-}NH_3^+ + R\text{-}NH\text{-}COO^-$

$$R\text{-}NH_2 + CO_2 + H_2O \longleftrightarrow R\text{-}NH_3^+ + HCO_3^-$$

Sterically Hindered Amine (KS-1)

$$2\ R\text{-}NH_2 + CO_2 \longleftrightarrow R\text{-}NH_3^+ + R\text{-}NH\text{-}COO^-$$

{Main Reaction} $R\text{-}NH_2 + CO_2 + H_2O \longleftrightarrow R\text{-}NH_3^+ + HCO_3^-$
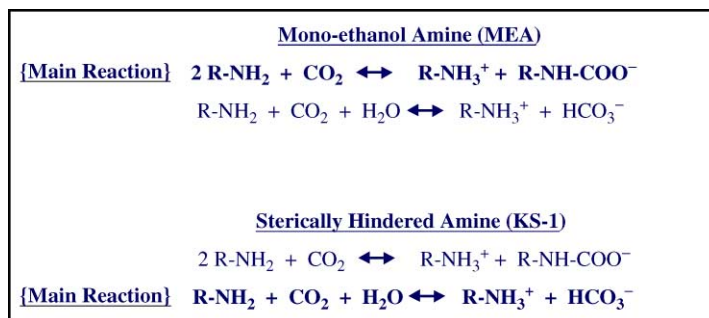
Figure 3: Reaction of hindered and unhindered amine.

The KS-1 solvent is not a corrosive substance in contrast to the corrosive MEA solvent, which requires the use of a corrosion inhibitor in operation. Table 1 shows the MHI in-house results of corrosion tests.

TABLE 1
CORROSION TEST RESULTS

	Test 1	Test 2
MEA	93.0	76.4
MEA + inhibitor	9.5	8.3
KS-1	3.1	3.6

Unit: mils per year; test condition: 130 °C, in presence of O_2.

Degradation of solvent is also significantly lower for KS-1 in comparison to MEA. Figure 4 displays both the in-house experimental results, as well as commercial results of solvent degradation for KS-1 solvents. As it can be seen, the rate of heat-stable salt (HSS) formation is significantly lower for KS-1. Proportionally frequent reclaiming operation is required for the MEA solvent. From this tendency, the rate of waste product formation, as well as difficulty of operation is significantly lower for the KS-1 process.
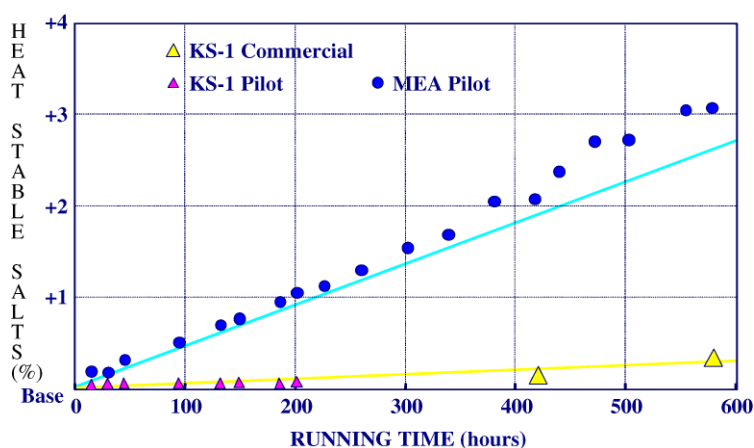
Figure 4: Heat-stable salt accumulation.

The AkerKvaerner gas/liquid membrane contactor technology

AkerKvaerner and W.L. Gore & Associates GmbH, Germany (Gore) have developed and tested process concepts and membranes and can deliver commercial membranes with custom characteristics for a particular application.

The AkerKvaerner gas/liquid (G/L) membrane contactor operates with liquid on one side and gas on the other. Unlike gas separation membranes where differential pressure across the membrane provides the driving force for separation, the pressure is almost the same on both sides of the AkerKvaerner membrane. Absorption into the liquid provides the driving force. The Gore membrane material (expanded PTFE, *e*PTFE) is virtually non-destructible under the operating conditions encountered in typical natural gas or exhaust gas applications, employing most of the common treating solvents (Figure 5).
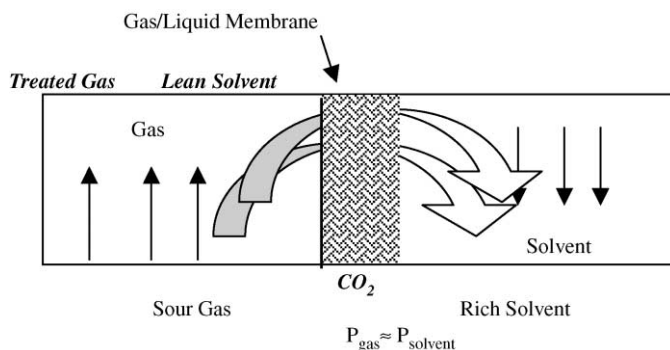
Figure 5: Gas/liquid membrane contactor in service as absorber.

The exhaust gas enters the membrane contactor where the CO_2 diffuses through a membrane into the lean amine solution, which chemically absorbs the CO_2 from the exhaust gas to meet treated gas specifications.

The gas and liquid flows are cross-flow for each module. However, the arrangement of multiple modules in series gives effectively a counter-current flow; the gas inlet with the highest CO_2 content meets the rich amine flow and the gas outlet (lowest CO_2 content) hits the lean amine flow.

The separation of components is caused by the presence of an absorption liquid on one side of the membrane, which selectively removes (absorbs) certain components from the gas stream on the other side of the membrane. The membrane provides a large contacting area without direct contact between the gas and the liquid.

The membrane should be highly permeable to the component whose removal is desired. The selectivity of the process is determined by the absorption liquid, which means that a highly selective separation can be obtained through a suitable choice of the absorption liquid. Figure 6 shows the principle of a membrane gas/liquid contactor. (Note: The figure does not represent a counter-current configuration, but is intended for explanation only.)

The difference in the way mass transfer takes place gives the following key advantages for the AkerKvaerner membrane technology;

1. High flexibility with respect to flow rates (liquid to gas ratios) and solvent selection.
2. Separating the phases eliminates the usual limitations of packed towers caused by foaming, flooding and entrainment of the liquid with the up-flowing gas.
3. The hollow fibre membranes give the possibility of a very high specific area for a membrane contactor. Depending on the fibre diameter, very large specific areas can be achieved for a membrane contactor.

