BENCH-SCALE DEVELOPMENT OF A HYBRID MEMBRANE-ABSORPTION CO_2 CAPTURE PROCESS: PRELIMINARY COST ASSESSMENT

Budget Period 1 Topical Report
Reporting Period: 10/01/2013 to 3/31/2014

submitted by

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ABSTRACT

This report describes a study of capture costs for a hybrid membrane-absorption capture system based on Membrane Technology and Research, Inc. (MTR)'s low-pressure membrane contactors and the University of Texas at Austin's 5 m piperazine (PZ) Advanced Flash Stripper (AFS; 5 m PZ AFS) based CO_2 capture system.

The report is submitted for NETL review, and may be superseded by a final topical report on this topic that will be submitted to satisfy the Task 2 report requirement of the current project (DE-FE0013118).

EXECUTIVE SUMMARY

This report describes results from a preliminary study of capture system costs. The project team developed capital and operating costs for two version for each of two hybrid configurations (hybrid-series and hybrid-parallel) and identified design variables which have the strongest influence on the overall cost of capture.

Capital costs for the 5 m PZ AFS CO_2 capture plant were provided by UT Austin. These costs were first developed as part of a techno-economic cost study in a separate UT Austin project (DE-FE0005654), which compared the 8 m PZ advanced flash stripper (AFS), short stripper (SS), and two-stage flash (2SF) configurations applied to the NETL Case 11 supercritical power plant. In that work, the equipment costs were developed for a NETL Case 11 plant scaled to 593 MWe gross electrical output. This work uses the same 593 MWe gross size in this analysis but will be rescaled to a 550 MWe net electrical output for the final techno economic analysis in BP3. For this project (DE-FE0013118), UT Austin adjusted and recalculated cost values to account for the change from 8 m to 5 m PZ concentration and for the process modifications required when applied to the hybrid-series and hybrid-parallel configurations.

Four cases were modeled in detail. The motivation for the hybrid capture design is to use the membrane to create improved capture conditions (higher inlet CO_2 concentration, reduced volume of flue gas (hybrid-parallel), and reduced removal requirements (hybrid-series)). These benefits primarily occur on the absorber (CO_2 capture) portion of the 5 m PZ AFS plant.

BENCH-SCALE DEVELOPMENT OF A HYBRID MEMBRANE-ABSORPTION CO_2 CAPTURE PROCESS

Budget Period 1 Preliminary Cost Assessment

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INTRODUCTION

The overall goal of this three-year project is to evaluate two variations of a hybrid membrane-absorption capture system combining the MTR cross-flow, air-swept Polaris membrane technology, which enriches flue gas to ~20% CO_2, with the UT Austin 5 m piperazine Advanced Flash Stripper (5 m PZ AFS) technology. MTR is being assisted in this project by UT Austin, which is performing characterization, optimization studies and testing of their 5 m PZ AFS system configured for operation in hybrid CO_2 capture applications.

This project builds upon recent MTR work funded by DOE NETL (DE-NT0005312, DE-FE0006138 and DE-FE0007553). In those programs, new MTR Polaris^{TM} membranes have been developed that are ten times more permeable to CO_2 than conventional membranes. A novel process design that uses incoming combustion air as a sweep stream to generate driving force for CO_2 capture has been designed, and demonstrated in slipstream field tests at a coal-fired power plant. And a low pressure membrane contactor has been developed for use in the high-gas-flow, low-pressure sweep operation.

This new project is focused on the evaluation, development and testing of a hybrid membrane-absorption CO_2 capture system for coal-fired power plant applications using MTR's high-gas-flow, low-pressure-drop membrane contactor. The ability of the membrane contactor to enrich flue gas from its normal content of 13-15% CO_2 into a smaller volume of gas containing 20-25% CO_2 – with minimal energy input – could reduce the cost of the final concentration process, regardless of whether it is absorption, adsorption, cryogenic, or membrane-based. In this project, we are evaluating the feasibility of hybrid systems combined with UT Austin's 5 m PZ AFS capture system.

This report describes a study of capture costs for a hybrid membrane-absorption capture system based on Membrane Technology and Research, Inc. (MTR)'s low-pressure membrane contactors and the University of Texas at Austin's 5 m piperazine (PZ) Advanced Flash Stripper (AFS; 5 m PZ AFS) based CO_2 capture system. The

The 5 m PZ AFS system costs were first developed by URS, Trimeric and UT Austin in the NETL sponsored R&D Project #DE-FE0005654. The CO_2 capture system in this project was UT Austin's 2 stage flash capture process using 8 m PZ. UT later updated these costs for this project (DE-FE0011318) for hybrid conditions and reduced 5 m solvent concentration.

The following section is excerpted directly from a UT Austin report (Frailie, 2014) which explains the method for developing the equipment costs used in the hybrid analysis for the 8 m PZ AFS system from which the 5 m PZ AFT hybrid costs are derived.

Hybrid System Integration

This analysis describes the methods for estimating the purchased equipment cost of major cost centers. Understanding the link between operating conditions and the capital cost of equipment is necessary to minimize the cost of CO_2 avoidance.

Method for Developing Cost

Scaling

The same methods employed in the 2012 DOE NETL Report for derating power generation and CO_2 capture facilities were employed in this study. The CO_2 source is a 550 MW_e supercritical pulverized coal power plant described in 2010 DOE Cost and Performance Baseline Case 11. Case 12 modifies the plant in Case 11 to be fit with the Econamine scrubbing system while maintaining 550 MW_e net production.

Derating

In addition to the 550 MW_e output of the power plant, the total steam turbine power includes the equivalent work of the steam heater or reboiler, compression, capture auxiliaries, and a balance of plant auxiliaries. The following is a summary of how each was calculated.

- <u>Compression</u> – The compression work values in the 2012 DOE NETL report were calculated using vendor quotes. The compression work in this study is calculated by scaling Aspen Plus® predictions for compressor inlet pressures to those in the 2011 DOE NETL Report. The compression work is first calculated using Equations 2.1 and 2.2 developed by Van Wagener (2011) in Aspen Plus® with 75% polytropic efficiency, intercooling of wet gas to 40 ^oC at a maximum compression ratio of 2.0 per intercooling stage, and no allowance for pressure drop through the intercoolers. P_{in} is the inlet pressure to the compression train in atmospheres.

$$W_{comps}(kJ/mol\ CO_2) = 4.572\log\left(\frac{148}{P_{in}(atm)}\right) - 4.096 \qquad P_{in} \leq 4.5atm \qquad \text{Equation 1}$$

$$W_{comps}(kJ/mol\ CO_2) = 4.023\log\left(\frac{148}{P_{in}(atm)}\right) - 2.181 \qquad P_{in} > 4.5atm \qquad \text{Equation 2}$$

  Using data from steam and equipment tables in the 2012 DOE NETL Report, the compression work was calculated for the PZ-SS and PZ-HT cases and compared to the compression work estimated by vendors. It was determined that the vendor estimates for compression work were categorically 20% less than those calculated using Equation 1 and Equation 2. To ensure consistency, compression work for the new configuration is calculated by Equation 1 and Equation 2 and reduced by 20%. This value is expressed as a percentage of the gross plant power and scaled directly when converting from CO_2 captured to CO_2 avoided.

- <u>Capture auxiliaries</u> – This includes the pump work and blower work required to overcome pressure drop and reach regeneration temperature. These works were all calculated using proprietary software. For this study the percentage of total electrical production devoted to capture auxiliaries is set equal to that of the other 8 m PZ cases, which is approximately 2.6%.

- Balance of plant auxiliaries – The percentage of total power plant electrical production devoted to plant auxiliaries was calculated in the 2010 DOE Cost and Performance Baseline Case 12 to be 5.3%.

Because each of these additional electrical requirements can be calculated directly from the total steam turbine power, the total steam turbine power can be calculated directly from the desired net electrical output.

Scaling Purchased Equipment Cost (PEC) to 2010 DOE Cost and Performance Baseline

Because of the lack of information in the 2010 DOE Cost and Performance Baseline concerning the calculation of PEC, factors were derived in the 2012 DOE NETL report to establish a common basis for the cost estimation methods. The two goals of this study were (1) to develop exponents that would allow for the calculation of PEC for the capture and compression plants as a function of total power plant electrical capacity, and (2) to determine the relative difference between 2010 DOE Cost and Performance Baseline and 2012 DOE NETL PEC estimates. The first goal was accomplished by calculating the PEC for a base case of 593 MW_e total power, assigning each component of the PEC an exponent on the basis of expected scalability, and weighting those exponents by the base case PEC to give a single exponent that represents the entire process. Separate exponents were calculated for capture and compression systems. The second goal was accomplished by replicating Case 12 using in-house costing methodology and comparing it to the DOE reported values. This study replicates this method and calculates its own scaling exponents for capture and compression.

