International Journal of Greenhouse Gas Control 96 (2020) 103014

Contents lists available at ScienceDirect

International Journal of Greenhouse Gas Control

journal homepage: www.elsevier.com/locate/ijggc

Experimental results of advanced technological modifications for a CO_2 capture process using amine scrubbing

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ARTICLE INFO

Keywords:
Split flow
CO_2 capture
Pilot plant
Interheating (do 5 max)

ABSTRACT

This paper provides a discussion of the experimental results obtained at the amine-based carbon capture pilot plant having the capacity of 200 m_n^3/h of flue gas, using 30 wt% ethanolamine solution as a solvent. The objective was to prove the superiority of application of advanced amine flow systems as well as of the novel stripper internal heater fulfilling pilot Technology Readiness Level. Standard process flow sheet, multi absorber feed and split-flow process modifications with and without stripper interheating were examined. A vast number of process parameters were recorded during the trials. The data were critically analysed, compared and presented in the paper. Demonstrated process flow modifications resulted in a reduction of the reboiler heat duty by about 5%, while using internal stripper interheatingby 9 ÷ 11 %, comparing to standard amine-based carbon capture plant.

1. Introduction

Carbon dioxide, as the most significant anthropogenic greenhouse gas, substantially contributes to global warming and climate change. Raising the level of anthropogenic CO_2 emissions in the last 150 years is especially dictated by growing energy consumption, which is compensated by fossil fuel combustion. In order to counteract the rising CO_2 level, low emissions, high-efficiency technologies should be implemented. These "clean coal technologies" should allow combusting fossil fuels without negative environmental impact (including carbon dioxide emissions reduction to atmosphere). One of the solutions for the reduction of carbon dioxide emissions is a post-combustion capture. There are many industrial technologies used to remove carbon dioxide from flue gases: physical and chemical absorption, adsorption, membrane processes, cryogenic and electrochemical processes (Rackley, 2009). The most mature technology, which can be implemented in existing large scale power plants, is chemical absorption. In the industrial scale, the chemical absorption CO_2 removal processes mainly use aqueous ethanolamine (MEA) solvents. The biggest advantages of MEA are low price, fast kinetics of CO_2 absorption and high mass transfer rates. However, the usage of MEA on the scale required in the power industry causes certain consequences. The power plant efficiency drops significantly, due to the high energy consumption of CO_2 removal process, resulting from the high regeneration energy requirement. Furthermore, MEA suffers from both thermal and oxidative degradation which makes it necessary to make up the system with fresh solvent or to reclaim the solution. Therefore, the decrease of regeneration energy is sought. Achievement of this goal is possible through replacement of MEA with other amines, having lower regeneration requirements, such as 2-amino-2-methyl-1-propanol (AMP), piperazine (PZ), 2-(2-aminoethyl-amino)ethanol (AEEA), 1,3-Diamino-2-propanol (AEP) (Wilk et al., 2017; Bernhardsen and Knuutila, 2017) etc. The other method is to modify the process flow sheet (Chakma, 1997; Cousins et al., 2011; Le Moullec et al., 2014; Spietz et al., 2014). Institute for Chemical Processing of Coal, in cooperation with two industrial partners: TAURON Polska Energia S.A. and TAURON Wytwarzanie S.A., designed, erected and operated carbon capture pilot plant for a searching solution in this field.

The purpose of the pilot plant research was to gain hands-on experience with carbon dioxide capture process and technology on real, hard-coal originated flue gases and to give opportunities to test novel solvent which was also developed in a part of the project.

The aim of this study was to prove the superiority of application of advanced amine flow systems, as well as of novel stripper internal heaters. Further goal included gaining experience in the usage of the 30 wt% MEA benchmark solvent used as a baseline for comparisons with further studied solvents.

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https://doi.org/10.1016/j.ijggc.2020.103014
Received 15 November 2019; Received in revised form 21 February 2020; Accepted 5 March 2020
Available online 19 March 2020
1750-5836/

Nomenclature

IH	interheating
MAF	multi absorber feed process configuration
HE	heat exchanger
S	standard process configuration
SF	split flow process configuration
PDU	process development unit

Subscripts

n	standard conditions (101.3 kPa, 0°C)

2. Experimental

2.1. Plant description

Design of the pilot plant is based on chemical post-combustion absorption process with several modifications. The designed pilot plant (Table 1) is a mobile unit, which allows its usage on various locations (Fig. 1).

The pilot plant consists of three containers: supervision, storage and technological. In the technological container, process equipment is located. During transport, oversized parts of the installation (columns), are placed in the storage container. All containers have typical dimensions, allowing the fast shipment and installation on site. The mobility of the pilot plant allows conducting tests on different sources of flue gases. The pilot plant is equipped with a deep desulfurization module, therefore desulfurization processes can also be tested. Desulfurization module is a necessity in locations which are not equipped with SO_2 cleaning systems and having sulphated flue gas. The sulphur dioxide in the flue gas stream has to be removed, due to its impact on the amine scrubbing module (degradation of amine) Anon. (2019a). The pilot plant gives the opportunity to test carbon dioxide capture from flue gases, using chemical absorption in various amine solutions, as well as gives a possibility to change process configuration. The latter allows investigation of the influence of the process configurations on CO_2 removal rate and energy demand.

2.2. Process description

Fig. 2 shows the pilot plant process flow diagram. Dedusting, cooling and SO_2 removal is conducted in the deep desulphurization module, while carbon dioxide is removed in the amine scrubbing module.

The flue gas is dedusted and cooled in direct water Venturi scrubber. Dedusted gas from the Venturi scrubber is treated in countercurrent desulphurization column, where sulfur dioxide is removed. A solvent used for desulphurization contains sodium bicarbonate and reduces SO_2 concentration in flue gas below 20 mg/m^3. Next, the gas enters an absorber and flows upward through its packing, where the CO_2 reacts chemically with semi-lean and lean amine solutions supplied from the stripper. The absorber is removing most of the CO_2 from flue gas.

The treated gas passes through a water wash absorber section for cooling and separating excess water, and finally evacuated through the top of the absorber.

The rich amine solution is pumped from the absorber's bottom into the stripper using three pipelines:

- through rich-semi-lean 1 heat exchanger (about 45 % of rich amine solution stream).
- through rich-lean and rich-semi-lean 2 heat exchangers (about 45 % of rich amine solution stream).
- directly to the top of the stripper without heating (remaining rich amine solution stream).

Solvent heat exchangers allow transferring heat from hot semi-lean and lean solutions to colder rich solution. The latter stream (without heating) has the same function as a rich-split process described in the literature (Stec et al., 2015a, b; Artanto et al., 2012).

The semi-lean and lean solutions are pumped back from a bottom of the stripper to the top of the absorber. The solvent streams are cooled and filtered, prior to the absorber.