Practical considerations, like pressure loss, limit the value to somewhere between 500 and 1000 m^2/m^3. This is five times greater than in a tower, where values of $100-250$ m^2/m^3 are common. The high specific area allows for the possibility for reductions in volume and weight for the contactor of typically $65-75\%$.

4. Flexibility with respect to orientation of the unit(s).
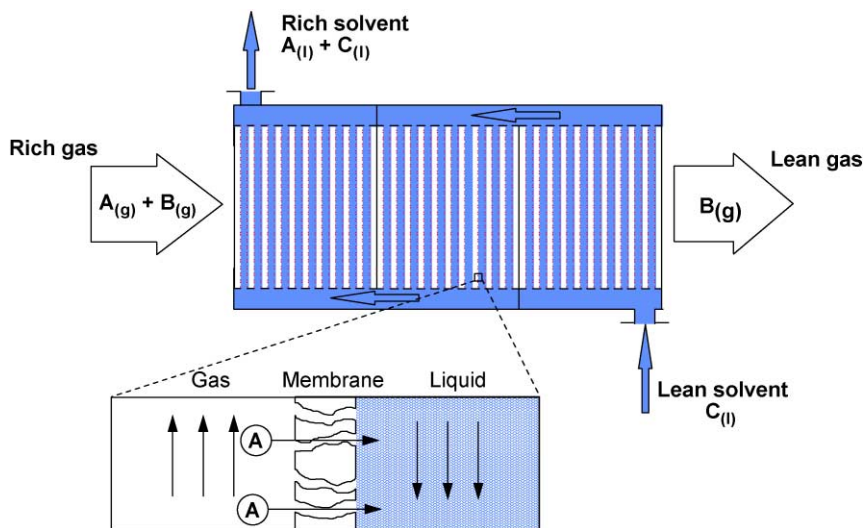
Figure 6: The principle of a membrane gas/liquid contactor used as absorber.

Outlet flue gas amine content, water wash

In the membrane contactor, some amine will vaporise and diffuse through the membrane. A water wash section is used to control the amine and water losses in the process. The water wash is used primarily in amine systems, especially at low absorber operating pressure, as the relatively high vapour pressure of the amine use may incur appreciable vaporisation losses.

The loss of water and amine is important with respect to the operational cost of the total process. Another important input to the design of the water wash is the environmental effect and regulations levied by the authorities related to emission of amines. There is, however, no clear statement from the authorities regarding this matter.

A water wash unit must be used to remove the last part of the amine from the gas phase.

A PFD is given in Figure 7.

In a conventional tower, the water wash is located in the top section of the absorption tower. The water wash section is typically $1/3-1/4$ of the total tower height. See Figure 8, as-built-drawing from the absorption tower at the AkerKvaerner pilot unit at K-Lab, Kårstø, Norway. The cost of a tower will not increase linearly with the height.

If existing water wash technologies should be used downstream the membrane contactor, the water wash unit will have a significant size (6L $\times$ 32W $\times$ 16H for a typical 350 MW unit), which is bigger than the membrane contactor itself (6L $\times$ 32W $\times$ 5H). The large size and weight (estimated to 400 tonnes) will significantly increase the construction and installation costs.
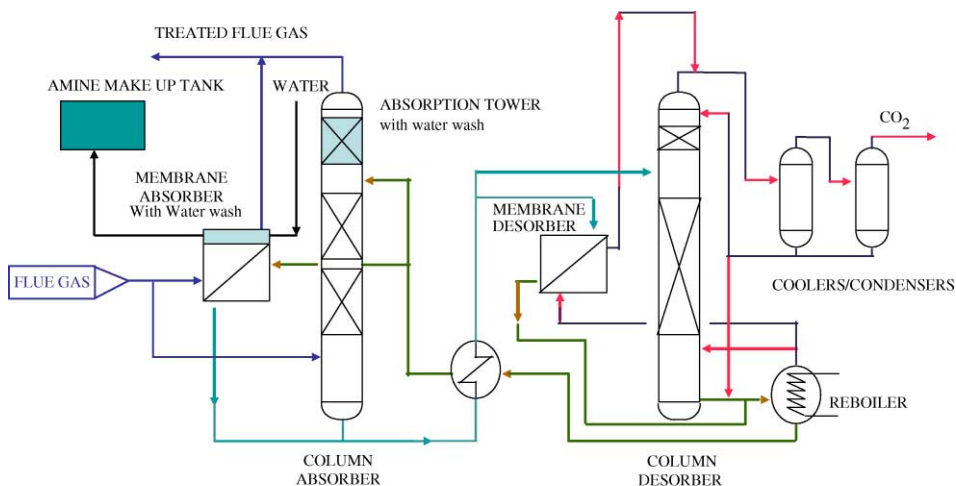
Figure 7: PFD of a CO$_2$ recovery unit with water wash. The wash water outlet will be transferred to the amine make up tank.

To be able to utilise cost reduction potential in the membrane contactor technology, a compact membrane water wash concept was developed by Kvaerner Process Systems a.s. (KPS). The water wash system contains a small membrane contactor where the water wash water (containing the solvent) is re-circulated and a bleed of the wash water is directed to the lean amine tank. See Figures 7 (PFD) and 8. If required, a small polishing module (using pure water) can be added after the water wash. The fresh water in the polishing section will absorb the rest of the amine. The fresh water is directed to the lean amine tank. A simulation tool is developed based on the existing and verified internal KPS simulation tool. The size of the membrane water wash was estimated to be 6L × 32W × 1H by using the simulation developed for the last version of the report. The concept has been confirmed by tests at SINTEF (Figure 9).

EXPERIMENTAL/STUDY METHODOLOGY

The Pilot Plant
The CO$_2$ recovery pilot plant is located within the Kansai Nanko Power Plant in Osaka Prefecture, Japan. Flue gas used at the CO$_2$ capture pilot plant is drawn from one of the three LNG-fired boilers for the power generators at the power plant facility. The boiler flue gas contains approximately 10 vol% of CO$_2$, and maintains a relatively stable operation during daytime of the summer and the winter seasons. Fortunately, this study was conducted during the winter season, with minimal fluctuation in flue gas CO$_2$ concentration.