Calculating PEC

A major goal of this study is to improve the methods for estimating PEC using Aspen Plus® predictions. Emphasis has been placed on accurately calculating the PEC of the major cost centers, which were determined to be the absorber, cross exchangers, reboiler or steam heaters, and the compressor. These process units can account for 80% of the plant PEC.

Absorber

The purchased equipment cost of the absorber was calculated by developing expressions for each column component that could then use Aspen Plus® predictions to estimate a total column price. Pricing information from Sulzer reported in Tsai (2010). Costs for the 304 SS baffle distributors and supports are from (Pilling, 2009), and packing (Pilling, 2008) is used to estimate the cost of column internals. Equation 3 calculates the total cost of distributors and their supports as a function of column diameter, D, and Equation 4 calculates the cost of packing per cubic meter as a function of specific area, a_p.

$$Distributor\ Purchased\ Cost\,(\$) = 7929(D)^{1.6031} \qquad \text{Equation 3}$$

$$Packing\ Cost\ (\$/m^3) = 12.14 * a_p + 337.15 \qquad \text{Equation 4}$$

It should be noted that Equation 3 and Equation 4 include factors to convert predictions from 304 SS to 316 SS. Shell price was estimated using Equation 5 from Peters, Timmerhaus, and West (5th Ed., 2003) for a 2 cm shell made of 316 SS as a function of shell mass, M, in kg.

$$Shell\ Cost\,(\$) = 10^{0.657 \times \log(M) + 2.65} \qquad \text{Equation 5}$$

In addition to the CO_2 absorption section, a 3 m water wash is assumed to be part of the column. The absorber intercooler is a plate-and-frame heat exchanger with 316 stainless steel plates. The SO_2 polisher is assumed to be part of the power plant, per the 2010 DOE Cost and Performance Baseline.

Rich-Lean Heat Exchangers

Pricing and performance information for plate-and-frame heat exchangers was obtained equipment vendors. Heat duties and log mean temperature differences predicted by Aspen Plus® can be used to calculate the required heat transfer area for a given overall heat transfer coefficient. The areas in this study are calculated using overall heat transfer coefficients provided by equipment manufacturers. There are, however, several methods for calculating overall heat transfer coefficients using Aspen Plus® predicted fluid properties in conjunction with the heat duties and temperature differences. This first requires the calculation of heat transfer coefficients for the liquids and the exchanger material, in this case 316 SS. Equation 6 from Hewitt et al. (1994) can be used to calculate the heat transfer coefficient for a liquid in the turbulent regime.

$$\alpha = \frac{0.4\lambda}{D_l}(\mathrm{Pr})^{0.4}(\mathrm{Re})^{0.64} \qquad \text{Equation 6}$$

In Equation 6 α is the heat transfer coefficient, λ is the thermal conductivity of the liquid, D_l is two times the plate spacing, Pr is the Prandlt Number, and Re is the Reynolds Number. Re and Pr are dimensionless numbers calculated using Equation 7 and Equation 8, respectively.

$$\mathrm{Re} = \frac{\rho V D_l}{\mu} \qquad \text{Equation 7}$$

$$\mathrm{Pr} = \frac{C_P \mu}{\lambda} \qquad \text{Equation 8}$$

In Equation 7 and Equation 8, ρ is the fluid density, V is the bulk fluid velocity, μ is the fluid viscosity, and C_P is the fluid heat capacity. Aspen Plus® predictions can be used to calculate each of these terms for the hot and cold ends of the cross exchanger, which can in turn be used to calculate the overall heat transfer coefficient, U, for either end of the cross exchanger by Equation 9.

$$\frac{1}{U} = \frac{1}{\alpha_h} + \frac{1}{\alpha_c} + \frac{1}{\lambda_p} \qquad \text{Equation 9}$$

In Equation 9 α_h is the heat transfer coefficient of the hot liquid, α_c is the heat transfer coefficient of the cold liquid, and λ_p is the thermal conductivity of the plate at the mean temperature. The hot end and cold end heat transfer coefficients, U_h and U_c, their respective temperature differences, ΔT_h and ΔT_c, and the heat duty, Q, predict a heat transfer area, A, by Equation 10.

$$A = \frac{Q}{\dfrac{\left(U_h \Delta T_c - U_c \Delta T_h\right)}{\ln\left(U_h \Delta T_c \big/ U_c \Delta T_h\right)}} \qquad \text{Equation 10}$$

Reboiler or Steam Heater

This study uses the method described in the 2012 DOE NETL report. An Aspen Plus®-predicted heat duty, a heat transfer coefficient consistent with a 2007 report submitted to SBIR (Fisher, 2007), and a 5 K log mean temperature difference (LMTD) are used to calculate the required heat transfer area for the convective steam heater that supplies the heat for regeneration. The heat transfer area is then used to determine the price of the process unit by scaling to the price of a high pressure convective steam heater predicted by PDQ$. A similar method was used to calculate the price of a reboiler with a comparable heat duty and temperature approach, and it was determined that the convective steam heater is less expensive.

Compressor

Compressor prices are scaled on the basis of inlet CO_2 vapor volume in MMSCFD and power consumption in MW to vendor quotes confirmed by proprietary software. Aspen Plus® calculates the energy required to compress each mole of CO_2 from the regeneration pressure to 15 MPa using Equations 2.1 and 2.2. This value is multiplied by the CO_2 removal rate to give the power consumption of the compressor train. The resulting price is assumed to include a skid package including electric motors, interstage coolers, and interstage separators. Pumps and dehydration units are priced separately by scaling to vendor quotes used in the 2007 SBIR Advanced Amine report.

All Other Process Units

Inlet gas blowers, centrifugal pumps, water-cooled heat exchangers, filters, tanks, and the reclaimer are all sized and priced using vendor quotes from the 2012 DOE NETL Report. Most of these process units are priced on the basis of vapor and/or liquid flow rates. The stripper is priced as two separate units: (1) a pressurized flash vessel, and (2) a packed column that promotes interaction between the vapor from the flash vessel and rich solvent from the cold rich bypass (CRB) and warm rich bypass (WRB). The flash vessel is priced using vendor quotes from the 2012 DOE NETL Report, and the packed section is priced using the same methodology used to price the absorber. Combined, these process units account for less than 20% of the final PEC.

Advanced Flash Stripper with Intercooled Absorber

This study proposes a base-case absorber design that tests both pump-around and in-and-out intercooling. Pump-around intercooling removes semi-rich solvent from one point in the column, cools it to 40^{o}C, and feeds the cooled solvent back at both a higher point in the column and just below the point that it was removed (Figure 1). Enough liquid is fed to the lower stage to avoid accumulation of solvent. This effectively splits the column into three sections: (1) a top section where lean solvent enters and scrubbed gas leaves, (2) a middle section containing 2–5 times more solvent than the top section, and (3) a bottom section containing the same amount of liquid as the top section from which the rich solution exits and the flue gas enters. A coarser packing must be used in the middle section to avoid excessive pressure drop from the higher liquid load. In-and-out intercooling removes all of the semi-rich liquid from the bottom of the top section, cools it to 40^{o} C, and feeds it to the top of the bottom section, effectively eliminating the middle section in the pump-around case. There are tradeoffs associated with lean loading, feed liquid flowrate, packing area, and pump-around rate (i.e., the amount of liquid fed back to the top of the second section). The absorber diameter is adjusted to achieve 70% flood in the bottom section.

Integrated Hybrid System Design

Two variations of each hybrid design (hybrid-series and hybrid-parallel) were investigated – see Figures 1 and 2.

