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Project Design and Control Considerations on Gas Sweetening Processes

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Amine plants for gas treatment are well-known processes to remove a series of compounds either upstream some plant operations particularly sensitive to these chemical species or before sending to vent/stack a process stream so as to match environmental regulations. There is a long experience in the design of amine plants and in their typical configuration of energy integrated absorber and stripper. Nevertheless, the design and the unit operations of current amine processes directly come from feasibility studies based on steady-state simulations and a series of process control considerations can be pointed out by means of the process dynamic simulation.

1. Introduction

Process design takes into account many technical and economical aspects to obtain the maximum profit by a minimum investment. Actually, it accounts for raw material costs, product market prices, purchase and installation costs for process units, reactors, instrumentation, etc. (Douglas, 1988). When the process design is particularly detailed, the resulting plant layout better matches the market requirements and, consequently, the net present value obtained by operating the plant is somehow increased.

Easy to realize that these feasibility studies of process design are one of the most important steps to set up a plant since an erroneous estimation in designing a single unit operation can easily lead to relevant losses not only in the capital investment, but even in the net operating margin, by also generating undesired longer breakevens and payback times, increasing the working capital, and reducing the return of investment. This pushed years ago to the development of tools and software to support at best the process engineer in the plant design leading to the current widespread commercial process simulators based on mass and energy balances, thermodynamic and hydraulic relationships and so on, which allow the aprioristic evaluation of the proper plant layout to satisfy the process specifications.

Unfortunately, the current steady-state process design approach is unable to provide the "real optimum" of the plant (Signor et al., 2007) and it is more and more evident the need for coupling the dynamic simulation with the process design. Even though it seems

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to be a paradox, as the dynamic simulation is time-dependent, whereas the process design is traditionally not, the "optimal" solutions obtained by steady-state simulations cannot account for some important aspects related to process unit dynamics and process transients. In other words, the optimal steady-state solution is "non-optimal" and sometimes unsafe or not so appealing from an economical point of view. This is not a surprise if we think that a steady-state simulation cannot anyhow account for process transients, process stability and plantwide controllability, market uncertainties and demand peaks, plant flexibility and plant operability, and off-spec periods (Manenti and Rovaglio, 2008; Lima et al., 2009; Manenti, 2009; Dones et al., 2010) to quote a few.

Conversely, dynamic simulation gives the possibility to account for all the aforementioned points and even to consider them not only for control and for operational purposes, but also even for the process unit design.

For example, the optimal solution coming from a steady-state simulation may make the plant unstable and, hence, hard to control during some specific process transients by leading to prolonged oscillations in some key-variables and, hence, enlarging the off-spec periods. An optimal process design based on the dynamic simulation accounts for these problematic issues, by overcoming the myopic design based on steady-state simulation.

A brief state-of-the-art of commercial tools for process simulation is given in paragraph 2. The case of Mono-Di-Ethanol-Amine (MDEA) process for H_2S removal is selected as case study and introduced in paragraph 3. Both steady-state and dynamic simulations are developed for design purposes. Some interesting solutions pointed out by the dynamic simulation are reported in paragraph 4.

2. Brief history and state-of-the-art of process dynamic simulators

The first process simulation packages were Speed-up, developed in '60s of the previous century by Sargent and co-workers (Sargent and Westerberg, 1964; Sargent, 1967). At the same time, some software societies were commercializing software solutions; for example, Esscor already had software to support electrical engineers in plant design in 1967. Nowadays, this family of pioneer software is still the kernel of the modern commercial packages for steady-state, but even dynamic, process simulation. Actually, Speed-up was the basis of Aspen (by Aspen Technology), whereas Dynsim (by SimSci-Esscor, Invensys Operations Management) includes the section *Electrical* coming from the original Esscor's package. These tools started spreading only in '90s as testified by the literature (Gani et al., 1992; Perregaard et al., 1992; Pantelides and Barton, 1993; Gani and Grancharova, 1997) by showing their effectiveness to analyze process dynamics and to verify control schemes. Only later the use of process dynamic simulators has started spreading from those fields traditionally time-dependent towards other fields completely time-independent such as process design (Luyben and Luyben, 1997; Signor et al., 2007). The current commercial software available for dynamic simulation has a complex architecture including multipurpose graphical interfaces, thermodynamic libraries, model libraries for unit operations, differential and differential-algebraic solvers, and support tools. Many simulators are field-proven by industrial applications in many areas from the oil & gas to the fine chemical, from the power generation to the petrochemical and their main advantage is in the intrinsic multipurpose nature, the user-friendly interface, and especially in large databases containing most common chemical species.