A reboiler of the stripper has a built-in electrical heating element to evaporate water and to strip carbon dioxide from the solvent. Hot vapours are used to regenerate and separate more carbon dioxide from the rich amine in upper sections of the regenerator. Hot vapours (containing mainly steam and CO_2) leaving the stripper, enter the internal condensation section at the top of the column where most of the water is removed. Remaining part of the water in the outlet gas stream is removed in an additional external condenser. The gas leaving the condenser contains almost pure CO_2. The condensate is directed to the stripper as reflux. Column sizes, packing heights, and packing elements of the mobile pilot plant columns can be found in our previous articles (Stec et al., 2015a, b).

2.2.1. Advanced flow systems

The plant construction provides a possibility of solvent flow modifications (Figs. 2 and 3). The simplest, standard configuration (Fig. 3A) enables comparison with other literature reports describing other pilot plants (CSIRO PCC (Bernhardsen and Knuutila, 2017) or CASTOR (Mangalapally et al., 2012a)). In the standard configuration with both lean and rich amine streams are merged and pumped into the absorber through a single inlet. Contact between the streams associated with the heat exchange occurs only in one rich-lean heat exchanger. However, presented pilot plant (Fig. 2) possess three heat exchangers designed and applied to allow easy comparison with more advanced solvent flow configurations. Thus, during the study standard configuration works with using all of the heat exchanger installed (Fig. 3A).

Fig. 3B shows the multi absorber feed flow configuration. After passing three heat exchangers and coolers (not shown) the lean amine is fed into two sections of the absorber: one at the top and second in the middle of the absorber. Injection of colder, compared to the standard configuration, lean solvent in the middle of the absorber causes a temperature decrease and thermodynamically favours the CO_2 absorption in the solution (Sanchez-Fernandez et al., 2014; Feron, 2016). Multi absorber feed enables a somewhat higher equilibrium loading. The multiple absorber feed configuration has been advantageous for low inlet pressure processes like carbon capture from flue gases (Spietz et al., 2014; Anon., 2019a).

Fig. 3C shows the most advanced split stream process configuration. The concept of the flow splitting was firstly suggested by Shoeld (1934). His patent aiming to remove H_2S from fuel gases using sodium phenolate. Shoeld suggested splitting the streams of both lean and rich amine. As he claimed that modification reduces steam usage by 50 % as compared with the conventional standard flow process. Shoeld's idea has been improved by several authors (Anon., 2019b; Condorelli et al.,

Table 1
Pilot plant design specification.

Parameter	Design parameter, unit
Flue gas capacity	200, m_n^3/h
Absorption efficiency	90 %
Nominal solvent flow	1570, dm^3/h
Lean amine temperature	40°C
Semi-lean amine temperature	40°C
Absorber top pressure	30, $kPa_{(g)}$
Stripper top pressure	30, $kPa_{(g)}$
Max. Reboiler heating power	63 kW

Fig. 1. Carbon capture pilot plant during tests at the Łaziska Power Plant in Łaziska Górne, Poland.

2001; Freguia et al., 2004). This configuration was previously tested in our Institute on the PDU scale (Stec et al., 2017), (Krótki et al., 2017).

In this configuration, an additional stream of semi-lean amine is included. The portion of amine is taken from the stripper liquid collector, cooled and pumped to the middle of the absorber (Polasek et al., 1983). The semi-lean amine solution from liquid collector entering the middle of the absorber absorbs a part of the CO_2 from the flue gas stream. However, the lean amine, which has very low CO_2 loading is injected into the top of the absorber section where remaining CO_2 is absorbed. Pumping lean amine to the absorber top section where the partial pressure of CO_2 is the lowest and semi-lean amine where the partial pressure of CO_2 is higher, makes the thermodynamic driving force constant along the column. This result in a lower concentration of CO_2 the treated gas. However, the overall circulation rate of the solvent is higher than for standard flow rate. It is expected that split stream configuration should reduce in reboiler heat duty from 5 to 30 % (Anon., 2019a).

2.2.2. Stripper interheating

Next process modification, implemented in the pilot plant, which is aimed to decrease energy consumption of CO_2 removal process, are two heat exchangers built-in the stripper for interheating (Cousins et al., 2011; Spietz et al., 2014). This modification, called generally a stripper inter-heating, primarily has been proposed by (Feron, 2016; Leites et al. (2003); Oyenekan and Rochelle, 2006). Diagram of a modified stripper inter-heated column for the split flow process is shown in Fig. 2, while for standard and multi absorber feed in Fig. 4A and B. Presented inter-heaters were patented in 2018 (Budner et al., 2018).

A rationale explaining the split stream process (Fig. 2) is as follows: a hot solvent of the lean amine from the stripper reboiler instead of the normal way to the rich-lean heat exchanger is redirected back to the stripper column section where two special heat exchangers are built. This two stripper internal heat exchangers – lower, separated with a liquid collector to collect semi-lean amine solution and upper internal heat exchanger. The internal heat exchangers are separated by a liquid collector where thermocouple and pressure transmitters are located. The internal heat exchangers are a shell and tube type heat exchangers with a packing placed inside the tubes. The tubes have a double wall (a pipe in a pipe) welded to two separate sieve bottoms. The heating amine flowing through the tubes is heating carbon dioxide and water vapours mixture in a countercurrent way (Knudsen et al., 2009a). The lean amine flows through lower internal heater while the semi-lean amine through an upper internal heater. After leaving stripper lean and semi-lean amine solutions are directed further to the heat exchanger island and to the absorber.

Used modification results in indirect heat transfer from hot solution to the stripper column instead of being exchanged in heat exchanger island outside of the column. Thanks to the direct exchange of heat in the stripper, ambient heat losses are reduced compared to the heat exchange accomplished in the external heat exchanger island. Furthermore, heat transferred inside of the stripper results in an increase of the column internal temperature and improves the solvent regeneration rate (Spietz et al., 2014). Therefore the reboiler heat duty is lowered (Cousins et al., 2011; Oyenekan and Rochelle, 2007).