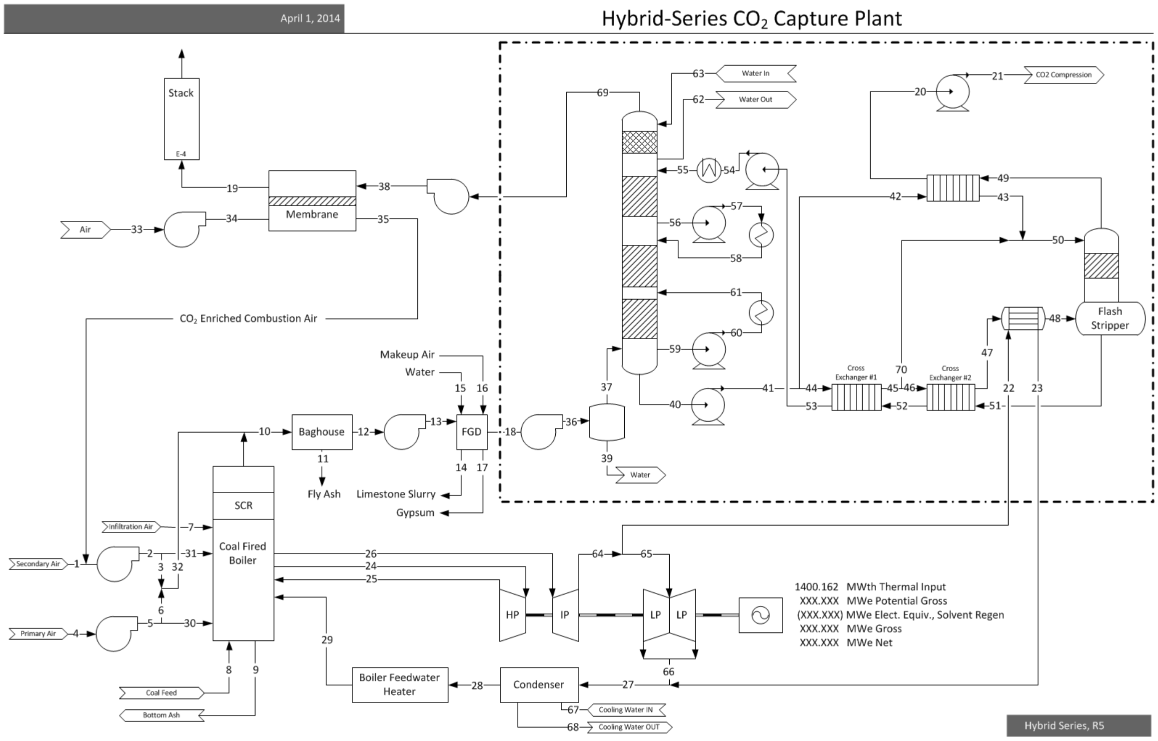

Figure 1: Process diagram of the Hybrid-Series capture system

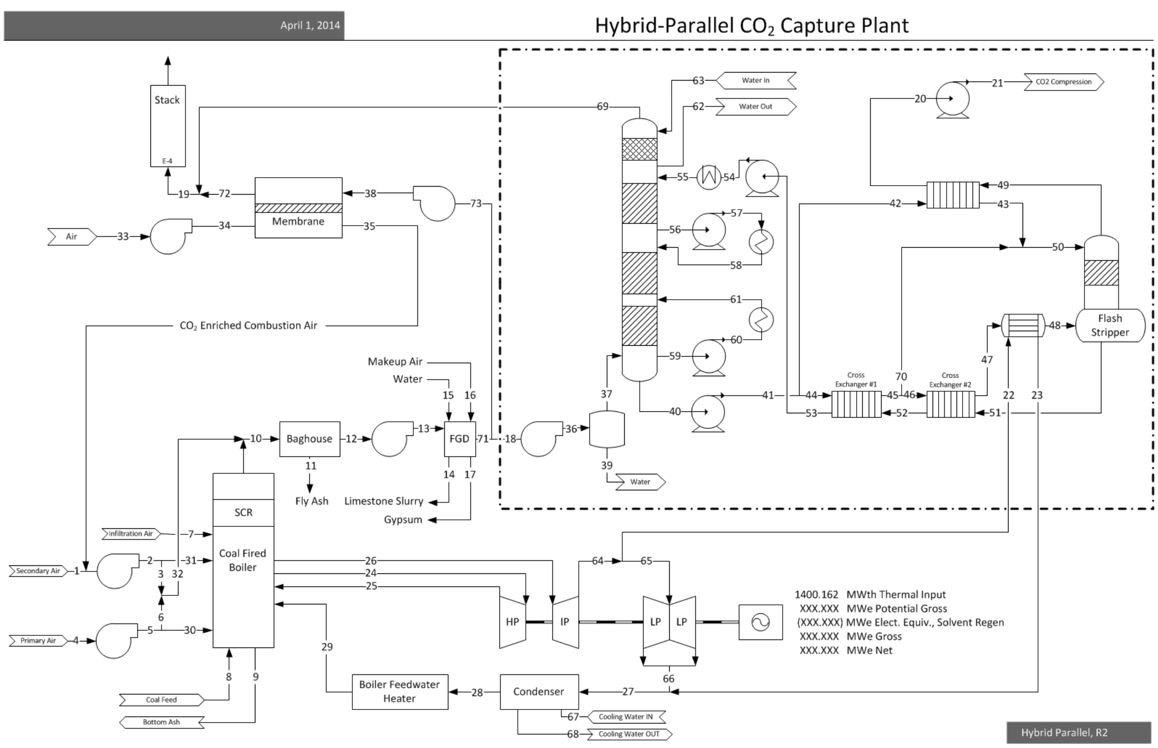

Figure 2: Process diagram of the Hybrid-Parallel capture system, Case #18 and Case #19

The focus of this initial cost assessment was on the capital cost (purchased equipment cost) of the 5 m PZ AFS capture plant. The process flow diagram shown in Figure 3 identifies the major components and steams which are detailed in Tables 1, 2, 3 and 4.

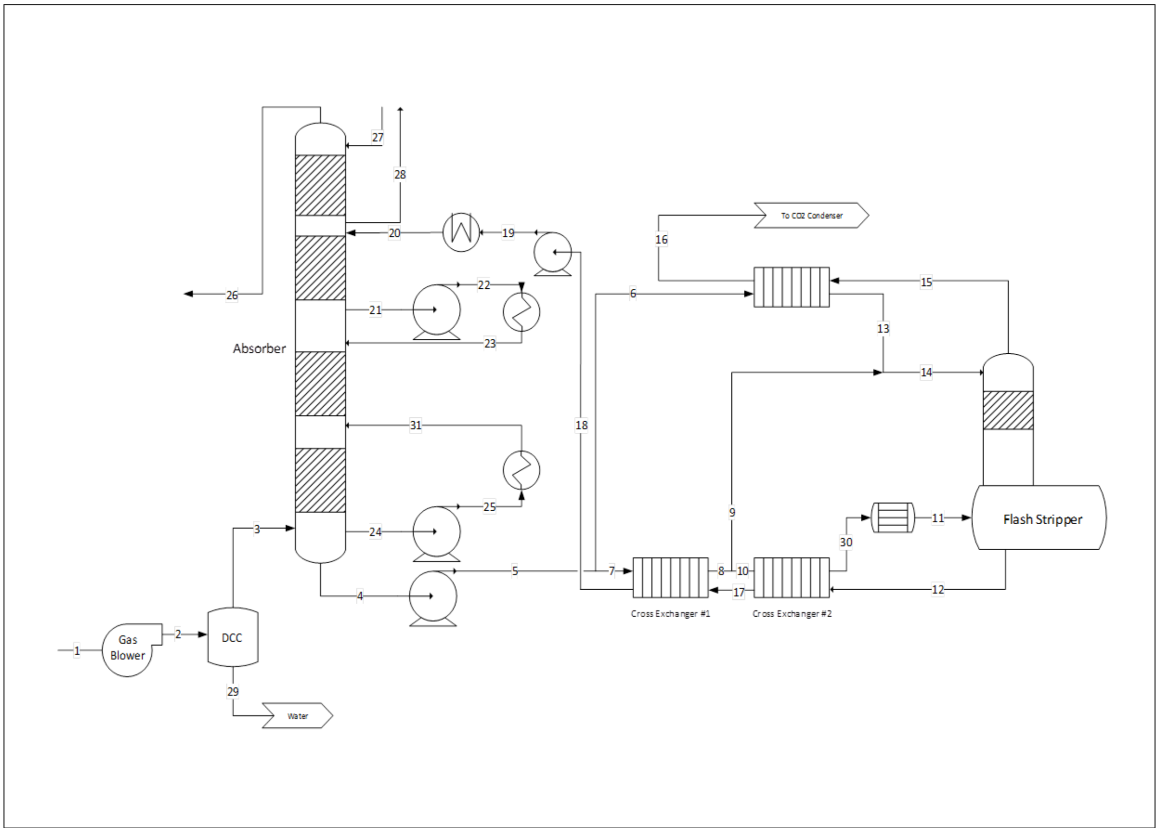

Figure 3: Process flow diagram for proposed advanced flash stripper with intercooled absorber