The CO$_2$ recovery plant
In the original flow scheme of the Nanko pilot facility, the flue gas CO$_2$ capture consists of two towers as its main constituents, the absorber and the stripper. Flue gas from the power plant boilers initially enters the flue gas cooling tower, and is cooled to approximately 40 °C before entering the flue gas blower. The flue gas then enters the absorber, where the solvent comes into direct contact with the solvent, and its CO$_2$ absorbed. The flue gas, with its CO$_2$ partially captured in the absorber, is exhausted through the top section of the tower, to the stack approximately 200 m in height after it is cleaned of by its residual amine and water vapour content in the washing section. Meanwhile, lean solvent enters the absorber through the upper section, and travels downward through the layers of absorber packing. The lean solvent captures CO$_2$ through direct contact with the flue gas on the surface of the tower packing, and becomes a rich solvent by the time it reaches the bottom of the absorber. At the tower bottom, the rich solvent is directed to the stripper through the solvent heat exchanger, where heat is recovered from the lean solvent from the stripper bottom. The rich solvent then enters the stripper, where heat is applied to the system through

Figure 8: 3D drawing of the absorption tower, pilot unit at K-Lab, Kårstø, Norway. Pink/dark colour illustrates the packing material sections with the water wash at the top.

the stripper reboiler in order to release the CO_2 from the solvent. CO_2 is exhausted through the top of the stripper, and the regenerated lean solvent is directed back to the absorber through the solvent heat exchanger once again.

The KPS membrane was made to replace the absorber for the current pilot test operation. Basic process configuration adjustment consisted of redirecting pipelines to the membrane contactor module that were originally directed towards the absorber.

The membrane contactor
The membrane contactor used at the Nanko pilot unit was designed to remove 30–60% of the CO_2 content in the exhaust gas stream. Since this was the first time the KS-1 was tested with the AkerKvaerner membrane, there were several uncertainties. The size of the contactor had to be large enough to give
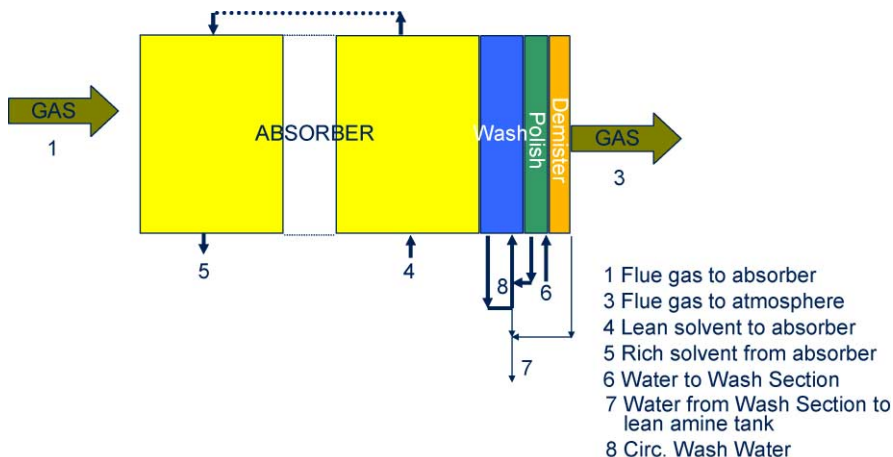
Figure 9: Membrane contactor with water wash (re-circulated water containing the solvent), polisher (fresh water) and demister.

statistically significant data, i.e. the data should be significantly higher than the uncertainties in the measurements. It was also important to keep the CO_2 liquid loading below the equilibrium, to eliminate the discussion of where the equilibrium was reached in the contactor.

From previous tests, it is known that some liquid is drained out from the gas side. The module was designed in such a way that the gas side could be easily drained, so the drainage could be measured and samples collected. The main purpose of these experiments was to obtain sufficient data to verify the simulation tool in order to be able to do a scale-up calculation to a commercial unit size (Figure 10). The membrane module design was based on:

Gas flow rate	575 Nm^3/h (operated at 555–840 Nm^3/h)
Liquid flow rate	1 m^3/h (operated at 0.48–1.44)
CO_2 removal rate	30–60% (approved removal rate at design conditions 35–47%)
Gas pressure drop	0.05 bar (operated at 0.01–0.015)
Liquid side pressure drop	0.1 bar (operated at 0.1–0.3)

A schematic flow diagram of a CO_2 removal from exhaust gas by membrane absorption system (G/L contactor) is shown in Figure 11.

A membrane contactor is packed in standardised modules for tests and industrial usage. To achieve the required outlet purity one can tailor-make a "process train" with several modules in series.

The Test Matrix
There are three main factors that constitute the test matrix. The factors are the observed data, the controlled parameters, and the analytical methodology. The controlled parameters were adjusted in accordance with the progression of the test, as various unknowns existed at the initial stages of the test operation.

Observed data
The primary objective of the test operation was to measure the performance of the combined processes of the KPS membrane contactor and the Kansai/MHI KS-1 process. In order to meet this objective, the focus

Figure 10: Test unit at the site.

was placed upon data regarding CO_2 capture performance and solvent regeneration energy. The objective data set is shown in the table below. Circumstantial data that explained experimental phenomenon were also focused as the test operation progressed. Data was observed through online instrumentation and laboratory analysis results.

Observed data are:

membrane contactor temperature profile;
membrane contactor outlet gas CO_2 concentration;
lean/rich solvent CO_2 loading;
product CO_2 flow rate;
stripper reboiler heat duty;
membrane contactor pressure drop (gas & solvent side).

Controlled parameters
The controlled parameters are listed in the table below. The inlet flue gas CO_2 concentration was adjusted by diluting the initial 10 vol% CO_2 flue gas with air (prior to the flue gas cooling tower) in order to simulate
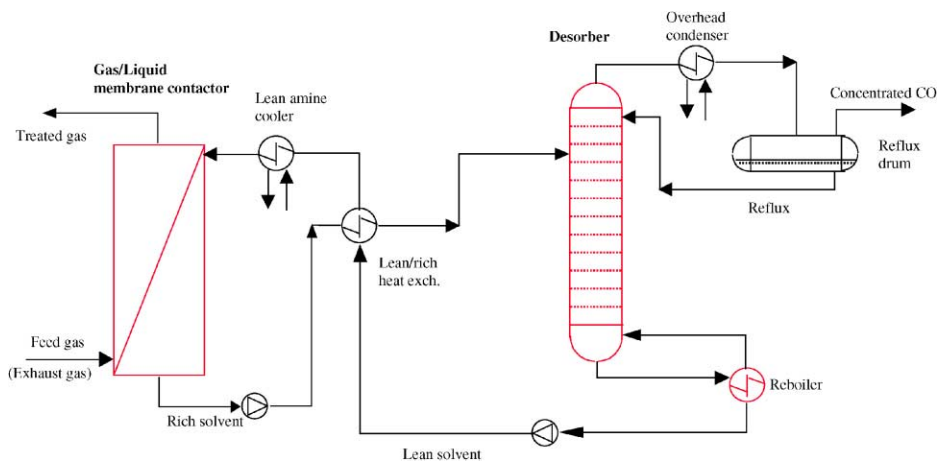
Figure 11: A schematic flow diagram of CO_2 removal from exhaust gas by membrane absorption system (G/L contactor).