2.3. Materials and media

This paper contains descriptions and results obtained during research campaigns carried out at the Łaziska Power Plant. The flue gases

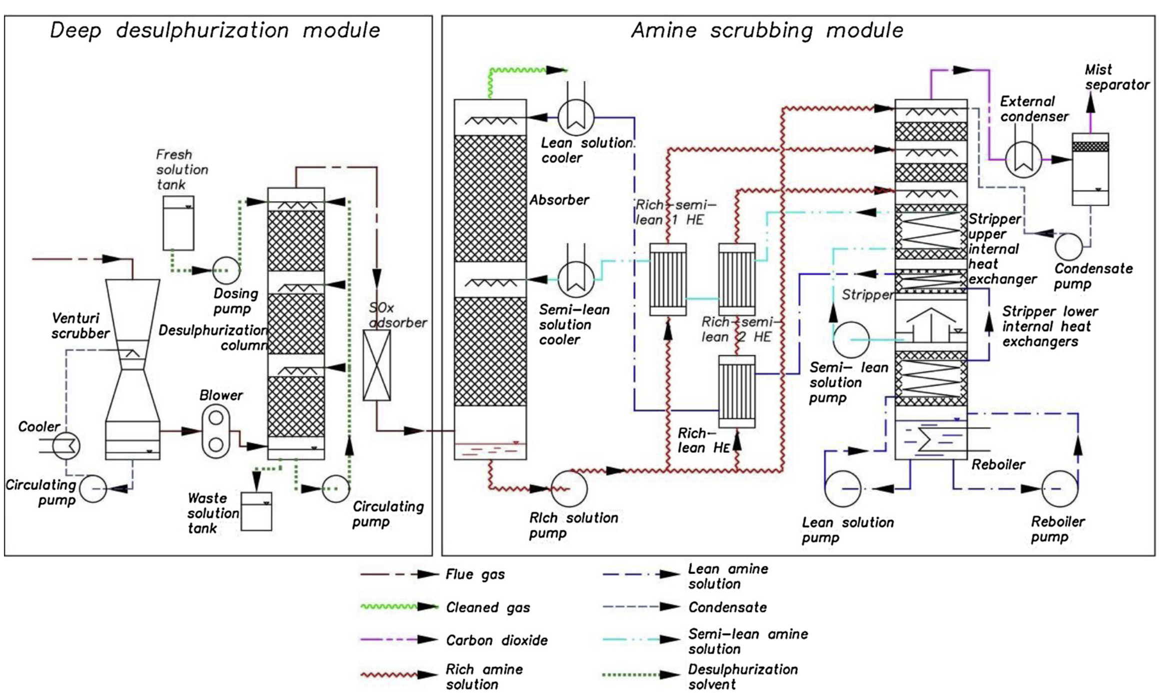

Fig. 2. Pilot plant flow sheet.

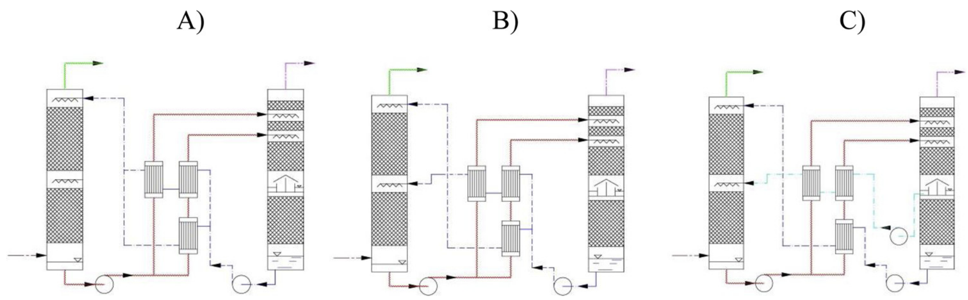

Fig. 3. Simplified process flow systems considered in this study: (A) standard system-S, (B) multi absorber feed- MAF and (C) split-flow process-SF.

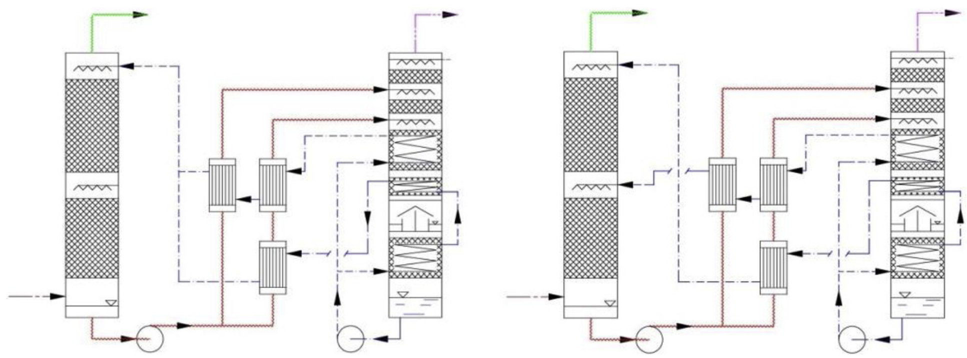

Fig. 4. Simplified process flow systems with interheating considered in this study: (A) standard system-S IH, (B) multi absorber feed-MAF IH.

were taken from hard coal pulverized-fuel bed boiler 225MW$_e$. Typical flue gas composition during study presents Table 2.

For the advanced technological study, a simple 30 wt% monoethanolamine (MEA) water solution with the addition of anti-foam Silpian W-3 was chosen. This choice was done because of the existing comprehensive database of physical and chemical properties, experience gained by the other pilot plant (Knudsen et al., 2009a; Lepaumier et al., 2011; Mangalapally et al., 2009, 2012b; Moser et al., 2011; Stec et al., 2015c). Concentrated monoethanolamine (MEA, CAS #000141-43-5) was obtained from Acros Organics. Silpian W-3 as an antifoaming additive was purchased from Silikony Polskie Sp. z o.o.

2.4. Measurements and analytics

The concentration of components in the gas streams was monitored on-line. Gas sampling probes were installed before and after desulfurization and absorption columns. Collected gas samples were transferred to two ULTRAMAT 23 analyzers (Siemens) where the concentration of CO_2, SO_2, O_2 and NO_x were measured. ULTRAMAT 23 analyzing O_2 with electrochemical oxygen measuring cell. CO_2, SO_2, NO_x were measured with infrared detection. Absorption column measurements (before and after column) are the most important measurements allowing CO_2 recovery calculation.

CO_2 loading of the solvent was determined by the densitometric method (Freguia et al., 2004; Krótki et al., 2017) using the Kyoto KEM DA-640 densitometer (density change with the content of CO_2 dissolved in solution). To verify MEA concentration in the solvent acid-base titration was carried out (Hartono et al., 2013). During the study, the stripper heater was regulated by a standard electric current Legrand 0046 84. Other measuring elements used in the pilot plant were written out in Table 3.

2.5. Testing methodology

Overall, over 500 h of tests with MEA use were carried out on the pilot plant in Łaziska Power Plant. The activities at the pilot plant have been divided into test campaigns, lasting approximately 100 h of continuous operation. Campaigns were divided into a shorter period called tests during which, the influence of a series of process parameters was investigated. During every test, key process parameters were registered with supervisory and data acquisition system (SCADA). Analysis of registered data allowed to select periods of a steady state lasting at least one hour (Fig. 5B). Steady state was determined when selected key parameters at the installation outlet:

- the absorber top outlet CO_2 concentration (light green line in Fig. 5) was constant (about ± 0.2 vol.%),
- the stripper CO_2 flow (blue line in Fig. 5) was constant (about ± 1.5 m^3/h),

and fluctuations of remaining parameters were negligible. For the chosen steady state the average of each parameter was used for further mass balance calculations. The results are presented in subsequent paragraphs. It is worth stressing, how important is to recognize steady state operation after changing process parameters.