Table 1 Process stream properties for the 5 m PZ AFS Plant in Case 13, hybrid-series configuration (593 MWe gross)

Stream Name	20*	4*	26	3	28	27	20	19	18	12	17	15
ASPEN Stream Name	**ABSLEAN**	**ABSRICH**	**G-OUT**	**GASIN**	**L-OUT**	**PUREWATE**	**LEAN**	**LEANCLD2**	**LEANCOLD**	**LEANHOT**	**LEANWRM**	**PRODUCT1**
Temperature C	40.00	42.21	40.00	40.00	40.00	40.00	40	47.6858163	47.7333159	150	121.07547	122.12122
Pressure kPa	101.325	101.325	101.325	120.000	101.325	101.325	100	100	837.173325	837.173325	837.17333	837.17333
Vapor Frac	0	0	1	1	0	0	0	0	0	0	0	1
Mole Flow kmol/min	8168.421	8161.387	1256.404	1404.600	4201.351	4200.000	8114.8611	8114.86693	8114.86698	8118.69313	8115.8828	200.1322
Mass Flow kg/min	210316.260	216967.082	36039.119	42719.043	75693.274	75664.179	209359.88	209359.883	209359.883	209359.883	209359.88	7607.3452
Volume Flow cum/min	194.440	196.227	32252.899	30431.157	76.283	76.258	193.45893	194.110575	194.064135	208.503751	203.34761	766.43888
Mass Flow kg/min												
H2O	134826.478	134377.013	1666.440	1556.210	75685.021	75664.175	133863.69	133806.382	133805.998	132977.073	133111.54	832.85353
CO2	0.327	9.706	4515.674	11303.739	6.534	0.000	0.3250801	0.58273777	0.58498478	168.971962	45.292105	6769.4348
MEA							0	0	0	0	0	0
MEA+							0	0	0	0	0	0
MEACOO-							0	0	0	0	0	0
HCO3-	607.151	1685.245			0.289		603.1468	797.247674	798.549162	3606.11527	3150.6878	0
PZ	5884.763	563.650	9.97E-10		1.46E-05	0.004	5852.14	6183.40426	6186.57843	11319.9071	9883.7366	3.9856712
PZCOO-2	10641.102	14270.678			3.71E-12		10635.443	9442.7448	9432.25355	1446.31305	2408.0537	0
PZCOO-	19776.500	5159.617			1.34E-08		19787.823	20186.6853	20189.8035	18978.7985	20047.555	0
PZH+	24986.181	20337.126			0.4122103		24982.374	24321.1696	24314.5104	19421.4712	20465.982	0
PZH+2							0	0	0	0	0	0
HPZCOO	13593.762	40562.976	1.52E-13		0.00024603		1.36E+04	14621.6646	14631.6031	21441.2312	20247.037	2.66E-07
N2		1.005	28841.913	28843.873	0.954		1.24E-03	1.24E-03	1.24E-03	1.24E-03	1.24E-03	1.01E+00
O2		0.066	1015.091	1015.221	0.064		1.16E-04	1.16E-04	1.16E-04	1.16E-04	1.16E-04	6.39E-02

Stream Name	16	4	13	6	5	7	11	14	9	8	10
ASPEN Stream Name	**PRODUCT2**	**RICH**	**RICHBPS2**	**RICHCBPS**	**RICHCLD2**	**RICHCOLD**	**RICHHOT**	**RICHTTL**	**RICHWBPS**	**RICHWRM1**	**RICHWRM2**
Temperature C	64.6339778	42.21	110.715521	42.3552677	42.35522	42.35527	140.9384	115.43912	117.151598	117.151598	117.1516
Pressure kPa	837.173325	101.325	837.173325	837.173325	837.1733	837.1733	837.1733	837.17333	837.173325	837.173325	837.1733
Vapor Frac	0.79330998	0	0	0	0	0	0.019783	0	0	0	0
Mole Flow kmol/min	200.085985	8161.3791	326.615766	326.455216	8161.38	7834.925	7022.846	1224.9407	898.329143	7839.96343	6941.634
Mass Flow kg/min	7607.34516	216967.07	8678.68275	8678.68275	216967.1	208288.4	184422	32545.062	23866.3776	208288.386	184422
Volume Flow cum/min	517.185838	196.22737	8.14834653	7.84756805	196.1892	188.3416	731.9433	30.658521	22.5106633	196.456697	173.946
Mass Flow kg/min											
H2O	832.144832	134311.51	5322.22347	5372.37325	134309.3	128937	113476.9	19949.37	14627.3598	127656.958	113029.6
CO2	6767.40085	9.3880046	7.44348041	0.37776196	9.445106	9.066287	3778.455	33.705062	26.4455673	230.797678	204.3521
MEA	0	0	0	0	0	0	0	0	0	0	0
MEA+	0	0	0	0	0	0	0	0	0	0	0
MEACOO-	0	0	0	0	0	0	0	0	0	0	0
HCO3-	2.40037132	1907.1109	246.436294	76.5791703	1914.538	1837.9	3950.855	954.51059	707.35195	6173.25337	5465.901
PZ	3.74E-03	5.73E+02	87.9264133	23.0318598	575.7739	552.7646	6141.84	350.61728	262.805668	2293.57673	2030.771
PZCOO-2	1.22E-04	1.42E+04	197.431025	565.190876	14130.29	13564.58	1965.824	681.16479	484.794674	4230.93535	3746.141
PZCOO-	1.41E-03	5.23E+03	409.181982	209.851511	5245.939	5036.436	14583.93	1564.5165	1154.90641	10079.1832	8924.277
PZH+	3.42919111	20586.007	827.961351	823.213286	20580.7	19757.12	17474.11	3108.596	2280.38828	19901.5705	17621.18
PZH+2	0	0	0	0	0	0	0	0	0	0	0
HPZCOO	8.94E-01	40192.244	1580.03632	1608.02215	40200	38592.53	23049.18	5902.4211	4322.20864	37721.0936	33398.88
N2	1.00724326	1.0084852	4.03E-02	4.03E-02	1.01E+00	0.968146	0.857212	0.1512728	0.11093338	0.96814582	0.857212
O2	6.39E-02	6.40E-02	2.56E-03	2.56E-03	6.40E-02	6.14E-02	5.44E-02	9.60E-03	7.04E-03	6.14E-02	5.44E-02

Table 2 Process stream properties for the 5 m PZ AFS Plant in Case 14, hybrid-series configuration (593 MWe gross)