Some simulators are Petro-SIM (by Profimatics, KBC Advanced Technology), OmegaLand (Yokogawa), AudySim (Trident), gPROMS (PSe), but, historically, only three software were particularly spread and with a considerable market slice: DYNSIM by SimSci-Esscor, an operating unit of Invensys Operations Management, HYSYS originally developed by Hyprotech, and ASPEN by Aspen Technology.

This was the picture up to 2004, when Aspen Technology, which was characterized by a decreasing market slice, acquired Hyprotech and, hence, HYSYS. Such an operation led to the fast development of ASPENHYSYS 2004, based on the original kernel of HYSYS with the aim of significantly improving sales. Facing a probable brilliant commercial strategy, Aspen Technology was condemned for monopoly by US Government and the newborn package was divided in four parts, each of them developed by a different society. Nowadays, two packages are born out from ashes of HYSYS: UNISIM, developed by Honeywell, and ASPENHYSYS, developed by Aspen Technology, both based on the same numerical kernel. Together with DYNSIM, these three packages still cover the largest worldwide market slide in process dynamic simulation.

3. MDEA plant

Apart from the process dynamic simulator adopted, process and process control engineers more and more frequently make use of the dynamic simulation for design purposes. The motivation is that each kind of event and perturbation generates dynamic behaviors that easily take to undesired (and sometimes dangerous) operating conditions such as vibrations, pressure drops, and flow peaks by causing instability and off-spec production periods.

3.1 Process description

The selected case in study is a natural gas sweetening process under construction in Middle East and specifically an amine process to remove sulfuric acid (H_2S). A qualitative process scheme is reported in Figure 1. The process consists of an absorber where the natural gas enters from the bottom and encounters a water/MDEA solution (60%/40%) to remove impurities (mainly H_2S) contained in the inlet stream. The natural gas exits from the top of the absorber, while the water/MDEA solution rich in H_2S exits from the bottom. The absorber operates at 35-40 atm and at a relatively low temperature to favor the removal operation. The liquid stream exiting the bottom of the absorber must be preheated before entering the regeneration splitter. Absorber and splitter are integrated for energy saving as well as to reduce variable costs. The stripper operates at lower pressure and higher temperature with respect to the absorber to favor the regeneration of the water/MDEA solution. Impurities removed form the natural gas stream exit the top of the stripper together with a relevant quantity of water. Thus, a make up of water is needed. Conversely, MDEA losses must be whenever prevented. In fact, in process simulation it is always assumed that the appropriate amount of MDEA is loaded at the beginning of the startup and no make-up is needed during the operations.

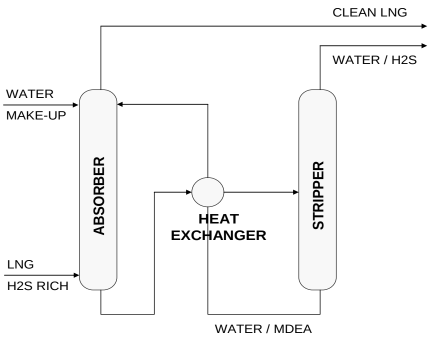

Figure 1: qualitative process scheme

3.2 Plantwide control scheme

The control scheme of gas sweetening processes usually consists of six loops: a flow controller regulates the water make up by accounting for the water amount exiting the top of the stripper; two pressure controllers, one for each column, acts on the gas streams exiting the top of the columns; a flow controller regulates the water/MDEA recycle stream from the stripper to the absorber; a level controller manages the bottom holdup of the absorber; a temperature controller regulates the top temperature of the stripper by acting on the stripper reboiler duty.

It is worth noting that no level controllers are needed for the liquid holdup at the bottom of the stripper. Actually, the combination of the flow controller on the water/MDEA recycle, the level controller at the bottom of the absorber, and the assumption that the amount of MDEA is always the same as no losses are allowed at the top of the stripper ensures that the level at the bottom of the stripper is indirectly controlled.