CO_2 concentrations of 3, 6, and 10%. Flue gas normal flow rate and solvent flow rate were adjusted in order to test performance levels at varying solvent–gas ratios. Finally, the stripper reboiler steam flow rate was adjusted in order to control the lean solvent CO_2 loading

Controlled parameters	Values
Inlet flue gas CO_2 concentration	3/6/10 (%)
Flue gas normal flow rate	555/760/840 (m^3/h)
Lean solvent flow rate	0.80/1.08/1.44 (m^3/h)
Stripper reboiler steam flow rate	Variable

Water Wash Experiments

The experimental set-up with the small membrane module is schematically shown in Figure 12, and is installed in a temperature-controlled cabinet. The feed to the recirculation loop is MEA vapour in nitrogen, which is generated when nitrogen is bubbled through two cells filled with pure MEA. The amine concentration in the feed corresponds approximately to saturation at the cell temperature (no droplets are seen). The flow rate of amine depends on the nitrogen flow rate, which was controlled by a mass flow controller (MFC).

The membrane contactor is operated counter-currently with the absorption liquid inlet (water or 5 wt% MEA = 827 mol MEA/m^3) at the bottom. The gas flow rate in the recirculation loop was regulated by a frequency regulation of the blower to 3–3.3 m^3/h.

The liquid flow rate was 0.200 L/min in all tests and the temperature was quite stable ∼40.6–41.2 °C. The process was controlled by a computer system (Labview). Samples are taken from the feed stream and from the outlet (purified) gas to determine the amine concentration. The amine vapour is absorbed in 0.05 M sulfuric acid, and analysed according to the Kjeldahl method (NS-ISO 5663) [1,2]. (The amine concentration is determined as the total concentration of nitrogen in a sample).

Figure 12: Experimental set-up.

The figure of the experimental test unit must not be mixed with a commercial unit. Recirculation of the gas is done to save gas expenses.

Observed data
The main objective of the test operation was to measure the performance of the KPS membrane water wash unit. The observed data set is shown in the table below. Data was observed through online instrumentation and laboratory analysis results.

Data to be observed:

liquid temperature;
inlet/outlet gas pressure;
gas inlet/outlet temperature;
inlet/outlet amine concentration in gas;
recirculated gas flow rate.

RESULTS AND DISCUSSION

Test Results Nanko
The duration of the test operation was from 15th January 2002 to 10th February 2000 while the span of continuous operation was from 17th January to 10th February. The total operation time of this test was 573 h and 20 min.

At the initiation of the test operation, various unknowns existed to both KPS and MHI. Factors such as membrane performance with KS-1 solvent were difficult to predict for both parties because there were no precedent experiments. After careful analysis of the test data, it was found that further optimisation can be made for the KS-1 process, possibly improving performance to a significant degree.

Material balance
Material balance of the operation data was taken in order to assure high confidence in the recorded data. CO_2 capture rates were calculated for flue gas, solvent, and product CO_2 flow rate. The general trend throughout the operation was that the CO_2 capture rates calculated from lean/rich solvent loading and CO_2 product flow rate (99.9% CO_2, product flow, from stripping reflux drum) was proximate. However, the CO_2 capture rate calculated from the difference between membrane inlet and outlet CO_2

concentrations of the flue gas was found to be significantly different to the other two data sets. Reliability of instrumentation was also taken into consideration; the CO_2 meter for the product CO_2 had the highest reliability. Therefore, the capture rate calculated from the CO_2 product was used as the basis for the material balance.

Heat duty comparison
The heat duty of the stripper reboiler was determined from the observed steam consumption value. In order to determine the reliability of the adjusted data, the stripper reboiler heat duty was calculated through simulation using the adjusted (from material balance) rich solvent values and flow rate. The simulation result was compared against the observed heat duty value.

CO_2 product purity
The purity of CO_2 product was found to be around 99.9 mol%. The data for all of the analyses during the Nanko Pilot testing is shown in Table 2.

TABLE 2
GAS ANALYSES RESULTS

Run No.	3	6	8	11	14	20	23
Date	1/20	1/23	1/26	1/29	2/1	2/7	2/10
CO_2 Purity (mol%)	99.87	99.89	99.93	99.94	99.93	99.85	99.95
N_2 (ppm)	1189	986	–	553	712	1418	515
O_2 (ppm)	94	201	–	72	84	91	62

The purity of product CO_2 at the Nanko Pilot test was slightly lower than MHI's expectations for conventional CO_2 capture process. Theoretically, the use of the membrane should lead to higher purity of the product CO_2 in comparison to conventional process because there is no direct contact of solvent and flue gas. The reason for the existence of N_2 and O_2 in the product CO_2 stream was due to the alteration made on the flow scheme during the testing period. This will not occur in a commercial unit; the purity of CO_2 product is expected to be higher than what was attained during the pilot test.

Test Results for Membrane Water Wash
The tests were performed from 4th November to 22nd November 2002 at SINTEF in Trondheim, Norway. Figure 13 shows the amine content in the gas inlet and outlet of the membrane water wash unit. This figure only shows the difference for the inlet and outlet amine concentrations. The parameters in these experiments are not equal. The uptake will vary with, for example, the flow rates so the experiments cannot be directly compared.
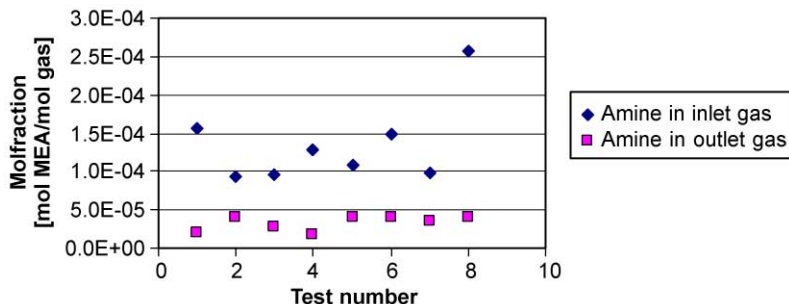

Figure 13: Amine content in the inlet and outlet gas from the experimental water wash unit.

The membrane water wash tests cannot be run with an optimised water wash unit, but with a small unit. Hence the outlet amine concentration could not reach 3 ppm like in the upscale calculation.

