RETROFIT DESIGN FOR GAS SWEETENING PROCESSES

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Gas sweetening using amines is an important gas treatment process. Existing gas sweetening units may be revamped to achieve a variety of aims, for example to expand processing capacity, to reduce operating costs, to accommodate a different feed, to reduce emissions levels, etc. Current practice for retrofit design is to use engineering judgement and experience and simulation studies to identify process modifications, but this approach can overlook beneficial options. Instead, a systematic approach for retrofit of gas sweetening units is proposed. The approach involves process simulation, using simplified models, and optimisation of both flowsheet configuration and operating conditions using stochastic optimisation, namely simulated annealing. The optimisation objective considers operating costs, and constraints in the problem formulation guide the solution towards retrofit designs that are practical and have low capital costs. The method is applied to an industrial case study where the CO_2 content in a natural gas feed has increased. Modifications that allow the sales gas specifications to be met are identified and evaluated.

KEYWORDS: simulated annealing, revamp, chemical absorption, MDEA

INTRODUCTION

Gas sweetening, the removal of acid gases (e.g. H_2S, CO_2) from a gas feed using chemical absorption, is a key operation in gas treatment. Figure 1 shows a typical process flowsheet. Retrofit design of these processes involves identifying suitable process modifications that will allow revamp objectives to be met, e.g. increasing plant capacity, reducing emissions levels, improving energy efficiency. Rising energy prices, increasing demand and increasingly stringent environmental regulations motivate gas producers to undertake process revamps. A few retrofit studies are reported in the literature (Daviet *et al.*, 1984; Gagliardi *et al.*, 1989; Spears *et al.*, 1996; Carter *et al.*, 1998).

Various modifications can be carried out to achieve retrofit targets. Operational changes are frequently an attractive option for process retrofits, because in most cases little or no capital expenditure is required. In gas sweetening flowsheets, these operational changes include changing the feed gas temperature and pressure, changing the amine inlet temperature, changing the amine concentration and changing the condenser temperature. More major changes include switching to a different solvent, to a mixed solvent or to a proprietary 'energised' solvent.

Over the years, the basic flow scheme shown in Figure 1 has evolved to include more complex structural options. These complex options include multiple-feed absorbers, single-stage flash units for absorbent recovery and liquid sidedraws from the absorption or

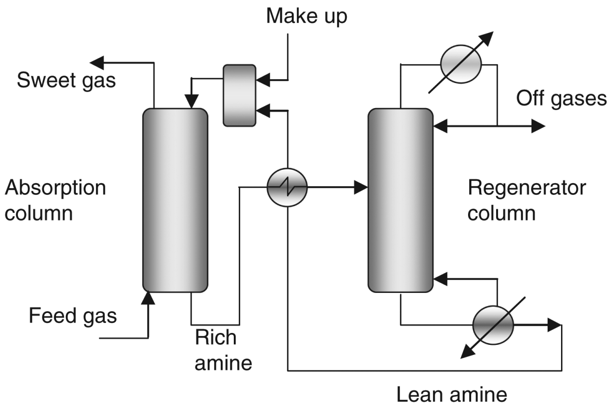

Figure 1. Basic flow scheme for a gas sweetening process using an alkanolamine solvent

desorption columns. Structural changes for retrofit of gas sweetening unit have been explored less than have operational changes. This is possibly because they are not as widely accepted, are more difficult to identify or evaluate, or are perceived as harder to engineer, especially when minimising downtime is critical.

In current practice for retrofit design, relatively few operational and structural changes are considered. Designs are typically based on engineering judgement, then the modified processes are simulated and analysed. Good results can be achieved, in terms of both process improvement and process economics. However, such an approach may not allow the strong interactions between key flowsheet design parameters to be explored. In order to find cost-effective and energy-efficient solutions to the retrofit design problem, a systematic approach is needed that can effectively generate and evaluate many design options and their combinations.

This work develops a systematic, automated method for retrofit design of gas sweetening units. A superstructure is developed which includes structural options, such as side-draws from and multiple feeds to the absorption and desorption columns, intermediate flash units, etc. Heat integration implications and opportunities are assessed during process simulation and optimisation. The structural options and operating conditions are optimised simultaneously. Stochastic optimisation is applied, to allow global, rather than local, optima to be found. The optimisation approach is known as simulated annealing. In this algorithm, new designs are generated randomly from a set of options, are simulated and are evaluated against an objective function. Designs with improved performance are accepted, but there is a randomised probability that inferior designs will also be accepted.