The conducted tests, both for advanced flow systems and internal stripper heaters performance were carried out for the various liquid to gas ratios (L/G) - Table 4. L/G variation ratio was generated only by

Table 2
Average flue gas composition and parameters from the hard coal-fired boiler at Łaziska Power Plant.

Component	Value	Unit
CO_2	13.2	vol%
H_2O	7.3	vol%
O_2	8.7	vol%
Impurities		
SO_2 content	100-200	mg/m^3
NO_x content	200-400	mg/m^3
Dust	30	mg/m^3
Parameters		
Temperature	up to 99	°C
Pressure	1	bar$_{abs}$

Table 3
Measuring element used in the pilot plant.

Process parameter	Measuring element
Gas flow rate	Standard flanged orifices with Aplisens APR-2000 ALW differential pressure converters
Solvent flow rate	Endress + Hauser Promag 50P15 flowmeters
Pressure	Aplisens APC-2000ALW sensors
Temperature	Limatherm PT-100 sensors

liquid flow rate alteration, thus the gas stream was kept constant. For each L/G ratio, the power of the electric heater of the reboiler was adjusted until the CO_2 recovery was close to 90 %. Conducting tests in this manner allowed to find the optimum L/G ratio at which the minimum reboiler heat duty was achieved (Mangalapally and Hasse, 2011; Knudsen et al., 2009b). The reboiler heat duty reported is expressed as a gross value (value contains actual heat duty and the heat losses). During the tests the unheated rich solvent (rich-split) stream flow rate was held constant being about 10 % of the total rich solvent stream. For advanced flow system study, the internal stripper heaters were always used. For stripper interheating study the optimum L/G ratio test was taken into considerations.

The lines at presented figures were used to connect experimental data, and these lines serve only to join the data points.

2.6. Error determination

After mass and energy balance calculations an estimation of the maximum absolute error was performed (Table 5). Repeatability of the results was found to be within the range of the measurement errors.

3. Results and discussion

3.1. Advanced flow systems

Fig. 6 shows the reboiler heat duty as a function of the ratio of the solvent and the flue gas mass flow (L/G ratio) for conducted experiments at 90 % CO_2 recovery. Three advanced flow systems are compared: standard, multi absorber feed and split flow.

The shape of the obtained test curves is similar for all absorbent flow systems. Compared to the others, the standard system flow (orange curve) was more favourable for a lower liquid to gas ratios (L/G $<$ 4). However, above L/G $>$ 4 MAF and SF were superior. Each flow system, in a wide range of L/G, showed minimum reboiler heat duty. This reboiler heat duty, for comparable CO_2 recovery values, can be recognized as optimal conditions to conduct the process (Knudsen et al., 2009b). Summary of the optimal process results, for various flow systems, is presented in Table 5.

For the standard flow system, the lowest reboiler heat duty was obtained for L/G about 3.41 kg/kg. Compared to standard, when using

Table 4
Pilot plant average test parameters.

Parameter	Value	Unit
Flue gas flow rate	285 ± 7	kg/h
Absorption temperature	40 ± 1	°C
Absorber top pressure	1.3 ± 0.03	bar_a
Stripper top pressure	1.3 ± 0.03	bar_a
Rich solvent flow rate to the stripper	various: (450–1700)	kg/h
Reboiler heater power	various: (50.4–63)	kW

Table 5
Maximum absolute error for selected parameters.

Parameter	Error	Unit
Instrumentation error		
Flue gas flow rate	5	kg/h
CO_2 in gas composition	0.2	vol%
Temperature	1	°C
Pressure	0.003	bar_a
Solvent flow rate	4	kg/h
Proportional error		
CO_2 recovery	2	%
Reboiler heat duty	0.088	MJ/kg_{co2}
CO_2 loading	0.005	mol_{co2}/mol_{amine}
Heating/ cooling power	0.22	kW

a larger liquid flow rate, the MAF and SF systems allow achieving lower regeneration energy values. For MAF and SF the reboiler heat duties were nearly equal, taking into account its maximal absolute error. However, the SF flow system allowed to achieve low reboiler heat duties in a wider range of L/G ratio values, making the configuration more flexible to operate.

Fig. 6 shows also test points, where 88 % CO_2 recovery wasn't obtained. This low CO_2 recovery points obtained was caused by reaching of maximum electric power to the stripper reboiler heater (63 kW), thus these test needed to heat up more the stripper reboiler.

At constant CO_2 recovery, the reboiler heat duty closely depends on solvent flow. Too small absorbent portion inflow into absorber can't remove the required amount of carbon dioxide from the flue gas as it is fully saturated with CO_2. Increasing the flow rate of solvent causes absorption of a new portion of CO_2 and also decreases the rich solvent loading. The 90 % CO_2 recovery can be achieved by reboiler heating power lowering, hence the reboiler heater power decrease. Further flow rate increase results in a stripper temperature decrease, thus the lean loading further increases leading to the CO_2 recovery decrease. The reboiler heater power must be improved. This results in reboiler heat duty increase

The described dependencies can be also visualized on the graph (Fig. 7) for the standard flow system. The rich solution decreases CO_2

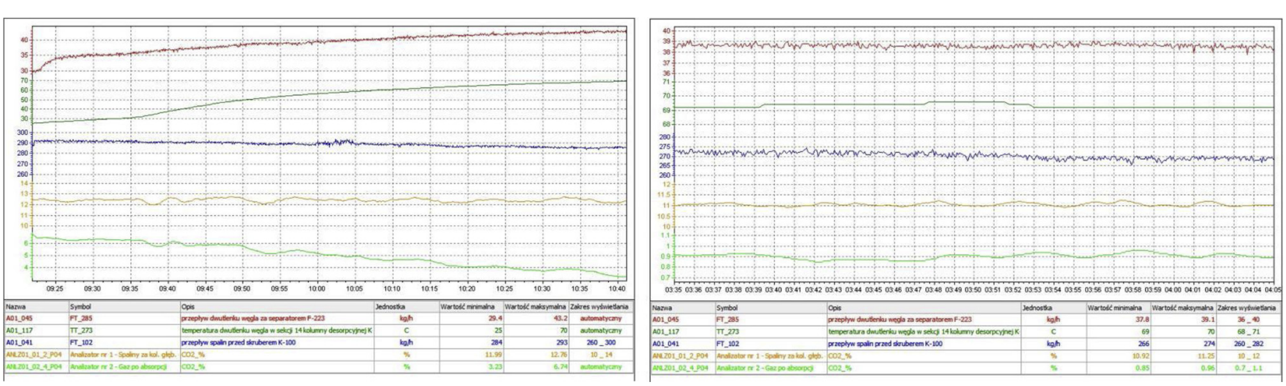

Fig. 5. SCADA key parameters screenshots from A) non-steady state, B) steady state.