The amine content in the outlet gas varies with the operational conditions, e.g. flow rates.

Theoretical Study
Simulation program for the membrane contactor
It has been assumed that the liquid flows is of laminar type inside the ribbon tubes. The diameter of the tubes is typically of the order 1 mm and the linear velocity is in the range $0.5 - 5$ cm/s. With density and viscosity of the systems in question higher than for water, the Reynolds number for the flow is well below $100 - 200$. This makes the assumption of laminar flow reasonable.

It has further been assumed that the flow inside the tubes is symmetrical and can be described by a Hagen–Poiseuille profile. Implicit in this lies the assumption of constant viscosity in the fluid. This is not completely correct as the viscosity of the used aqueous amine systems increases with increased CO_2 loading. This means that the viscosity normally will increase toward the tube walls. However, a CFD-study has shown that the effect on the velocity profile is negligible (Figure 14).
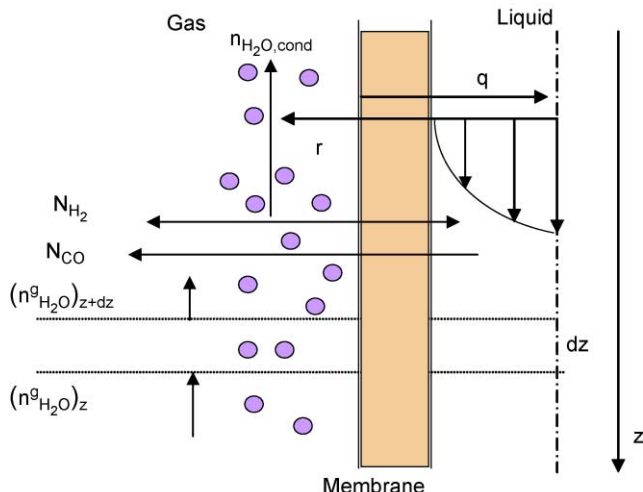
Figure 14: Flow model for the membrane absorber.

The gas flow has been assumed turbulent and counter- or co-current to the liquid flow. Typically the gas, in low-pressure applications, flows at linear velocities of a few m/s and the characteristic dimension, being the space between ribbon tube layers, is about $1 - 2$ mm. This gives a Reynolds number of about 2000 indicating that the flow may be in the transition region between laminar and turbulent flow. In high-pressure applications, the Reynolds number will normally be higher.

The assumption of counter- or co-current flow is made to make the modelling problem tractable. The model is thus not a direct representative of the often-used flow situation in a module, which is cross-current. However, normally the concentration changes in a single module are modest, and an average of a counter- and a co-current calculation will give a good estimate. The modules themselves are normally placed in a counter-current fashion [3]. The simulation tool is further described in a PhD thesis by Hoff [4] and in a paper from AIChE 2000 by Hoff et al. [5].

Comparison of the Test Results and the Simulations

Figure 15 shows the simulated and the experimental values calculated as recovery rate in kg CO_2 removed per hour for all the tests.

Nanko pilot unit, KS-1 and membrane contactor
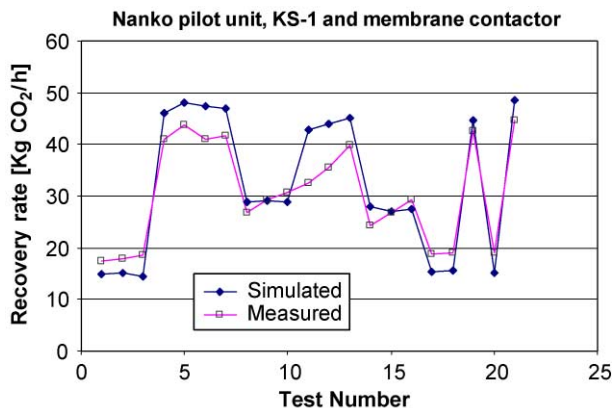
Figure 15: Simulated and measured values for the Nanko pilot unit tests.

A comparison between the simulated and experimental results shows that deviations from most of the experimental results are between 0 and 20% (two around 30%). It is also clearly seen that the discrepancies are systematic. At low partial pressures, the experimental results show a higher absorption rate than the simulations, and at high CO_2 partial pressure, the experiments give lower absorption rates than the simulations.

The experimental and simulation results for the tests at SINTEF are shown in Figure 16.

The analysed values for MEA content in the membrane inlet gas are used together with the measured outlet values. This method is in very good experimental agreement (dev. $<5\%$) in all cases apart from one experiment showing the lowest experimental mass transfer rate.

Figure 16 shows the amine rate (mol amine/hour) through the membrane as a parity plot with the simulation results on the X axis and the experimental results on the Y axis. If the model is 100% accurate, all the points would fall on the 45° line ($Y = X$). The deviation from the parity line ($Y = X$) shows that the model under predicts (more conservative) at higher driving forces (higher rates), but over predicts at low driving forces. Error bars indicate the uncertainty in the simulated results. As can be seen, apart from two points, all the others are within the experimental uncertainty.

Gas Side Pressure Drop

In a conventional process for separation of CO_2 from exhaust gas, a blower is necessary to overcome the pressure drop in the absorber. When designing a contactor, the gas pressure drop must thus be taken into account. Numerous tests and simulations are made to find the information needed for the design of a large-scale membrane contactor for exhaust gas treatment. The membrane used at Nanko is a typical large-scale laboratory module.

Table 3 shows the gas side pressure drop measured for the different membrane configurations, flow patterns and spacers.

A commercial membrane contactor unit for exhaust gas treatment will be designed with the new spacer design. Using the new spacer, the energy input to the blower will be reduced by 75% (membrane contactor including a membrane water wash). For a typical 28.5 MW gas turbine, the electricity consumption will be reduced by 450 kW (from 600 to 150 kW).
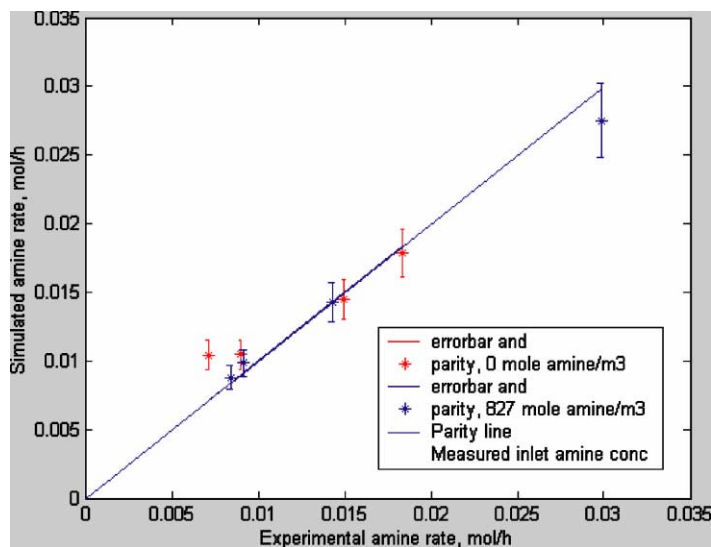
Figure 16: Parity plot with simulated and measured amine rates for the water wash tests at SINTEF. Experimental values for gas inlet are used for comparison.