RETROFIT DESIGN BY SUPERSTRUCTURE OPTIMISATION

Patil (2005) developed a superstructure-based optimisation approach for grassroots synthesis and design of gas sweetening flowsheets. Many structural options are included in the superstructure. The superstructure and operating conditions are optimised simultaneously to generate flowsheets meeting problem specifications. Figure 2 shows the superstructure developed by Patil (2005) which includes the following structural options for synthesis and design of gas sweetening flowsheets:

- *Multiple feeds to the absorption column*: Multiple feeds effectively allow intermediate cooling of the absorbent which reduces solvent losses to the gas phase and enhances absorption of acid gases.
- *Multiple sidedraws from the absorption column*: In the absorption column, the bulk of the acid gases are removed in the bottom few stages. The remaining traces of acid gases are then removed. After removing these traces, the solvent still has a relatively low concentration of acid gases, so part of this liquid can be withdrawn from the absorption column and recycled back, without requiring solvent regeneration. This option reduces the reboiler and condenser duties of the desorption column.
- *Sidedraw from the desorption column*: This arrangement is known as a split-loop arrangement. Part of the liquid stream is withdrawn from an intermediate stage of the desorption column and fed back to an intermediate stage of the absorption column, reducing the reboiler and condenser duties. However, this 'semi-lean' absorbent is less pure than the lean solvent produced in the desorption column reboiler, and is therefore less able to absorb acid gases. Thus there are trade-offs between the quality of the sweet gas and energy demand.

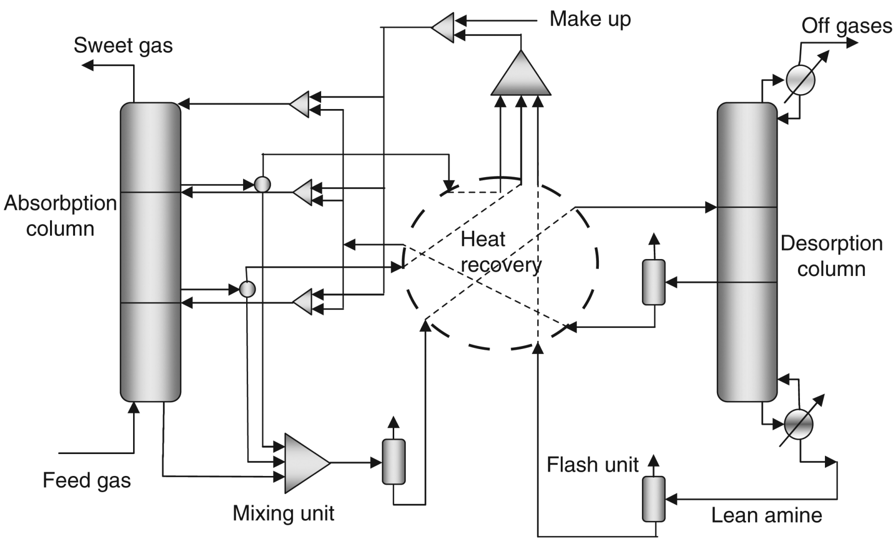

Figure 2. Superstructure for optimisation of gas sweetening flowsheets (Patil, 2005)

- *Flash units*: The absorption column operates at a high pressure, while desorption takes place close to atmospheric pressure. An intermediate flash can exploit this pressure difference to provide an energy efficient method of removing some of the acid gases from the rich solvent stream. The flash unit reduces the flow rate of acid gases to the desorption column and so reduces its reboiler and condenser duties.
- *Heat integration*: Various streams from the absorption and the desorption columns need to be cooled or heated. Recovery of available process heat can reduce operating costs significantly.
- *Column design*: The design of individual columns (operating pressures, feed temperature, etc.) impacts on the performance and energy demand of the overall flowsheet.

Patil (2005) has demonstrated the application of this optimisation approach for a number of gas sweetening process synthesis and design problems. This work extends the approach to the retrofit of gas sweetening units. The flowsheet structure and operating conditions can be optimised to de-bottleneck an existing plant.

RETROFIT CASE STUDY

The optimisation-based design approach will be described through its application to an industrial case study. The existing gas sweetening unit uses methyl diethanolamine (MDEA) to selectively remove H_2S, rather than CO_2, from natural gas. Gas feeds from different wells and of considerably different quality are treated. Ambient conditions are warm: the amine inlet temperature is normally 56°C. Table 1 presents the 'current' plant operating data.

Table 1. Operating data for the absorption and desorption columns of an industrial gas sweetening plant

Absorption Column		Desorption Column	
Number of trays	20	Number of trays	20
Column pressure, bar	55	Column pressure, bar	3
H_2S in gas feed, ppm	380	Condenser temperature, °C	30
CO_2 in gas feed, mol%	4.8	Feed tray location	4
Gas feed temperature, °C	43.9	Feed temperature, °C	97.2
Amine conc. in absorbent, wt%	36.0	H_2S efficiency, %	39#
Amine inlet temperature, °C	55.7	CO_2 efficiency, %	2.7#
Amine flow rate, $kmol \cdot h^{-1}$	691.5		
Feed gas flow rate, $kmol \cdot h^{-1}$	1096		
H_2S efficiency, %	56.0#		
CO_2 efficiency, %	8.0#		

#specified values.