Stream Name	20	4	26	3	28	27	20	19	18	12	17	15
ASPEN Stream Name	**ABSLEAN**	**ABSRICH**	**G-OUT**	**GASIN**	**L-OUT**	**WATER**	**LEAN**	**LEANCLD2**	**LEANCOLD**	**LEANHOT**	**LEANWRM**	**PRODUCT1**
Temperature C	40.00	40.67	40.00	40.00	40.00	40.00	40	46.3954465	46.509636	150	139.38725	131.97692
Pressure kPa	101.33	101.33	101.33	120.00	101.33	101.33	100	100	1846.52644	1846.52644	1846.5264	1846.5264
Vapor Frac	0	0	1	1	0	0	0	0	0	0	0	1
Mole Flow kmol/min	25531.99	25535.12	1256.32	1404.60	4191.87	4200.00	25506.513	25506.6101	25506.6121	25545.2148	25532.444	181.89798
Mass Flow kg/min	673708.14	680532.32	36036.88	42719.04	75522.14	75664.18	673247.39	673247.39	673247.42	673247.42	673247.42	7284.9141
Volume Flow cum/min	611.97	613.94	32250.69	30431.16	76.11	76.26	611.50021	613.069455	612.731542	655.660174	649.75444	316.71034
Mass Flow kg/min												
H2O	420772.249	420208.612	1666.329	1556.210	75514.400	75664.175	420315.35	420045.277	420040.5	415854.145	415928.67	498.31893
CO2	9.884	45.266	4516.027	11303.740	6.520		9.8631767	14.1241828	14.2115918	1713.09901	1151.0537	6781.7437
MEA							0	0	0	0	0	0
MEA+							0	0	0	0	0	0
MEACOO-							0	0	0	0	0	0
HCO3-	4102.222	6145.490			0.082		4096.1121	5010.83983	5027.01807	19206.1865	18953.783	0
CO3--							0	0	0	0	0	0
H+							0	0	0	0	0	0
KHCO3(S)							0	0	0	0	0	0
PZ	3630.004	1155.466	8.24E-11		1.20E-06	0.004	3629.2835	4139.761	4153.5116	15052.5712	13615.803	1.3454627
PZCOO-2	46373.946	43731.528			2.48E-14		46365.859	42745.3226	42654.2628	7569.4483	9008.0739	0
PZCOO-	26223.752	12121.662			3.16E-10		26230.44	28193.7067	28241.6598	43206.7299	42632.326	0
PZH+	70506.655	61233.224			0.118		70494.256	69459.7862	69423.0554	64249.4995	64957.975	0
HPZCOO	102089.420	135887.516	1.25E-14		2.03E-05		102106.18	103638.526	103693.123	106395.694	106999.69	1.46E-07
N2		3.332	28839.590	28843.874	0.952		0.0429057	0.04290571	0.04290571	0.04290571	0.0429057	3.2887188
O2		0.221	1014.936	1015.221	0.063		0.0041592	0.00415923	0.00415923	0.00415923	0.0041592	0.2173013

Stream Name	16	4	13	6	5	7	11	14	9	8	10
ASPEN Stream Name	PRODUCT2	RICH	RICHBPS2	RICHCBPS	RICHCLD2	RICHCOLD	RICHHOT	RICHTTL	RICHWBPS	RICHWRM1	RICHWRM2
Temperature C	47.392419	40.67	111.272655	41.0157129	41.01585	41.01571	144.2977	131.02787	134.908878	134.908878	134.9089
Pressure kPa	1846.52644	101.325	1846.52644	1846.52644	1846.526	1846.526	1846.526	1846.5264	1846.52644	1846.52644	1846.526
Vapor Frac	0.85298329	0	0	0	0	0	0.004259	0	0	0	0
Mole Flow kmol/min	181.882362	25535.093	255.508263	255.351048	25535.11	25279.75	24120.26	1534.1277	1278.69159	25318.0933	24039.4
Mass Flow kg/min	7284.91413	680532.29	6805.32288	6805.32288	680532.3	673727	639700.4	40831.94	34026.6145	673726.965	639700.4
Volume Flow cum/min	207.093378	613.94269	6.3752602	6.13658782	613.6588	607.5222	798.05	38.809699	32.4397067	642.306193	609.8665
Mass Flow kg/min											
H2O	498.070109	420067.68	4159.31717	4200.5091	420050.8	415850.4	390930.4	24921.379	20765.0503	411147.995	390382.9
CO2	6781.05636	44.155151	7.36564003	0.44669004	44.67759	44.22231	5202.622	91.640349	87.4508455	1731.52675	1644.076
MEA	0	0	0	0	0	0	0	0	0	0	0
MEA+	0	0	0	0	0	0	0	0	0	0	0
MEACOO-	0	0	0	0	0	0	0	0	0	0	0
HCO3-	0.8427511	6622.7635	206.312569	66.7956307	6679.979	6612.767	19547.1	1354.8098	1138.37375	22539.8003	21401.43
O2	0	0	0	0	0	0	0	0	0	0	0
PZ	0.00012567	1171.5567	58.1251779	11.8837707	1188.288	1176.493	11083.12	459.85584	403.956259	7998.33394	7594.378
PZCOO-2	4.84E-06	43510.476	151.527957	433.368828	43338.74	42903.51	8097.492	649.40164	507.892786	10056.2772	9548.384
PZCOO-	4.38E-05	12260.108	286.075993	123.666255	12365.39	12242.96	36311.67	1883.9805	1596.3713	31608.1518	30011.78
PZH+	1.20361737	61784.463	641.110213	617.617038	61763.34	61144.09	60618.17	3863.7044	3217.24567	63701.4642	60484.22
HPZCOO	0.23510426	135067.53	1295.45307	1351.00004	135097.5	133749	107906.5	7606.9562	6310.09847	124939.95	118629.9
N2	3.28871884	3.3316226	0.03331622	0.03331622	3.331623	3.298306	3.131725	0.1998974	0.16658113	3.29830635	3.131725
O2	0.21730125	0.2214605	0.0022146	0.0022146	0.22146	0.219246	0.208173	0.0132876	0.01107302	0.21924588	0.208173

Table 3 Process stream properties for the 5 m PZ AFS Plant in Case 18, hybrid-parallel configuration (593 MWe gross), 60% flue gas split ratio

Stream Name	20	4	26	3	28	27	20	19	18	12	17	15
ASPEN Stream Name	ABSLEAN	ABSRICH	G-OUT	GASIN	L-OUT	WATER	LEAN	LEANCLD2	LEANCOLD	LEANHOT	LEANWRM	PRODUCT1
Temperature C	40.0	44.8	40.0	40.0	40.0	40.0	40	50.6144107	50.649763	150	115.12018	115.90558
Pressure kPa	101.325	101.325	101.325	120.000	101.325	101.325	100	100	607.195138	607.195138	607.19514	607.19514
Vapor Frac	0	0	1	1	0	0	0	0	0	0	0	1
Mole Flow kmol/min	6232.118	6240.275	707.974	886.221	4794.664	4800	6178.543	6178.54522	6178.54523	6179.86498	6178.7804	237.27473
Mass Flow kg/min	157608.000	165478.470	19410.409	27186.333	86378.796	86473.347	156631.02	156631.019	156631.018	156631.018	156631.02	8847.1378
Volume Flow cum/min	147.628	150.037	18178.804	19198.907	87.057	87.152	146.64038	147.375589	147.351076	158.628215	153.77373	1239.8112
Mass Flow kg/min												
H2O	102925.202	102787.732	938.683	985.451	86375.989	86473.343	101962.55	101926.066	101925.923	101540.263	101635.66	1107.4481
CO2	0.071	6.491	77.670	7804.754	0.228		0.0693467	0.16581459	0.16634542	58.24802	10.515493	7732.3767
MEA							0	0	0	0	0	0
MEA+							0	0	0	0	0	0
MEACOO-							0	0	0	0	0	0
HCO3-	263.650	1216.861			0.540		260.29123	383.866199	384.352087	1690.5806	1367.4854	0
PZ	10032.412	569.664	4.91E-08		0.001	0.004	10023.46	10328.9622	10330.5942	14258.529	12926.284	6.5068094
PZCOO-2	4603.627	10729.904			9.98E-10		4577.6715	3793.21581	3790.05497	552.662284	1036.7626	0
PZCOO-	17788.151	4725.003			2.19E-06		17827.94	17958.6934	17958.6194	14668.8023	16162.924	0
PZH+	17041.115	15790.003			0.771		17036.889	16507.3724	16504.8162	12872.5766	13909.524	0
PZH+2							0	0	0	0	0	0
HPZCOO	4953.775	29652.006	2.28E-13		0.001		4942.1477	5732.67736	5736.49349	10989.3566	9581.8668	4.13E-07
N2		0.761	17829.487	17831.443	1.195		0.0002395	0.00023954	0.00023954	0.00023954	0.0002395	0.7611834
O2		0.045	564.569	564.685	0.072		2.08E-05	2.08E-05	2.08E-05	2.08E-05	2.08E-05	0.0449765