TABLE 3
PRESSURE DROP ON THE GAS SIDE

Membrane module	Gas flow (Nm³/h)	Gas temperature (°C)	Pressure drop (mm WC, water column)	
			At test start	During test
Liquid on tube side, standard spacer (used at the pilot test in Japan)	150	30–37	8	9–13
Liquid on tube side new spacer	150	30–36	ND	ND
Mixing geometry, standard spacer	150	32–38	9	9–18

ND: not detected.

The effect of the reduced pressure drop is not included in the over all calculation of the process, and the reduced operation cost (more electricity will be available for sale) is not taken into account in the cost calculations.

Scale Up
Design basis (gas turbine exhaust, 350 MW)
The design basis and conditions are displayed in Table 4.

(1) Flue gas supply temperature of 270 °C is adopted as a typical temperature at the outlet of the waste heat boiler.
(2) CO_2 delivery pressure of 0.6 barg is adopted as a typical pressure at the outlet of a general CO_2 recovery facility.

TABLE 4
DESIGN BASIS FOR 350 MW

Power capacity (MW)	350
Flue gas flow rate (Nm^3/h)	1,766,779
Flue gas supply temperature (°C)	80
CO_2 concentration (mol%)	3.98
NO_x concentration (ppm)	15.0
CO_2 recovery (T/D)	2817
CO_2 recovery rate (%)	85
CO_2 delivery press. (barg)	153
CO_2 delivery temperature (°C)	50

Process flow sheet

A study of a conventional MEA process was conducted in conjunction with the respective cases for comparative purpose. The term "conventional MEA process" refers to a CO_2 capture process that utilises MEA for the absorbing solvent, and an absorbing tower as the medium where the absorption of CO_2 takes place. Both the solvent and absorption medium differ from the KPS/Kansai/MHI combined process. This study was conducted in order to verify the advantages borne from the use of the high-performing KS-1 solvent over MEA, and the use of the membrane contactor over the absorbing tower.

Main equipment

Basic difference in main equipment between the KPS membrane/KS-1 combined and the conventional process is the replacement of the absorber with the KPS membrane contactor module.

Cost estimate

The cost estimate of CO_2 capture and compression plant is based on the technical information included in this report and is subject to ± 30% of accuracy. This cost estimate was reached by using MHI's in-house cost data except cost of critical equipment for which vendor's quotation is applied, on a current cost basis. Shop prefabrication is considered for the flue gas cooler, the flue gas absorber (KPS G/L Membrane Contactor Module), ducts, pipe rack, structural steel (stage for heat exchangers), and the solvent storage tank to minimise on-site construction work at the field. The scope of cost estimate includes the compression section.

The summary of each category in the estimate is as follows:

Engineering: Includes the cost for basic and detail engineering work for the plant.
Procurement: Includes the cost of all the materials and equipment, assuming that they are procured on the world-wide basis.
Transportation: Includes the cost of ocean transportation of all materials and equipment to the nearest port of entry.
Site Construction & Commissioning: Includes the cost of all civil work, field installation and commissioning work.

Major assumptions. The following assumptions have been made in the preparation of the estimate:

1. International Standards such as API, ANSI, ASME, IEC are acceptable for design and manufacturing of all equipment and materials. No special requirements over such standards are considered.
2. Equipment and material vendors will be selected from world-wide sourcing. No special requirements for vendor selection are considered.
3. Piping prefabrication is assumed to be done at the field, therefore, cost of piping prefabrication is not included in the cost estimate.
4. Inland transportation from the nearest port to the construction site is excluded from the cost estimate.

Exclusions. The following items have been excluded from the estimate:

1. escalations;
2. import duties and taxes;
3. soil investigation;
4. fire fighting facilities;
5. lighting system;
6. handling system of waste from the Reclaimer (if required);
7. safety equipment, such as eye washer, body shower (if required);
8. SCADA/telecommunication system (if required);
9. operation spare parts.

Total capital cost
The total cost for both equipment and installation is displayed in Table 5. As it can be deduced from the table, the total capital cost for the combined process (using a conventional water wash) is lower than the conventional process using the MEA solvent.

TABLE 5
TOTAL COST COMPARISON BETWEEN COMBINED AND CONVENTIONAL
PROCESSES (INCLUDING CONSTRUCTION AND COMMISSIONING COST)
350 MW CASE

1 Million US$	Membrane /KS-1 process	Conventional /MEA solvent
Total	116.0	125.9

Operational cost
Utilities and chemical consumption will be the focus of operation cost study due to the fact that all other types of operational costs will have insignificant effects on the comparison between the KS-1/membrane combined case and the conventional MEA case.

Research and experience have shown that impurities contained in flue gas affect the rate of solvent loss. This effect is reduced through the replacement of the absorber with the membrane contactor. Although the extent of solvent loss reduction is not understood, we have hypothesised that the membrane contactor will reduce this effect by 1/3; this value is applied in the calculation of solvent loss for the membrane/KS-1 combined process.

All cost estimations in this study are based on the utilities unit cost below and the operation cost study was conducted based on the utilities consumption values listed in Table 6.

In comparison to the conventional MEA process, the KPS membrane/KS-1 combined process has a significantly reduced operation cost calculated both as annual operating cost and cost per tonne of CO_2 as shown in Table 10. The basis for one year period is 330 days at 90% production load.

The key constituents of operation cost reduction of the KS-1/Membrane process are the following:

Steam consumption: Reduction from lower regeneration energy requirement.
Electricity consumption: Lower pump electricity consumption from lower solvent flow rate.
Chemical consumption: Lower solvent make-up required (solvent cost correlated), NaOH consumption reduced from minimal solvent reclamation operation, land lower activated carbon requirement from reduced solvent flow rate.
Cooling (sea) water consumption: Reduced cooling requirement from lowered solvent flow rate and heat of reaction.