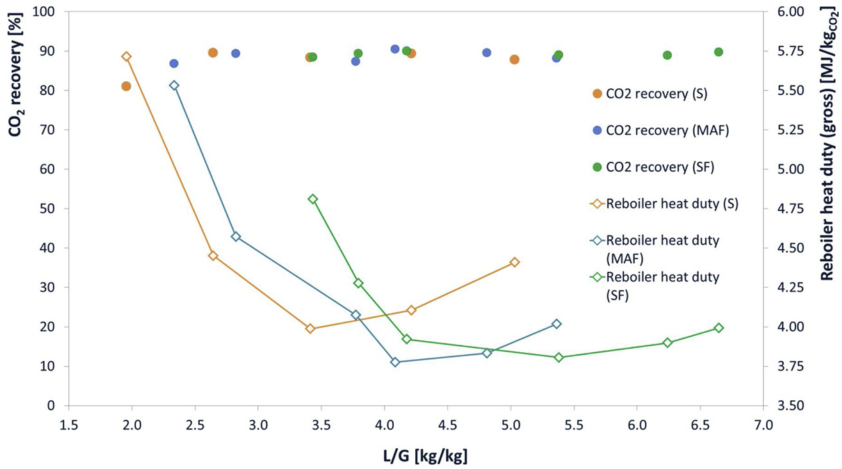

Fig. 6. The effect of L/G ratio on CO_2 recovery and reboiler heat duty for different process flow systems; S- Standard system, MAF- multi absorber feed, SF- split-flow process.

loading, except stripper temperature decrease, can be explained by the deterioration of both columns hydraulics in the higher parts of the columns as the performance of a packed column is very dependent on the maintenance of liquid and gas distribution throughout the packed bed. Increasing the solvent volume inflow to columns distributors may interfere in poor liquid distribution throughout the packed bed. Also, the higher liquid holdup on the packed bed disrupts drop formation what influence on CO_2 absorption in solution (Moser et al., 2011).

For MAF (Fig. 8), comparing to the standard flow system, there is no such radical decrease in rich solution CO_2 loadings level observed, pointing that a higher amount of CO_2 is being absorbed. Furthermore, above L/G = 3.5, for the similar L/G ratios the differences between CO_2 loadings were even slightly higher, what in fact results in lower reboiler heat duties. For the MAF flow system, the inflow of colder, lean solvent into the middle of the column causes column temperature decrease, what thermodynamically favours the CO_2 absorption similarly as can be observed for absorber intercooling (Shoeld, 1934; Krótki et al., 2017; Leites et al., 2003; Śpiewak et al., 2015).

For SF (Fig. 9), taking into account the maximal proportional error, it can be said that the rich solution CO_2 loadings were almost constant. The lean solution CO_2 loadings were the lowest from all studied flow systems observed. The semi-lean CO_2 loading values were rather close to rich CO_2 loading values for higher values of L/G pointing poor operation of the higher section of the stripper at high solvent loads.

The proposed Shoeld idea of split-stream operation was that the semi-lean amine solution directed to the middle part of the absorber, contacts gas having high CO_2 concentration (high partial pressure of CO_2) (Shoeld, 1934). Most of the CO_2 is captured in the lower part of the absorber by semi-lean amine stream. The lean solution purifies gas from the remaining amount of acidic component in the upper part where its partial pressure is low. The usage of split flow equalizes the driving force of absorption through absorber height. In our case for a low L/G ratios up to 4.0, the condition has been met. However, for higher L/G ratios (above 4.0), the CO_2 absorption was quite different. The CO_2 absorption through absorber column height was low in the lower part and high in higher parts of the absorber. Within high solvent flow rates, the semi-lean amine does not regenerate to an expected level. The semi-lean solution values of 0.547-0.56 mol_{CO2}/mol_{MEA} at 40°C seems to be near full loading for 30 wt% MEA solution according to (Bernhardsen and Knuutila, 2017; Anon., 1993). This indicates, that installation working limit was affected by CO_2 desorption process.

Table 6 shows the summary of the optimal process results for various flow systems, for which the lowest reboiler heat duty was achieved. For the optimal conditions with a great simplification can be said that using MAF and SF comparing with S flow systems reboiler heat duty can decrease respectively about 5,3 and 4,6%.

3.2. Stripper interheating

In the tests described below, apart from the solvent flow

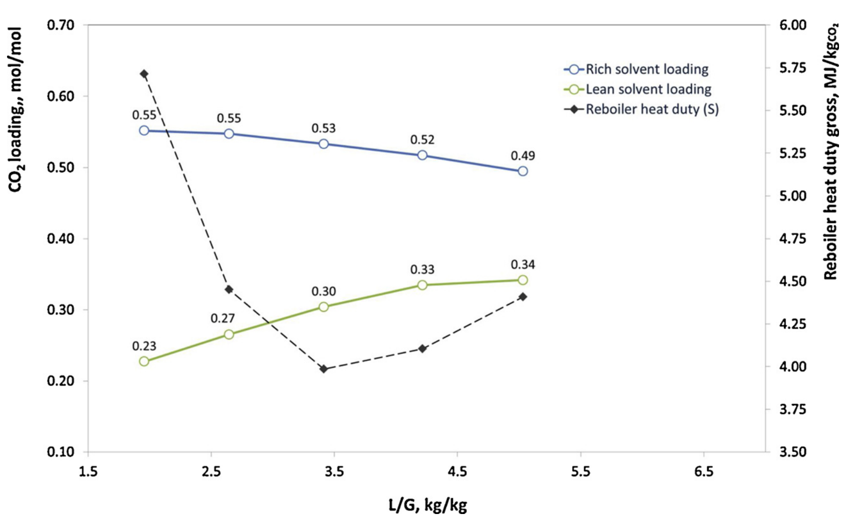

Fig. 7. The effect of the L/G ratio on CO_2 loading and reboiler heat duty for a standard solvent flow system at constant CO_2 removal efficiency.

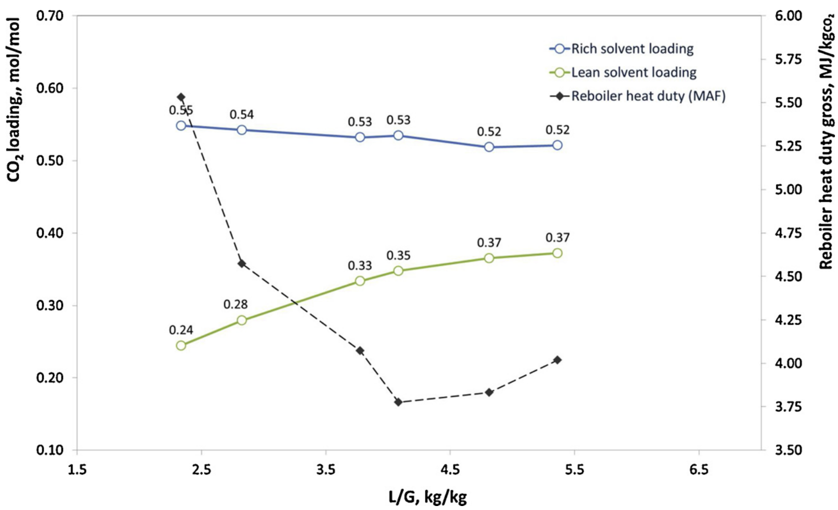

Fig. 8. The effect of the L/G ratio on CO_2 loading and reboiler heat duty for a multi absorber feed solvent flow system at constant CO_2 removal efficiency.

modifications, the internal heat exchangers (Leites et al., 2003) in the stripper were used. The tests allowed to show an influence of the internal heat exchangers on the process for advanced flow systems. According to the literature using a heat-integrated stripper reduces the energy penalty associated with regenerating amine solutions (Bernhardsen and Knuutila, 2017; Freguia et al., 2004; Knudsen et al., 2009b; Anon., 1993). To investigate interheating performance, a novel stripper invented and patented in 2018 by co-authors of this article (Stec et al., 2017), was tested.