Stream Name	16	4	13	6	5	7	11	14	9	8	10
ASPEN Stream Name	**PRODUCT2**	**RICH**	**RICHBPS2**	**RICHCBPS**	**RICHCLD2**	**RICHCOLD**	**RICHHOT**	**RICHTTL**	**RICHWBPS**	**RICHWRM1**	**RICHWRM2**
Temperature C	71.1366476	44.82285	104.577011	44.9227939	44.92272	44.92279	143.5227	109.92233	111.053563	111.053563	111.0536
Pressure kPa	607.195138	101.325	607.195138	607.195138	607.1951	607.1951	607.1951	607.19514	607.195138	607.195138	607.1951
Vapor Frac	0.78453924	0	0	0	0	0	0.049564	0	0	0	0
Mole Flow kmol/min	237.199416	6240.268	436.962234	436.818806	6240.269	5803.45	3821.22	2497.1468	2060.18946	5805.98847	3745.799
Mass Flow kg/min	8847.13779	165478.46	11583.4924	11583.4924	165478.5	153895	99287.08	66191.389	54607.8929	153894.971	99287.08
Volume Flow cum/min	859.576934	150.03682	10.8488775	10.5011865	150.0169	139.5158	1149.077	62.221587	51.3734811	144.77981	93.40633
Mass Flow kg/min											
H2O	1106.31999	102720.03	7131.61322	7190.32065	102718.9	95528.55	61470.11	40728.402	33597.0677	94682.6452	61085.58
CO2	7729.06219	6.2165695	6.74935706	0.43717212	6.245256	5.808144	3395.074	48.237256	41.7045567	117.531023	75.82647
MEA	0	0	0	0	0	0	0	0	0	0	0
MEA+	0	0	0	0	0	0	0	0	0	0	0
MEACOO-	0	0	0	0	0	0	0	0	0	0	0
HCO3-	3.82086832	1446.1415	300.348192	101.506325	1450.081	1348.584	1416.065	1796.4564	1495.16501	4213.64685	2718.482
PZ	0.01513604	580.58075	121.213018	40.7634116	582.3367	541.571	5747.634	741.24002	620.151693	1747.70023	1127.549
PZCOO-2	0.00052988	10604.402	293.238219	741.303929	10590.05	9848.752	642.0203	1523.2529	1231.57945	3470.81481	2239.235
PZCOO-	0.00624328	4801.9814	570.757221	336.713918	4810.202	4473.485	9029.199	3331.3134	2759.83117	7777.706	5017.875
PZH+	5.46155866	16042.332	1110.98977	1122.7294	16038.97	14916.26	8765.355	6355.8614	5244.61947	14780.2912	9535.672
PZH+2	0	0	0	0	0	0	0	0	0	0	0
HPZCOO	1.64513554	29275.97	2048.52749	2049.66118	29280.9	27231.21	8821.138	11666.302	9617.51049	27103.8932	17486.38
N2	0.76118341	0.7614232	0.05329962	0.05329962	0.761423	0.708124	0.456854	0.3045693	0.25126964	0.70812354	0.456854
O2	0.04497654	0.0449974	0.00314981	0.00314981	0.044997	0.041848	0.026998	0.0179989	0.01484912	0.04184753	0.026998

Table 4 Process stream properties for the 5 m PZ AFS Plant in Case 19, hybrid-parallel configuration (593 MWe gross), 60% flue gas split ratio

Stream Name	20	4	26	3	28	27	20	19	18	12	17	15
ASPEN Stream Name	ABSLEAN	ABSRICH	G-OUT	GASIN	L-OUT	WATER	LEAN	LEANCLD2	LEANCOLD	LEANHOT	LEANWRM	PRODUCT1
Temperature C	40.0	42.4	40.0	40.0	40.0	40.0	40	47.6821322	47.7350373	150	122.102069	119.858747
Pressure kPa	101.325	101.325	101.325	120.000	101.325	101.325	100	100	914.045675	914.045675	914.045675	914.045675
Vapor Frac	0	0	1	1	0	0	0	0	0	0	0	1
Mole Flow kmol/min	10,008	10,016	727	904	4,795	4,800	9971.80009	9971.80932	9971.80941	9977.34779	9973.38441	217.287629
Mass Flow kg/min	258,621.358	266,402.294	20,064.210	27,755.180	86,383.371	86,473.346	257966.713	257966.713	257966.713	257966.713	257966.713	8435.96081
Volume Flow cum/min	238.466	240.802	18,677.699	19,573.758	87.061	87.152	237.803244	238.594582	238.531817	256.107225	250.033438	756.499862
Mass Flow kg/min												
H2O	165161.708	164909.622	964.522	1004.704	86380.427	86473.343	164514.799	164437.464	164436.891	163331.941	163496.424	781.261492
CO2	0.535	14.360	401.181	8049.382	1.144		0.53152945	0.93745324	0.94131973	244.68289	70.2559543	7650.78029
MEA							0	0	0	0	0	0
MEA+							0	0	0	0	0	0
MEACOO-							0	0	0	0	0	0
HCO3-	829.737	2134.130			0.224		825.609067	1087.54364	1089.48321	4831.94315	4274.83981	0
PZ	5942.476	603.665	1.75E-09		5.04E-05	0.004	5938.04901	6335.14544	6339.34544	12409.229	10745.6663	2.58874485
PZCOO-2	14156.096	17307.199			5.93E-12		14138.1238	12586.6107	12571.4181	1969.35152	3212.72021	0
PZCOO-	22516.611	5758.530			3.16E-08		22542.9857	23112.7645	23117.8765	22686.4591	23735.9483	0
PZH+	30711.929	24456.732			0.320		30705.6352	29893.3552	29884.1927	24203.6619	25375.1008	0
PZH+2							0	0	0	0	0	0
HPZCOO	19302.270	51216.724	4.08E-14		0.00013067		19300.9783	20512.8896	20526.5619	28289.4423	27055.7552	2.87E-07
N2		1.253	18088.688	18091.121	1.180		0.00191314	0.00191314	0.00191314	0.00191314	0.00191314	1.25108013
O2		0.079	609.819	609.974	0.075		0.00017788	0.00017788	0.00017788	0.00017788	0.00017788	0.07920733

Stream Name	16	4	13	6	5	7	11	14	9	8	10
ASPEN Stream Name	**PRODUCT2**	**RICH**	**RICHBPS2**	**RICHCBPS**	**RICHCLD2**	**RICHCOLD**	**RICHHOT**	**RICHTTL**	**RICHWBPS**	**RICHWRM1**	**RICHWRM2**
Temperature C	70.062825	42.3700411	109.273402	42.5321019	42.5309023	42.5321019	141.781037	116.339909	117.963555	117.963555	117.963555
Pressure kPa	914.04568	101.325	914.045675	914.045675	914.045675	914.045675	914.045675	914.045675	914.045675	914.045675	914.045675
Vapor Frac	0.8304737	0	0	0	0	0	0.01834921	0	0	0	0
Mole Flow kmol/min	217.25761	10015.5382	300.615935	300.466205	10015.5402	9715.07399	8514.05698	1603.55701	1302.94913	9722.00504	8419.0559
Mass Flow kg/min	8435.9608	266402.281	7992.06851	7992.06851	266402.282	258410.215	223777.932	42624.368	34632.2968	258410.215	223777.918
Volume Flow cum/min	546.12667	240.802284	7.49047509	7.2225228	240.750649	233.528238	786.49689	40.1484171	32.6588372	243.68517	211.026332
Mass Flow kg/min											
H2O	780.78421	164807.665	4898.29772	4944.13631	164804.617	159860.407	137578.092	26106.6294	21208.7447	158249.864	137041.12
CO2	7649.4592	13.7511528	7.00471913	0.41517475	13.8377343	13.4239837	4456.72615	49.3301991	42.6796825	318.456093	275.77641
MEA	0	0	0	0	0	0	0	0	0	0	0
MEA+	0	0	0	0	0	0	0	0	0	0	0
MEACOO-	0	0	0	0	0	0	0	0	0	0	0
HCO3-	1.6165409	2479.432	229.955281	74.7001772	2489.75602	2415.30572	4996.71245	1286.12423	1054.77013	7870.20786	6815.43773
PZ	0.0026241	617.226728	74.2465994	18.6249149	620.865646	602.205582	7017.5893	435.841284	361.810208	2699.66079	2337.85058
PZCOO-2	3.11E-05	17134.8401	185.635922	513.020511	17100.7883	16587.6632	2429.71471	874.805411	690.958025	5155.60989	4464.65187
PZCOO-	0.000645	5865.71365	358.132124	176.585786	5886.05745	5709.60709	17201.9558	1971.86952	1613.11187	12036.2963	10423.1844
PZH+	2.3091449	24847.6984	758.032908	745.261104	24841.6945	24096.7757	21204.1016	4053.13234	3294.49042	24581.967	21287.4766
PZH+2	0	0	0	0	0	0	0	0	0	0	0
HPZCOO	0.4581511	50634.6236	1480.7237	1519.28456	50643.3341	49123.5342	28891.9184	7846.42235	6365.56074	47496.8762	41131.3155
N2	1.2510801	1.25299275	0.03758978	0.03758978	1.25299275	1.21540297	1.05251391	0.20047884	0.16288906	1.21540297	1.05251391
O2	0.0792073	0.07938517	0.00238155	0.00238155	0.07938517	0.07700362	0.06668355	0.01270162	0.01032007	0.07700362	0.06668355