TABLE 6
UTILITIES CONSUMPTION

Utilities consumption	Unit	Membrane/ KS-1	Conventional/ MEA	Membrane/ KS-1 (US$/T-CO$_2$)	Conventional/ MEA (US$/T-CO$_2$)	Utilities unit cost
MP steam 37 barg, 370 °C	(T/h)	161.7	172.1	12.00	12.78	8.71 (US$/T)
LP steam 6 barg, 200 °C	(T/h)	19.3	64.2	0.84	2.79	5.11 (US$/T)
Electricity	(KWH/h)	1107	1324	0.38	0.45	0.04 (US$/KW)
Solvent	(kg/h)	33.41	288.70	1.94	4.57	
KS-1 solvent						6.50 (US$/kg)
MEA solvent						1.80 (US$/kg)
NaOH	(kg/h)	1.95	11.68			0.17 (US$/kg)
Activated carbon	(T/year)	25.9	38.3			3.04 (US$/kg)
Cooling sea water	(T/h)[a]	14,633	18,341	1.25	1.56	0.01 (US$/T)
Reclaimer waste handling	(kg/h)	33.4	577.4	0.04	0.69	140 (US$/T)
Total				16.45	22.84	

Total- and unit cost for 350 MW case; T is metric ton.
[a] $\Delta T = 10$ °C base.

Membrane Water Wash, 350 MW

The membrane water wash upscale calculations (simulations) are based on the simulation program (programmed in Matlab) verified by the tests. The design basis used for the two cases is given in Table 7 and the simulation results are shown in Table 8. Three cases were simulated:

both water wash (with recycled water) and polishing step (with fresh water);

TABLE 7
DESIGN BASIS AND MEMBRANE TYPE FOR MEMBRANE WATER WASH UPSCALE 350 MW

Design basis	Unit	Case		
		1	2	3
Gas flow rate	Nm^3/h	1,676,250	1,676,250	1,676,250
Liquid recycle flow rate	m^3/h	565	–	–
Liquid flow rate polishing/single step	m^3/h	19.5	19.5	10
Gas inlet temperature	°C	52.7	52.7	52.7
Fresh water inlet temperature	°C	38	38	38
Gas MEA inlet concentration	ppmv	124	124	124
CO$_2$ inlet concentration	vol%	0.5	0.5	0.5
H$_2$O inlet concentration	vol%	11.9	11.9	11.9
Water recycle MEA concentration	mol/m^3	330		
MEA concentration inlet fresh water	mol/m^3	0.02	0.02	0.02

polishing step only (with fresh water);
same as 2 but with only half of the fresh water flow rate.

See also explanations in Figures 17 and 18. An inlet MEA gas concentration of 124 ppm is used and an outlet concentration of 3 ppm. The outlet amine concentration can be dependent of economic considerations, environmental and governmental requirements.

TABLE 8
RESULTS FOR MEMBRANE WATER WASH UPSCALE 350 MW

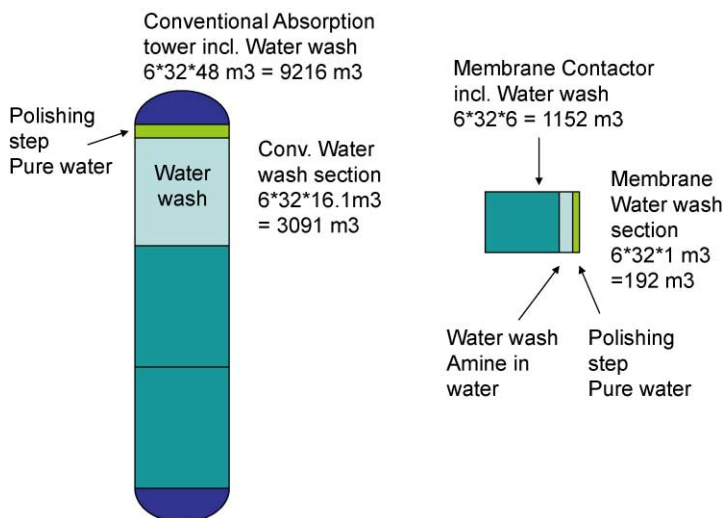| Results from Matlab	Unit	Case		
		1	**2**	**3**
Gas outlet temperature	°C	51.3	52.2	52.3
Liquid outlet Temperature	°C	51.7	51.8	52.8
Gas outlet MEA conc.	ppm	3.0	3.0	2.8
MEA liquid outlet loading	wt% amine	2.1	2.9	5.6
MEA recycle/outlet	mol amine/m^3	342	465	910
MEA recovery rate	T/D	13.3	13.3	13.3
Size	m^3	148	135	150

Figure 17: Upscale (350 MW) size of a conventional tower compared to a membrane contactor.

Cost estimate
Based on the calculation of size and weight of the conventional absorber, KPS has in cooperation with cost experts in AkerKvaerner and MHI estimated the installed cost of a conventional absorber and a membrane contactor. The cost estimate is based on fabrication in Norway and installation on an existing gas terminal on the south west coast of Norway. The membrane contactor will be fabricated in Germany and transported on trucks to the site. The membrane modules will then be installed in a prefabricated frame. The weight of each membrane module is approximately 10 tonnes.
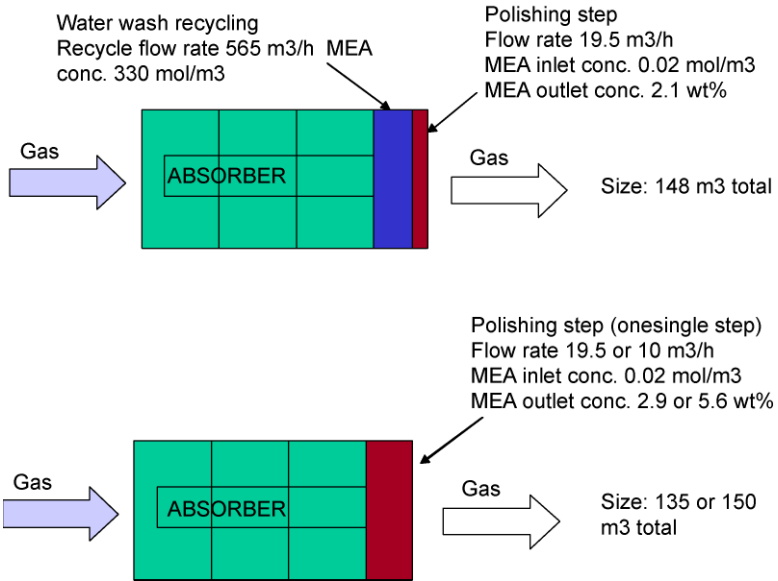
Figure 18: Illustration of the different upscale cases, water wash and polishing steps.