Table 2. Operating scenarios of gas sweetening unit

Scenario	Inlet H_2S, ppm	Outlet H_2S, ppm	Inlet CO_2, mol%	Outlet CO_2, mol%
Normal	380	2–3	3.5	1.9
Current	380	2–3	4.8	2.8
Target	380	<6	4.8	2.2

In the 'current' scenario, the CO_2 concentration in the feed gas has increased suddenly (compared to 'normal' operation), resulting in high levels of CO_2 in the sweet gas. Table 1 shows the current values of H_2S and CO_2 in the feed gas. The design specification for the treated gas from the absorption column is 2.2% (max) CO_2. However, as shown in Table 2, the outlet from the absorber contains 2.8% CO_2. The inlet and outlet concentrations of acid gases in normal and current operating modes, and the 'target' for retrofit are given in Table 2. The gas sweetening plant configuration is shown in Figure 3.

OPERATING LIMITS OF EXISTING UNITS

During retrofit design, it is important to account for limits associated with existing equipment or equipment configurations. In the proposed approach, these limits are imposed as optimisation constraints:

- The lean amine recycle stream is air-cooled. Therefore the minimum possible temperature of the lean amine stream depends on the ambient temperature. In the operator's

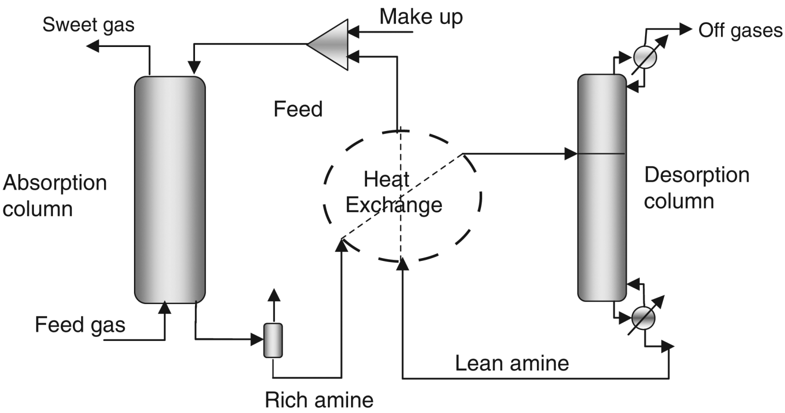

Figure 3. Flow configuration of the current gas sweetening unit

experience, a minimum temperature of 50°C is achievable most of the year. During winter, temperatures as low as 35°C are achievable.

- The normal temperature of feed gas is around 26°C; it is heated to 42°C before entering the absorber. In the view of the operator, the difference between the temperatures of the feed gas and lean amine should be in the range 8 to 15°C to avoid foaming due to condensation of hydrocarbons in the bottom of the absorber.
- The lean amine pump is a positive displacement pump with a fixed flow rate of 78 USGPM (17.7 $m^3 \cdot h^{-1}$). Any increase in flow rate will require a larger capacity pump.
- The current concentration of MDEA is 36%. It is a common practice to keep the amine concentration as low as possible to minimise corrosion and cost of inhibitors. Although the plant was designed for 50% MDEA concentration, the amine concentration cannot be increased to more than 40% as no corrosion inhibitors are used.
- In the existing absorption column, there are three amine inlets: at the top of the column (stage 1) and at stages 6 and 10; currently only the amine inlet on stage 1 is used.

VAPOUR–LIQUID EQUILIBRIUM (VLE) AND PROCESS MODELS

The validity of the optimisation results depends heavily on the validity of both the process and VLE models. In this case study, we have used a simplified simulation model developed by Patil (2005) for the simulation of the absorption and desorption columns. The model modifies the Kremser group method (King, 1980) by including component efficiencies to account for non-equilibrium effects of the chemical absorption reactions and of the mass transfer. Patil (2005) demonstrated the applicability of this model for simple and complex gas sweetening flowsheets against published plant data and HYSYS simulation results. To represent the vapour–liquid equilibrium of the H_2S-CO_2-MDEA-H_2O system, the VLE model of Kent and Eisenberg (1976) model has been modified (Malik, 2005). Malik (2005) tested the new VLE model against experimental data and demonstrated the application of the model to various commercial operating units. The incentive for using these models is that they are simple enough to be used for flowsheet optimisation, yet accurate enough to provide meaningful results.

The models of Patil (2005) and Malik (2005) are applied to simulate the 'normal' and 'current' operating cases; results are compared to plant data and to HYSYS simulation results in Tables 3 and 4. The results show that the model predictions are in reasonable agreement with the plant data, and that the model captures the key differences in process performance for the two operating modes.