The base-case stripper contains both a CRB and WRB (Figure 1). A fraction of the cold rich solvent exiting the bottom of the absorber is heated by the product gas in a cross exchanger with a 20^o C LMTD before being fed into the top of the stripper. The remaining rich solvent is heated to its bubble point by a warm solution from the bottom of the stripper in a cross exchanger. Another portion of the warm rich solvent is bypassed, mixed with the CRB stream, and fed directly into the top of the stripper. The remaining rich solution is heated first by the hot lean liquid in a cross exchanger and finally by a steam heater to 150^o C and flashed into the bottom of the column. The two liquid-liquid cross exchangers are designed to have a combined 5^o C LMTD, as defined by Equation 11. Q is the heat duty of an exchanger, and the subscripts 1 and 2 refer to the two heat exchangers in series.

$$LMTD_{TOT} = \frac{Q_1 + Q_2}{\frac{Q_1}{LMTD_1} + \frac{Q_2}{LMTD_2}} \qquad \text{Equation 11}$$

There are three tradeoffs: (1) the amounts of solvent removed in the CRB and WRB; (2) the total height of packing in the stripper; and (3) the lean loading. Increasing bypass decreases steam losses in the product stream but decreases the amount of heat recovered in the cross exchangers. Increasing stripper height increases the amount of CO_2 removed in the packing but increases the cost of the column.

Intercooled Absorber Optimization

The lean loading optimization is based on energy performance and, therefore, is more closely associated with the advanced flash stripper. The liquid flow rate and packing area are optimized simultaneously. Assuming 90% removal, the liquid flow rate is a function of the packing area. As the packing area increases the liquid flow rate decreases until it reaches a minimum. As the packing area decreases the column approaches an isothermal condition with an infinite liquid flow rate. Between these extremes exists a case that balances the capital cost of packing area and the operating cost of circulating solvent. Ultimately a techno-economic analysis is needed to determine this point, but experience suggests that the optimum liquid flow rate is between 1.05 and 1.3 times the minimum liquid flow rate. As a first-order approximation this study always uses a flow rate equal to 1.2 times the minimum. With the liquid flow rate set, the packing area is minimized by adjusting the location of the intercooling. For an absorber with in-and-out intercooling this is relatively straightforward. An absorber with pump-around intercooling has three sections, and the middle section has less packing area per unit volume. This optimization is performed using the Aspen Plus® optimization tool. An optimum pump-around rate for coal-fired applications was approximated by Sachde to be five times the inlet vapor flow rate.

Advanced Flash Stripper Design and Optimization

The advanced flash stripper is designed to reduce the equivalent work by reducing both steam losses and sensible heat requirement. Equation 12 calculates the equivalent work, W_{EQ} (kJ/mol CO_2) as a function of reboiler duty, Q_i, reboiler temperature, T_{reb}, pump work, W_{pump}, and compressor work, W_{comp}. The sink temperature, T_{sink}, is assumed to be 40 0C. Equation 13 and Equation 14 calculate W_{comp} as a function of inlet pressure, P_{in}.

$$W_{eq} = \sum_{i=1}^{n_{reboilers}} 0.75 Q_i \left(\frac{T_i + 5K - T_{\sin k}}{T_i + 5K} \right) + W_{pumps} + W_{comps} \qquad \text{Equation 12}$$

$$W_{comps}\left(kJ/mol\ CO_2\right) = 4.572 \log\left(\frac{148}{P_{in}(atm)} \right) - 4.096 \qquad P_{in} \le 4.5 atm \qquad \text{Equation 13}$$

$$W_{comps}\left(kJ/mol\ CO_2\right) = 4.023 \log\left(\frac{148}{P_{in}(atm)} \right) - 2.181 \qquad P_{in} > 4.5 atm \qquad \text{Equation 14}$$

Contacting cold rich liquid with the hot product gas will reduce both the vapor pressure of water in the product and the hot side approach on the main cross exchanger. The amount of cold and warm liquid that is bypassed determines the extent to which these values are reduced. There is, of course, a limit to how much liquid can be bypassed usefully. Higher lean loading cases will not have enough steam to strip the CO_2 from the colder liquid entering the top of the column. If too much liquid is bypassed the desired lean loading will not be achievable. Lower lean loading cases will have higher concentrations of steam, but there must be enough liquid exiting the main cross exchanger to avoid a temperature pinch on the hot side of the exchanger. The equivalent work is minimized by adjusting the relative flow rates in the CRB and WRB without violating these physical constraints.

Calculating Cost of CO_2 Avoided

In order to compare the effects of process conditions on CAPEX and OPEX, both expenses must be expressed in dollars per metric ton of CO_2 captured. The PEC can be generally converted to these units using Equation 15.

$$\$/MT\ CO_2 = \frac{\alpha \times \beta \times PEC}{Total\ MT\ captured\ per\ year} \qquad \text{Equation 15}$$

In Equation 15 α converts the PEC to a total capital requirement (TCR) and β annualizes the cost. Literature values for α range from as low as 2 to as high as 10, depending on the process unit in question. The 2010 DOE Cost and Performance Baseline results in a value of 2.8. The annualizing factor, β, takes into account return on investment (10%), taxes (35% of return on investment), depreciation (3–10%, depending on plant lifetime), and maintenance (2–3%). Typical values of β range from 0.1 to 0.3.

RESULTS AND DISCUSSION

Optimum Design Configuration

The cold rich bypass (CRB) draws off 4.5 mol % of the total liquid leaving the absorber. The warm rich bypass (WRB) draws off 11 mol % of the remaining liquid or 10.5 mol % of the liquid leaving the absorber. In all evaluated cases the optimum design resulted in a larger WRB than CRB. This is primarily due to the physical limits of the liquid-vapor heat exchanger (X4). Because liquids exhibit much greater heat capacities, the amount of liquid that can be bypassed without pinching on the hot end of the cross exchanger is relatively small. The amount of vapor generated is nearly constant across all cases, so the maximum amount of liquid in the CRB is also nearly constant.

The two main cross exchangers, X2 and X3, have a combined LMTD of 5 K. The LMTDs of X2 and X3 are 4.6 K and 6.8 K, respectively. As the lean loading increases, the LMTD of X2 decreases and the LMTD of X3 increases. This is a consequence of the CRB and WRB. Because it has lost CO_2 and some H_2O, the mass flow rate on the lean side of the exchanger is less than the mass flow rate on the rich side in the absence of bypasses. This mass imbalance will cause the temperature approach on the hot side to be larger than on the cold side. Bypassing rich solvent will reduce this imbalance, the hot side temperature approach, and the steam that must be supplied to the steam heater to account for the sensible heat. Because the opportunity for steam recovery by bypassing rich solvent decreases as lean loading increases, the LMTD of X3 increases.

The heat duties of X2 and X3 are 830 MW and 280 MW, respectively. This ratio increases as the lean loading increases. Because X2 heats the rich solvent to its bubble point, the heat duty of X2 is determined primarily by the regeneration pressure. As the lean loading increases from 0.2 to 0.35 moles of CO_2 per mole of alkalinity, the rich loading only increases from 0.38 to 0.41 moles of CO_2 per mole of alkalinity. Therefore, the bubble point temperature of the rich solvent is determined primarily by the pressure. Because the regeneration pressure increases as the lean loading increases, so does the bubble point temperature. This increases the duty of X2 relative to that of X3.