The conventional tower has a significant size ($6L$ m $\times$ $32W$ m $\times$ $48H$ m) and weight (1200 tonnes). Due to the significant size, AkerKvaerner would propose to construct the tower in segments at a yard, ship the segments to the site on barges and then mount the segments together at the site. The construction work and installation work at the site is a large operation. The cost estimation is based on experience from other projects in Norway. The construction and installation work is estimated to be 85 NOK/kg (9.50 USD/kg). The packing weight is estimated and the cost of the packing is 108 NOK/kg (12 USD/kg). The installation cost of the packing is estimated to be 50 NOK/kg (5.5 USD/kg). The total cost estimate of the absorber including water wash is shown in Table 9.

TABLE 9
COST, WEIGHT AND SIZE ESTIMATE FOR CONTACTORS INCLUDING WATER WASH FOR
350 MW UNIT

	Membrane contactor including water wash	Conventional absorber including water wash
Size (m)		
Length	6	6
Width	32	32
Height	6 (= 5 + 1 (ww))	48 (= 32 + 16 (ww))
Weight (tonnes)		
Frame	30	–
Tower and packing	–	1200
Membrane absorber	175	–
Membrane water wash	45	–
Total weight	250	1200
Total installed cost	15.72	20.63

CONCLUSIONS

The results of an upscale calculation proved that merits of respective technologies gave a synergetic effect for the combined process. (This is also the case for the equipment cost if the not yet proven membrane water wash is used. Due to this, the cost study for the membrane water wash is not a $\pm 30\%$ estimate.)

The cost, size and weight of equipment were relatively lower for this combined technology with membrane water wash, compared to a conventional CO_2 capture facility using the MEA solvent.

(1) Capital cost savings for the combined technology compared to conventional MEA technology:
    (a) Lower corrosion rates of the KS-1 solvent compared to MEA.
    (b) The cost, size and weight of equipment were considerably lower for the membrane water wash, compared to a conventional water wash unit.
(2) Installation cost savings for the combined technology compared to conventional MEA technology:
    (a) Smaller and lighter equipment, especially the membrane contactor and membrane water wash compared to conventional towers.
(3) Operational cost savings for the combined technology compared to conventional MEA technology:
    (a) Lower chemical consumption since KS-1 has a much lower degradation rate than MEA. Furthermore, the membrane contactor prevents direct gas to solvent contact, thereby reducing the absorption rate of impurities from the flue gas.
    (b) Lower circulation rate for KS-1 compared to MEA (due to higher loading capacity (CO_2 uptake) of the amine and smaller size of the membrane contactor compared to conventional technology).
    (c) Lower regeneration energy for KS-1 compared to MEA.
    (d) Reduced energy consumption in blower, due to reduced pressure-drop in the membrane absorber and membrane water wash.

In conclusion, the result of an upscale calculation of a combined process facility showed that costs were equal or lower for "equipment" and lower for installation and operation in contrast to a conventional MEA facility.

The weight and size is also reduced for a combined unit compared to a conventional unit.

The results and conclusions from the work described in this chapter, (the combined technology KS-1 and membrane contactor) is an important step to come closer to a more environmental solution for the gas and coal fired powerplants (Tables 10–12).

TABLE 10
THE COST OF THE MEMBRANE/KS-1 SOLVENT COMBINED PROCESS WAS FOUND TO BE LOWER THAN THAT OF CONVENTIONAL MEA PROCESS

Item	Unit	350 MW combined (membrane water wash)[a]	350 MW Conv. MEA
Flue gas flow rate	Nm^3/h	1,766,779	1,766,779
CO_2 recovery	T/D	2817	2817
Capital cost[b]	[mill US$]	116.0[a]	125.9
Operational cost	[US$/T-$CO_2$]	16.45	22.84
Operational cost	[MillionUS$/year]	13.774	19.134

[a]The cost study for the membrane water wash is not a $\pm 30\%$ estimate since the technology is not yet proven.
[b]Capital cost includes equipment, installation and commissioning costs.

TABLE 11

THE REDUCED SIZE AND WEIGHT OBTAINED BY USE OF A COMBINED PROCESS WITH A MEMBRANE WATER WASH SHOWS BENEFIT COMPARED TO A CONVENTIONAL MEA PROCESS; HOWEVER THIS DEPENDS ON CHOICE OF LOCATION

Service	Unit	350 MW combined	350 MW Conv., MEA
Flue gas flow rate	Nm^3/h	1,766,779	1,766,779
CO_2 recovery	T/D	2817	2817
Absorber	m	$7L \times 32W \times 5H$	$6L \times 32W \times 48H$
Absorber	tonnes	250	2300

TABLE 12

A MEMBRANE WATER WASH UNIT WAS TESTED IN THE NEXT STEP OF THIS PROGRAM. COST, SIZE AND WEIGHT OF THE MEMBRANE WATER WASH PROCESS WERE FOUND TO BE MUCH LOWER COMPARED TO A CONVENTIONAL WATER WASH PROCESS. THE MEA INLET CONCENTRATION IS 124 PPMV AND THE OUTLET IS 3 PPMV

Item	Unit	350 MW membrane water wash	350 MW conventional water wash	Savings membrane compared to conventional
Treated gas flow rate	Nm^3/h	1,676,250	1,676,250	
Amine recovery	T/D	13.3	13.3[a]	
Capital cost[b]	[mill US$]	3.1	6.9	55%
Water wash unit size	m^3	$6L \times 32W \times 1.0H$[c]	$6L \times 32W \times 16.1H$	94%
Water wash unit dry weight	tonnes	45	400[d]	89%

[a] Value calculated by Mitsubishi (MHI), value not confirmed, but assumed to be the same.
[b] Capital cost includes equipment, installation and commissioning cost.
[c] Demister and polishing step added to the size ($= 192 \ m^3$, conservative value. Largest simulated case $= 150 \ m^3$).
[d] Total tower weight/3 $= 400$ tonnes.

NOMENCLATURE

N_i Mass transfer flux across membrane (mol/m^2, s)
n_i Mass flux along membrane (mol/m^2, s)
r Variable radius (m)
r_{CO_2} Reaction rate for CO_2 (mol/L, s)
z Variable length (m)

Subscripts
i Component i (H$_2$O and MEA)

Superscripts
g Gas phase

ACKNOWLEDGEMENTS

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