Fig. 10 is an expansion of Fig. 6. The dotted lines indicate a change of reboiler heat duty for a different advanced flow system without internal heat exchangers. It can be noticed that for standard and split flow configuration with L/G ratio increase the reboiler heat duty differences increased. For MAF configuration up to L/G = 3.5, the reboiler heat duty was even higher when the heat exchangers were turned on. We also suppose that the same situation might occur when more test at low L/G ratio for standard and split flow configuration have been performed. Unfortunately, this test wasn't performed due to process installation restrictions. In spite of this, for the optimum L/G ratio, what was assumed, at the lowest reboiler heat duty and at least for 90 % CO_2 recovery, the reboiler heat duty was always lower having a stripper interheating on. Thus, it can also be said that the largest differences in the optimum L/G ratio were observed.

3.2.1. *Standard flow*

According to literature, the use of internal heat exchange can cause an increase in the temperature of the stripper by approximately 5°C (Stec et al., 2015a; Oyenekan and Rochelle, 2007). Fig. 11 shows temperature profiles as a function of the packing height of the stripper for standard process configuration with and without interheating. The temperature profiles were chosen for the tests at the optimum L/G ratios equal 3.41 and 3.51 respectively. Analysis of the temperature profile in the stripper for standard flow shows that the highest temperature difference between, occurs above 10 m in the stripper condensation section, what is above the upper inter-heater height. Unfortunately, having interheating there was no increase, but the decrease of temperature for about 5-6°C for this stripper section. This is explainable because part of the heat from the reboiler is exchanged in stripper internal rather than in external heat exchangers. Hence, 3-4°C temperature bulge at $3 - 8$ m stripper height can be observed. A confirmation can be found in Table 7, comparing heat exchanged between rich-semi-lean 1, rich-semi-lean 2, rich-lean and internal stripper heat exchangers. Moreover, taking into account almost equal total solvent heat exchanged and a significant difference in total solvent cooled, it can be confirmed superiority if interheating for standard flow system.

3.2.2. *Multi absorber feed*

Similar trends in the temperature profiles for multi absorber feed

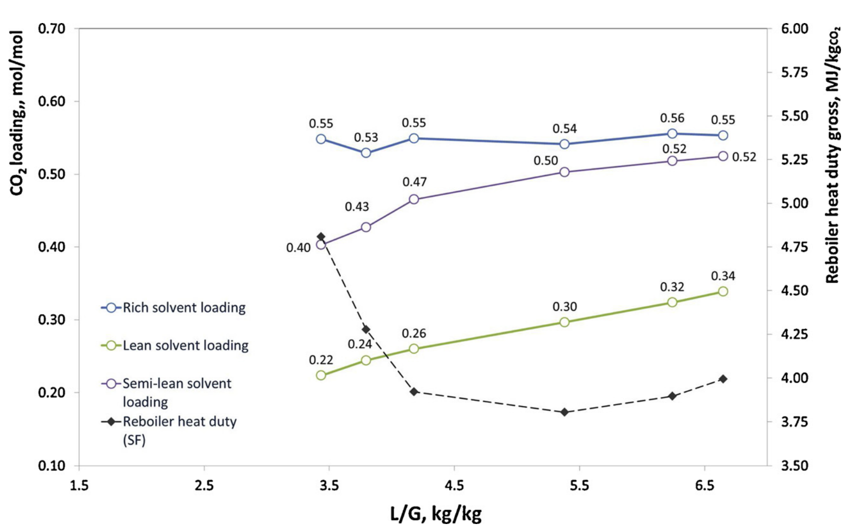

Fig. 9. The effect of the L/G ratio on CO_2 loading and reboiler heat duty for a split solvent flow system at constant CO_2 removal efficiency.

Table 6
Summary of the optimum process results for various flow systems.

Process variable	Unit	Value		
		S	MAF	SF
CO_2 recovery	%	88.3	90.4	89
Reboiler heat duty (gross)	MJ/kg_{CO2}	3.99	3.78	3.81
CO_2 in gas composition	vol%	13.34	13.37	13.29
Absorber top pressure	kPa_a	26	30	26.8
Stripper top pressure	kPa_a	30	30	30
L/G ratio	kg/kg	3.41	4.08	5.38
MEA concentration	%mass	30	30	30
Overall rich amine flow	kg/h	970	1175	1511
Rich amine mass flow–top stripper inlet	kg/h	125	125	125
Lean amine mass flow–top absorber inlet	kg/h	929	573	721
Lean amine mass flow–middle absorber inlet	kg/h	0	573	753
Semi-lean amine mass flow	kg/h	–		
Rich amine loading	mol_{CO2}/mol_{MEA}	0.53	0.53	0.54
Lean amine loading	mol_{CO2}/mol_{MEA}	0.3	0.35/0.35	0.3
Semi-lean amine loading	mol_{CO2}/mol_{MEA}	–	–	0.5
Top stripper temperature	°C	32.2	28.5	29
Lean amine temperature in stripper bottom	°C	109	109	109
Top absorber temperature	°C	46	51	44
Lean amine temperature at absorber inlet	°C	40	40	40
Lean/semi-lean amine temperature at absorber inlet	°C	–	40	40
Reboiler heating element electric power	kW	50.4	50.4	48.7