Table 5: Equipment table for advanced flash stripper with intercooled absorber

Description	No. Trains	Type	Cost Source
Inlet Gas Blower	1	Centrifugal blower; SS or alloy process-wetter components	Verbal quote from vendor for blower; PDQ$ for motor
Absorber	1	Packed Tower (316SS Mellapak 250X/125X); Section heights = 4.25, 4.60, and 0.64 m; 316 SS Shell and Distributors	Vendor quotes for individual components
Absorber Intercooler	1	Plate and Frame; 316 SS; 5 psi pressure drop	Vendor Quote
Absorber Intercooler Pump	1	Centrifugal ; 316 SS	PDQ$
Rich Amine Pump	1	Centrifugal ; 316 SS	PDQ$
Rich Amine Carbon Filter	1	316 SS with Teflon Gasket	PDQ$
Particulate Filter	1	316 SS with Teflon Gasket	PDQ$
Amine Cross Exchangers	1	Plate and Frame; 316 SS	Vendor Quote
Lean Solvent Cooler	1	Plate and Frame; 316 SS	Vendor Quote

Table 6: Equipment table for advanced flash stripper with intercooled absorber

Description	No. Trains	Type	Cost Source
Stripper	1	HP Flash Vessel and 5 m packed column (316 SS Mellapack 250X)	Vendor quotes for i tower and PDQ$ for HP flash vessel
Convective Steam Heater	1	Shell and tube; 316 SS tubes and carbon steel shell	PDQ$
Compressors	1	Centrifugal; multistage; 316 SS	Vendor Quote
Overhead Condenser	1	Plate and Frame; 316 SS	Vendor Quote
Overhead Accumulator	1	Horizontal vessel; 316 SS	PDQ$
Makeup Amine Tank	1	Fixed roof tank	PDQ$
Makeup Amine Pump	1	Centrifugal	PDQ$
Water Tank	1	Fixed roof tank	PDQ$
Water Pump	1	Centrifugal	PDQ$
Surge Tank	1	316 SS horizontal vessel	PDQ$
Lean PZ Pump	1	316 SS Centrifugal	PDQ$
Reclaimer	1	Similar reclamation system to Case 12	Scaled vendor quote
Dehydration Unit	1	TEG unit	Scaled vendor quote

Scaling Purchase Equipment Cost (PEC) to 2010 DOE Cost and Performance Baseline

A scaling exponent was calculated according to the method outlined in the 2012 DOE NETL Report. If it is assumed that increasing the size of the plant will increase the size of the process unit, a multiplier of 0.6 is used. If increasing the size of the plant will necessitate the purchase of additional units, a multiplier of 1.0 is used. The weighted prices of each process unit are added together and divided by the total cost of the process at 593 MW_e. The result is used as a scaling exponent for calculating the purchased equipment cost of the capture and compression units at given power plant electrical capacity, CAP, using Equation 16 and Equation 17, respectively.

$$PEC\ Capture = 66{,}881{,}000\left(\frac{CAP}{593MW}\right)^{0.77} \qquad \text{Equation 16}$$

$$PEC\ Compression = 12{,}198{,}000\left(\frac{CAP}{593MW}\right)^{0.62} \qquad \text{Equation 17}$$

The PEC of the PZ-AFS configuration is slightly more than that of the PZ-2SF for a 593 MW_e gross electrical generation. The greatest difference between the configurations is PEC for the cross exchangers and the convective steam heaters. The PZ-AFS attempts to reduce steam requirement by (1) recovering steam in the product stream by bypassing cold rich solvent, and (2) reducing the hot-side temperature approach and, thus, the portion of the steam heater duty associated with the sensible heat of the solvent. The reduction in steam heater PEC and increase in cross exchanger PEC are due to a redistribution of heat duties. The increase in absorber PEC is a result of the pump-around intercooling configuration. The absorber in the PZ-AFS requires approximately 35% less packing area to capture 90% of the CO_2 from the 593 MW_e case. However, the pump-around intercooling configuration requires an additional set of distributors and supports, as well as a larger heat exchanger and pump for the additional liquid load in the middle section of the column. Pump-around intercooling improves the solvent capacity and, thus, should also reduce the steam heater duty. This analysis suggests that there is no net effect on CAPEX associated with the configuration, but the reduction in OPEX improves the cost of CO_2 avoidance.

The PEC of the compressor train is scaled to inlet vapor flow rate and pressure. The average inlet pressure of the PZ-2SF configuration is greater than that of the PZ-SS or PZ-AFS configurations. The pressure of the PZ-AFS (7.8 bar) is slightly greater than that of the PZ-SS (7.4 bar), which accounts for the slight reduction in compressor train PEC.

Table 7: Prices of unit operations for 5 m PZ AFS

	Hybrid-Series		Hybrid-Parallel	
Description	**Case 13**	**Case 14**	**Case 18**	**Case 19**
Inlet Gas Blower	2,841,000	2,841,000	2,841,000	2,841,000
Absorber	16,194,446	25,826,915	13,200,010	15,228,055
Absorber Intercooler	2,479,825	2,610,342	2,218,790	2,871,376
Absorber Intercooler Pump	1,992,787	5,181,246	1,651,166	2,334,408
Rich Amine Pump	869,559	2,742,454	668,891	1,070,226
Rich Amine Carbon Filter	181,793	277,083	170,773	181,583
Particulate Filter	136,000	136,000	136,000	136,000
Rich/Lean Amine Exchanger	22,971,006	73,611,633	16,445,152	28,191,689
Lean Solvent Cooler	1,174,654	3,393,444	913,620	1,435,688

CAPEX Summary

The hybrid PZ process is categorically less expensive than the MEA-Econamine process in Case 12. The prices in Table 3.4 are reflective of relative differences in both CAPEX and OPEX. Because the plants have been derated to 550 MW_e net power production, the thermal efficiency of the CO_2 capture plant determines the gross power plant capacity. The contribution of CAPEX and OPEX to the total plant PEC requires a closer analysis. For example the PZ-SS and PZ-2SF cases only differ by 0.6% in required power plant capacity, but the PZ-2SF PEC is 8.3% less than that of the PZ-SS. The decrease in PEC between the cases is almost entirely due to the decrease in CAPEX from using two flash vessels and steam heaters rather than a stripper and reboiler. Table 3.2 suggests that the PEC of the advanced flash stripper is nearly identical to that of the two-stage flash. The decrease in CAPEX is due to the improved energy performance.

Another difference between the cases worth noting is the actual percentage of CO_2 that is being captured compared to a 550 MW_e plant without CO_2 capture and compression equipment. The CO_2 scrubbing processes are designed to capture 90% of the total CO_2 in the plant flue gas. If the capacity of the plant is being increased to guarantee 550 MW_e net power production, the CO_2 scrubber has to be scaled to accommodate the added capacity. Because 90% of the total CO_2 in the plant flue gas is being capture, 10% is being emitted. Increasing the capacity of the plant will increase the magnitude of that 10% and, thus, decrease the percent CO_2 avoided. Table 3.8 summarizes the CO_2 avoided across the four cases in this study.

CONCLUSIONS AND FUTURE WORK

- The advanced flash stripper with intercooled absorber represents an improvement in both capital and operating costs over the short stripper and two stage flash configurations reported in the 2012 DOE NETL Report for a supercritical pulverized coal power plant with 550 MW_e net capacity.

- The added power plant capacity required to avoid 90 % of the CO_2 and maintain 550 MW_e net capacity was reduced to 141.9 MW_e, and the purchased equipment cost was reduced to $167.5 MM.
- The main contributors to the capital cost of CO_2 capture and compression are the absorber, cross exchangers, reboiler, and compressor.
- The capital cost of the cross exchangers is highly dependent upon the calculation of heat transfer coefficients.
- When evaluating the impact of CO_2 avoidance on the cost of electricity more attention should be paid to the impact of assumptions relating purchased equipment cost to total capital requirement.

REFERENCES

Department of Energy (DOE) National Energy Technology Laboratory (NETL). "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity", Revision 2, November 2010, DOE/NETL 2010/1397.

Fisher KS. "Advanced Amine Solvent Formulations and Process Integration for Near-Term CO_2 Capture Success", June 2007, Grant No: DE-FG02-06ER84625.

Frailie, P. Excerpts from PhD dissertation, "Modeling Absorber/Stripper Performance with MDEA/PZ". University of Texas at Austin. April 2014.

Hewitt GF, Shires GL, Bott TR. "Process Heat Transfer." 1994: CRC Press, Inc.

Peters MS, Timmerhaus KD, West RE. "Plant Design and Economics for Chemical Engineers, 5th Edition." 2003: McGraw Hill.

Sexton AJ. "Techno-Economic Analysis for CO_2 Capture by Concentrated Piperazine with Regeneration by High Temperature Two Stage Flash: Budget Period 1", June 2012, Cooperative Agreement No: DE-FE0005654.

Van Wagener DH. *Stripper Modeling for CO_2 Removal Using Monoethanolamine and Piperazine Solvents.* The University of Texas at Austin. Ph.D. Dissertation. 2011.