4. Process dynamics and control considerations

If the steady-state simulation is nowadays the predominant tool for process design, the solutions for plant operability and controllability are still based on process experience. If years ago there was no possibility to check these solutions before their implementation by the field, today there is the possibility to use the dynamic simulation to validate a priori their effectiveness. For the sake of conciseness, we briefly propose a series of considerations directly coming from the dynamic analysis of the MDEA plant described above.

A first consideration deals with the duty provided to the reboiler of the regeneration tower (splitter). The reboiler and the related control configuration are qualitatively reported in Figure 2. On the left side of Figure 2 (control scheme No. 1), the flow rate control acts on the inlet steam to the tube side of the reboiler. The regulation of the duty supplied to the stripper is defined by a different pressure value within the tube side. Assumed a constant pressure of the steam network, the valve on the steam line could increase/decrease the pressure drop on the steam line itself and thus induce a corresponding decrease/increase in the reboiler pressure with a variation in the condensation temperature. Hence, the heat exchange is characterized by a variation of the thermal driving force to control the system. On the right side of Figure 2 (control scheme no. 2), the alternative control scheme acts on the outlet condensate flow. The regulation of the duty is performed by regulating the liquid holdup in the heat exchanger (hence, by partially wetting the tube bundle) to modify again the effective heat exchange. The control scheme No. 1 always has a certain pressure drop even when the valve is completely open. Conversely, the control scheme No. 2 has the advantage of operating at the maximum pressure available since no pressure drops are present upstream the reboiler. From a process design point of view, a reboiler significantly smaller could be used if the control scheme No. 2 is adopted. Since such a configuration is starting spreading in the process industry, it is important to underline that the dynamic simulation easily highlights some possible shortcomings of the control scheme No. 2. First of all, the control action is somehow slower by using the control scheme No. 2 against the control scheme No. 1. In addition, the control scheme No. 2 might lead to some instabilities during process transients (Signor et al., 2007).

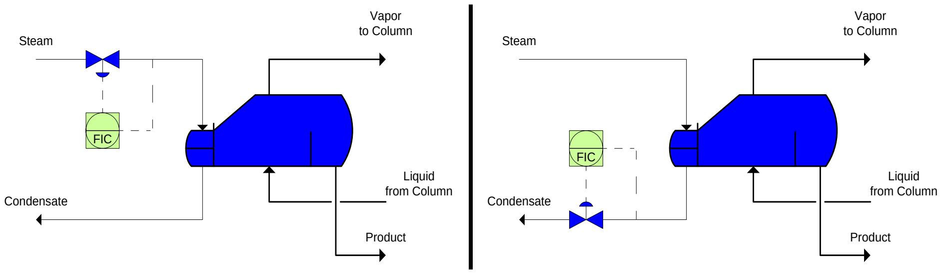

Figure 2: duty control on vapor inlet flow (steam pressure on tube side) on the left – control scheme no. 1; duty control on condensate discharge flow (effective wet area on tube side) on the right side – control scheme no. 2

Another consideration deals with the temperature controller adopted to regulate the top temperature of the stripper by acting on the bottom reboiler duty.

Before discussing this point, it is important to remark that many process control solutions for chemical and oil & gas processes are dictated by the experience rather than by a rigorous validation. In addition, if we account for the traditional inertia of process industry to implement new solutions, it is not surprising if many control schemes designed some decades ago are still used in modern processes and in on-going process designs. Specifically, in the case of gas sweetening processes, the temperature controller used for the stripper is the direct consequence of this scenario. Actually, gas sweetening processes were originally designed to process and remove a large amount of H_2S. This meant that a relevant molar fraction of H_2S was present on the stripper trays and the temperature was sensitive to it.

On the other hand, the current major task of some gas sweetening processes is to remove a very small fraction of H_2S to satisfy the more and more stringent environmental regulations. In these specific cases, since the H_2S fraction is in the order of few ppm, the temperature controller is no longer sensitive to possible variations in the H_2S molar fraction and the composition is practically constant with 60 % of water and 40 % of MDEA. Thus, the temperature controller is useless. Whereas it is not possible to check this consideration via steady-state simulations, the implementation of grade and load changes in the process dynamic simulation clearly confirms it.

5. Conclusions

The paper provides some process design and control considerations coming from the use of dynamic simulation to investigate process operability and controllability. It is mainly aimed at emphasizing some interesting solutions against those well-established solutions dictated either by the experience or by means of the steady-state analysis. Possible process design and control improvements for MDEA plants are proposed.

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