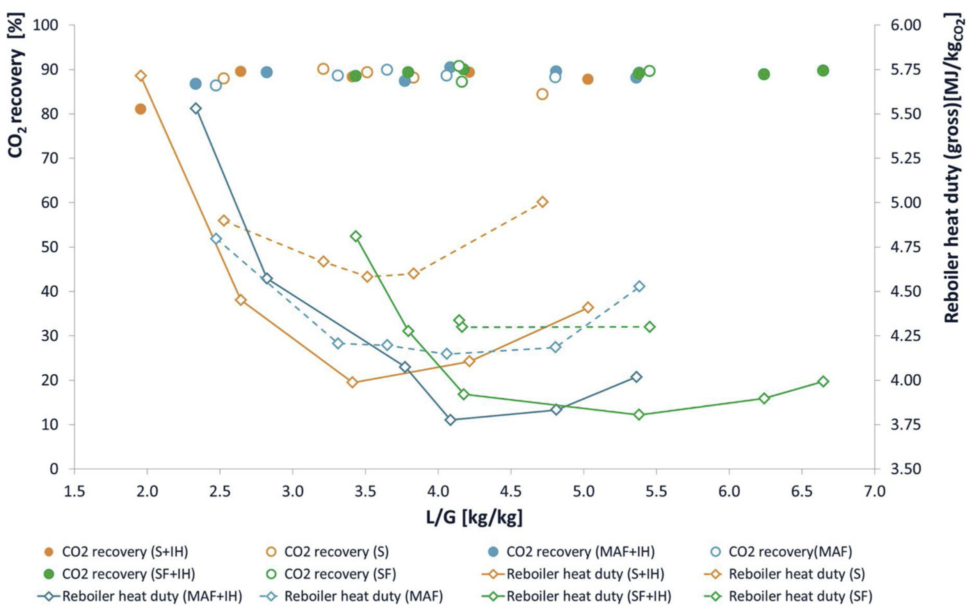

Fig. 10. CO_2 removal efficiency and reboiler heat duty comparison results for a different advanced flow system with and without stripper interheating.

process configuration can be observed in Fig. 12. The highest temperature difference between temperature profiles in the condenser section of stripper also occurred. However, the condenser section temperature was slightly lower than for standard process configuration in both cases of interheating, even if the solvent circulation was higher than for the standard flow sheet (L/G = 4.06–4.08). Thus the external condenser showed lower cooling power (Table 8).

Interheating the stripper in this process configuration increase the total solvent heat exchange while the reboiler heating element electric power was on the same level. The total solvent cooled have been lower with interheating.

3.2.3. *Split flow*

Splitting the stream, causes further enlargement of overall circulation rate up to an L/G ratio to about 5.4, therefore even higher heat values were exchanged through exchangers (Table 9). More heat was also exchanged through the stripper internal heat exchangers. The lower internal heater exchanged almost 13 kW, while the higher reached a value of 8.23 kW. The heat exchanges rates resulting mainly from temperature changes were visible also on stripper temperature profiles (Fig. 13). The differences between temperatures showed about 8°C for lower and 2-3°C for upper interheater.

For split flow even more, up to about 46°C, lowered the stripper condenser section temperature was noticed. As a result, the external condenser consumed less cooling water, thus very low heat was wasted. In spite of high L/G ratio, a comparison to standard and MAF process flow sheet total amount of heat was transferred for cooling.

Without interheating, visibly temperature decrease can be noticed in the vicinity to non-heated rich solution inlet (rich 3 amine inlet). This decrease was dictated by lower temperature in the absorber column bottom as the absorption between semi-lean amine and CO_2 was lower and less reaction heat was produced.

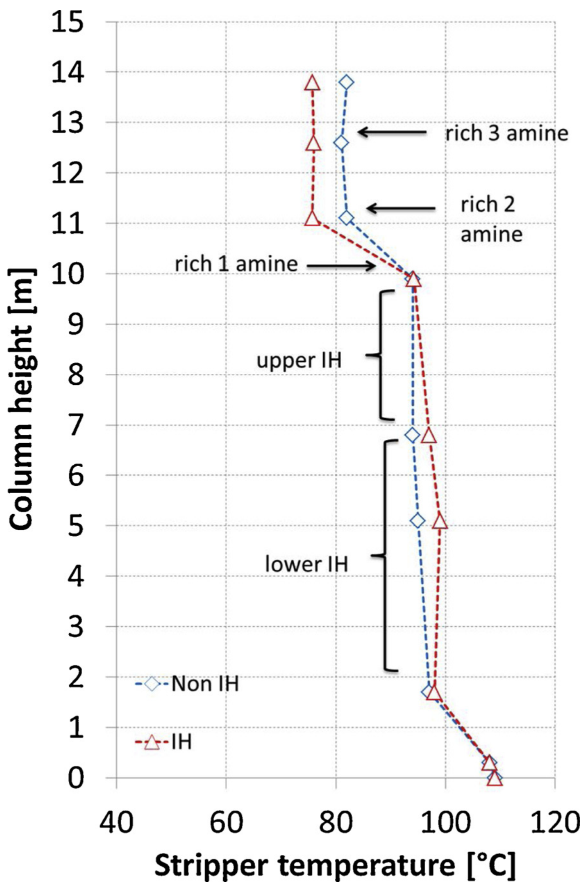

Fig. 11. Standard (S) process configuration stripper temperature profiles.

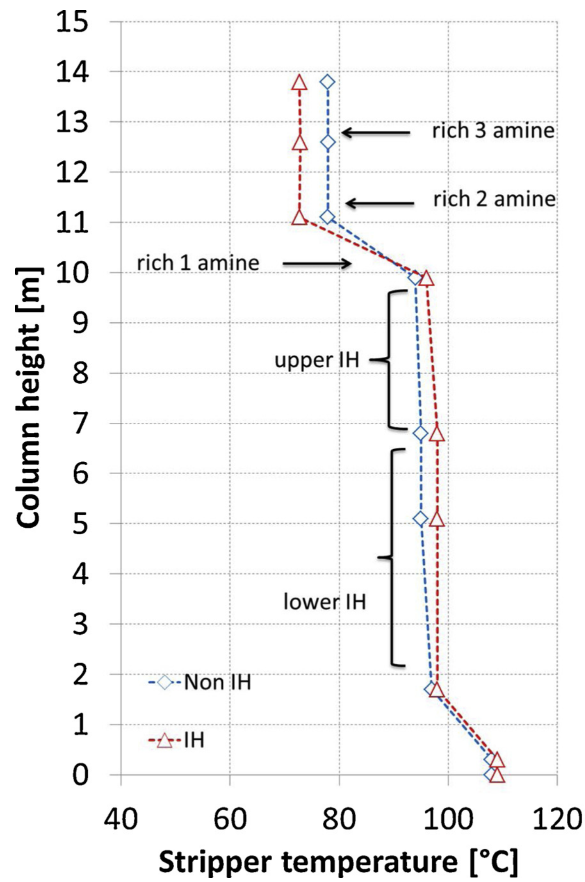

Fig. 12. Multi absorber feed (MAF) process configuration stripper temperature profiles.

Table 7
Selected equipment comparison of heat exchanged with and without stripper interheating for standard process configuration.