PROCESS OPTIMISATION

The stochastic optimisation framework is applied to generate, simulate and evaluate flowsheet design alternatives. The objective function is to minimise the operating costs; a penalty function accounts for the CO_2 purity specification. Only variables which incur relatively small structural and operational changes and insignificant capital cost are selected as optimisation variables. These variables are the solvent and gas feed temperatures, the amine concentration and the split fraction to the absorption column. Simulated

Table 3. Comparison of model results for case study: 'normal' operation

	Plant	Model	HYSYS
H_2S in sweet gas, ppm	2–3#	0.34	13
CO_2 in sweet gas, %	>2.0	1.97	1.8
Rich amine temperature, °C	60.6	63.1	61.2
H_2S load in lean solvent*	–	4e−5	4.9e−8
CO_2 load in lean solvent*	–	0.03295	0.003617
Boilup ratio§	0.18	0.18	0.18

#calculated; *moles of acid gas per mole of amine; §ratio of vapour to product flow rates in reboiler

Table 4. Comparison of model results for case study: 'current' operation

	Plant	Model	HYSYS
H_2S in sweet gas, ppm	2–3#	0.5	57
CO_2 in sweet gas, %	2.8	3.08	2.8
Rich amine temperature, °C	60.6	65.6	62.2
H_2S load in lean solvent*	–	3e−5	4.9e−8
CO_2 load in lean solvent*	–	0.03397	0.003617
Boilup ratio§	0.18	0.18	0.18

#Calculated; *moles of acid gas per mole of amine; §ratio of vapour to product flow rates in reboiler

Table 5. Optimisation results for the retrofit case study

Optimisation variable	Base case	Optimised case
Solvent feed temperature (°C)	55.7	50.5
Gas feed temperature (°C)	43.9	36.0
Amine concentration (wt%)	36.0	39.8
Amine fed to stage 1 (fraction)	1.0	0.1
Amine fed to stage 6 (fraction)	0.0	0.35
Amine fed to stage 10 (fraction)	0.0	0.55
CO_2 in treated gas (mol%)	2.8	2.2

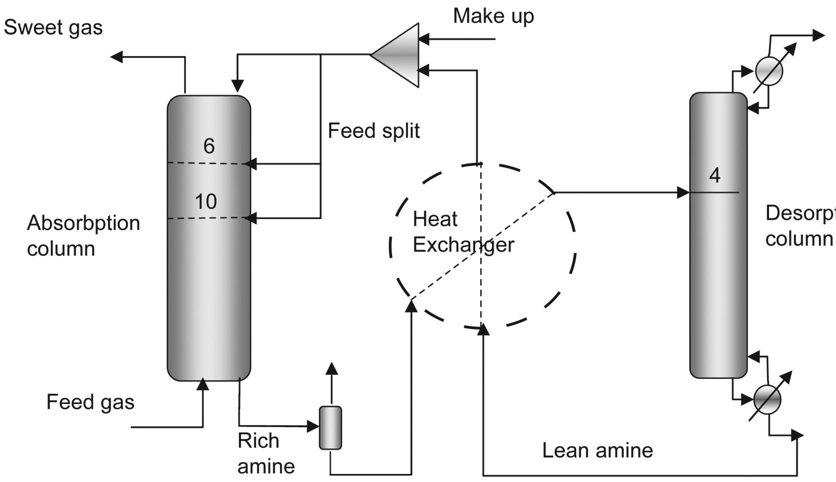

Figure 4. Modified flowsheet for retrofit design

annealing is used as an optimisation technique to optimise the operating and structural options (Malik, 2005). Table 5 presents the optimisation results. The retrofit target (CO_2 in treated gas no greater than 2.2 mol%) is met by changing operating conditions and making structural modifications. Figure 4 presents an optimised flowsheet: by cooling the gas inlet to 36°C and the absorbent to 50°C, and introducing the cooler absorbent at stages 1 (10%), 6 (35%) and 10 (55%), the absorption column can more effectively remove CO_2. Although the absorbent flow rate is unchanged, the increase in amine concentration from 36% to nearly 40% also provides additional absorption capacity.

CONCLUSIONS

Current practice for the retrofit design of gas sweetening plants is not systematic, using time-consuming, manual simulations and ambiguous engineering and operating insights. A new design procedure, based on optimisation of a set of structural options and a range of operating conditions, generates optimised designs. Operating costs and operational constraints and heat integration calculations are included in the problem formulation. Simplified process and vapour–liquid equilibrium models reduce the side of the optimisation problem.

The method is demonstrated using an industrial case study in which the acid gas composition has increased, compared to normal operating conditions. Practical constraints and limits at the plant are accounted for by constraining optimisation variables. As a result, the design generated is cost-effective and practical. The design method can be for a variety of retrofit scenarios.

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