Equipment	Unit	Process flow sheet	
		S	S + IH
Reboiler heating element electric power	kW	63	50.4
Rich-semi-lean 1 heat exchanger	kW	22.79	17.76
Rich-semi-lean 2 heat exchanger	kW	4.14	2.34
Rich-lean heat exchanger	kW	25.01	19.6
Stripper higher internal heat exchanger	kW	0	6.04
Stripper lower internal heat exchanger	kW	0	5.57
Total solvent heat exchanged	kW	51.94	51.31
Lean solvent cooler	kW	−18.42	−12.99
Semi-lean solvent cooler	kW	–	–
External condenser	kW	−3.75	−3.22
Total solvent cooled	kW	−22.17	−16.21

Table 8
Selected equipment comparison of heat exchanged with and without stripper interheating for multi absorber feed process configuration.

Equipment	Unit	Process flow sheet	
		MAF	MAF + IH
Reboiler heating element electric power	kW	50.4	50.4
Rich-semi-lean 1 heat exchanger	kW	24.07	19.99
Rich-semi-lean 2 heat exchanger	kW	4.78	2.46
Rich-lean heat exchanger	kW	22.23	21.07
Stripper higher internal heat exchanger	kW	0	6.89
Stripper lower internal heat exchanger	kW	0	6.89
Total solvent heat exchanged	kW	51.08	57.3
Lean solvent cooler	kW	−11.74	−10.45
Semi-lean/lean solvent cooler	kW	−10.39	−9.35
External condenser	kW	−2.13	−2.18
Total solvent cooled	kW	−24.26	−21.98

4. Summary and conclusion

The objective of the MEA test phase proven the superiority of application advanced amine flow systems as well as of the novel stripper internal heater. The effects of following solvent flow systems: standard, multi absorber feed and split flow, on a reboiler heat duty and CO_2 removal have been described in detail. Each flow system showed minimum reboiler heat duty in a wide range of L/G with an assumed 90 % CO_2 recovery level. This indicates an optimum L/G ratio. Obtained results show, that MAF and SF were superior over the standard process flow system. The lowest reboiler heat duty was received for the MAF process configuration. Similarly, low reboiler heat duty was demonstrated for the SF configuration. Furthermore, the SF flow system allowed to achieve low reboiler heat duties in a wide range of L/G ratio values, thus made this configuration more flexible to operate. For the optimal conditions with a great simplification can be said that using MAF and SF comparing with S can be decreased respectively about 5,3 and 4,6% of reboiler energy.

For various flow configurations, it was experimentally shown, that for systems having interheating, the reboiler heat duties can be reduced by about 9–11%. In the interheated stripper, a part of the heat from the reboiler was exchanged inside of the stripper rather than in external heat exchangers, which reduces heat losses. During usage of the interheated stripper, the temperatures increased in in the stripper and decreased in the stripper condensation section. It allowed lowering duty of the external cooler. The lowest external condenser cooling stream needed for the split flow was achieved, while multi absorber feed and

Table 9
Selected equipment comparison of heat exchanged with and without stripper interheating for split flow process configuration.

Equipment	Unit	Process flow sheet	
		SF	SF + IH
Reboiler heating element electric power	kW	54.39	48.72
Rich-semi-lean 1 heat exchanger	kW	28.39	23.84
Rich-semi-lean 2 heat exchanger	kW	5.18	0.75
Rich-lean heat exchanger	kW	38.39	28.12
Stripper higher internal heat exchanger	kW	0	8.23
Stripper lower internal heat exchanger	kW	0	12.98
Total solvent heat exchanged	kW	71.96	73.95
Lean solvent cooler	kW	−12.4	−8.65
Semi-lean/lean solvent cooler	kW	−10.48	−8.67
External condenser	kW	−1.59	−0.56
Total solvent cooled	kW	−24.47	−17.88

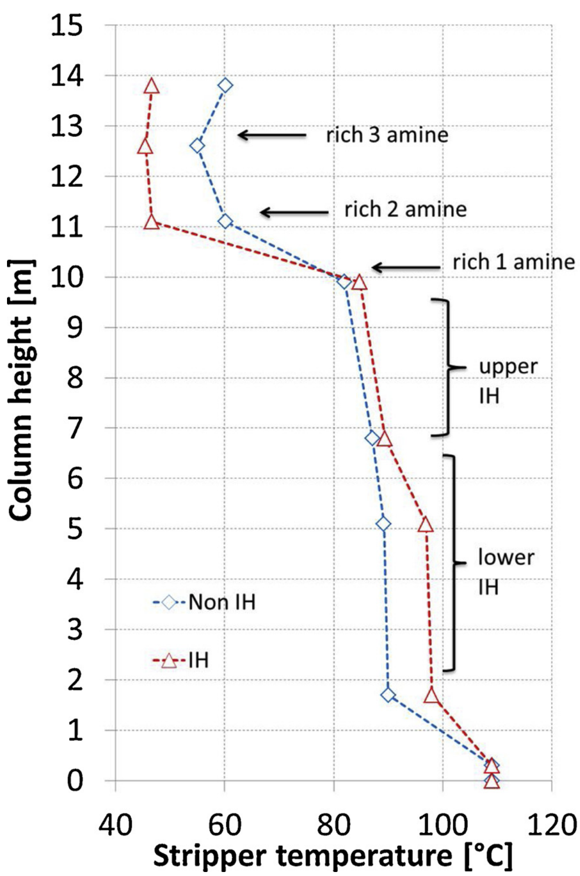

Fig. 13. Split flow (SF) process configuration stripper temperature profiles.

standard were next.

It can be expected that split-flow modification coupled with a new solvent would further decrease the energy demand the amine-based post-combustion CO_2 capture process.

Together with the process comparison, a complete economic assessment should be conducted, because, in the end, the decision on the implementation of interheated multi absorber feed or split-flow process will be determined by the overall economics of the system.

CRediT authorship contribution statement

Aleksander Krótki: Conceptualization, Methodology, Software, Data curation, Writing - original draft, Writing - review & editing. **Lucyna Więcław Solny:** Conceptualization, Writing - original draft, Writing - review & editing. **Marcin Stec:** Methodology, Software, Data curation, Writing - original draft, Writing - review & editing. **Tomasz Spietz:** Investigation, Data curation, Formal analysis. **Andrzej Wilk:** Investigation, Formal analysis. **Tadeusz Chwoła:** Investigation, Validiation, Formal analysis. **Krzysztof Jastrząb:** Supervision.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Acknowledgements

The results presented in this paper were obtained during research co-financed by the National Centre for Research and Development in the framework of Contract SP/E/1/67484/10 – Strategic Research Programme – Advanced technologies for energy generation: Development of a technology for highly efficient zero-emission coal-fired power units integrated with CO_2 capture.

This activity has received funding from the European Institute of Innovation and Technology (EIT). This body of the European Union's Horizon 2020 research and innovation programme.

Project: CO_2 methanation system for electricity storage through SNG production (CO2-SNG) Reference: 30_2014_IP108_CO2-SNG.

Appendix A. Supplementary data

Supplementary material related to this article can be found, in the online version, at doi:https://doi.org/10.1016/j.ijggc.2020.103014.

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