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HYDROCARBON PROCESSING®

Optimize capacity for large ethylene oxide reactors

Many factors must be considered in fine-tuning unit design

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Ethylene oxide (EO) is an important chemical intermediate, with annual global consumption of approximately 30 million tpy. Most of the EO is hydrolyzed to produce ethylene glycol (EG), but other important products are also made from EO, including ethanolamines, glycol ethers and various ethoxylates.

EO is produced commercially via a vapor-phase reaction of ethylene and oxygen over a silver-based catalyst. This reaction is exothermic, and the unselective reaction to complete combustion is even more so. Conversion must be kept low to ensure high selectivity to EO. Accordingly, heat must be removed from the reacting process gas efficiently within the reactor to allow good control at low conversion. The reactor is effectively a multitubular heat exchanger, with catalyst loaded inside the tubes and a coolant on the shell side. Current commercial reactors use boiling water as the coolant and generate steam directly within the reactor unlike earlier designs that used an intermediate organic fluid, either boiling or non-boiling.

A key performance measure is EO selectivity, defined as moles of EO produced in the reactor per 100 moles of ethylene converted. Improvements in catalyst over time have led to higher EO selectivity and accordingly with less heat removed per ton of EO produced. The reduction in heat removal has allowed some relaxation in variables such as productivity and tube diameter, and it became easier to meet certain technical constraints. However, this has allowed the process engineer more scope in improving the process economics with great care.

Modern commercial EO reactors have tubes that range in diameter from about 1.5 in. to 2 in. (38 mm to 50 mm)

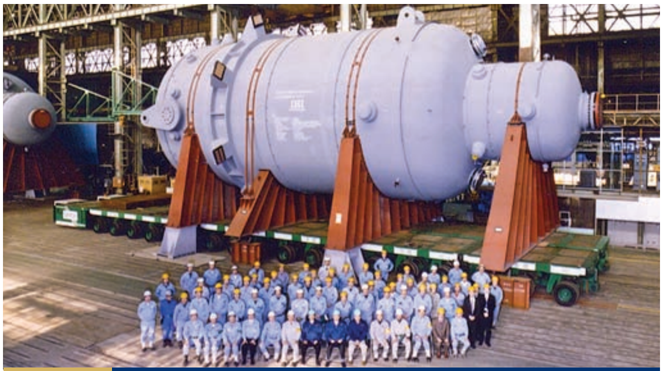

FIG. 2 Photo of a typical large modern EO reactor.

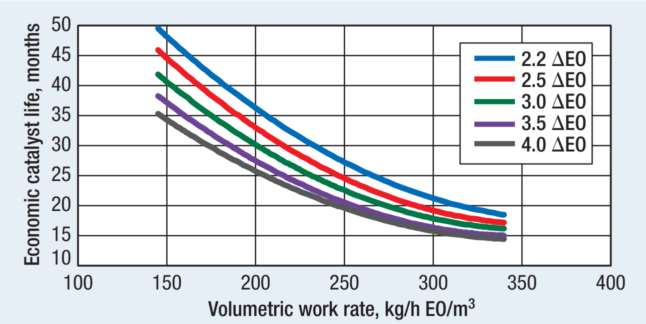

FIG. 1 Economic service life of EO catalyst.

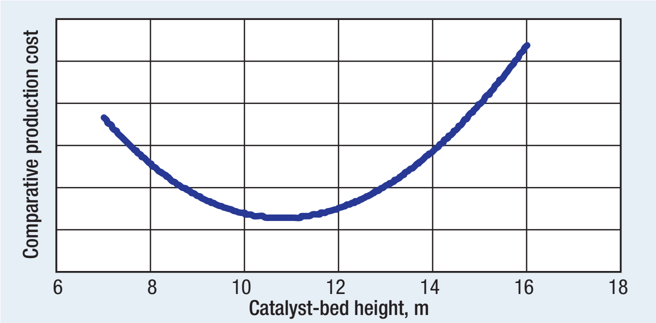

FIG. 3 Production costs vs. bed height for EO reactor.

for the outside diameter (OD). Tube-wall thickness is controlled by external pressure from the boiling water side and is proportional to diameter. Therefore, increasing the tube diameter does not necessarily reduce the total weight of tubes for a fixed catalyst volume. The point of diminishing returns in reactor cost has been reached.

Designing for normal operating conditions at steady state does allow increasing the tube diameter as catalyst selectivity increases. However, this must be weighed against the benefits of a robust design with smaller diameter tubes that has a greater ability to handle process disturbances without creating "hot spots" that can damage catalyst and lead to other problems such as deteriorating product quality.

CONVERSION AND PRODUCTIVITY

In common industry terms, conversion is often described in terms of ΔEO (moles of EO produced per 100 moles of reactor inlet gas, often in the range of 2 to 4). Productivity is defined as work rate or kg/hr of EO produced per metric ton (weight work rate) or per cubic meter (volumetric work rate) of catalyst. Volumetric work rate is often in the range of about 150 to 300.

Selectivity to EO decreases as ΔEO increases; raw material costs (ethylene and oxygen) increase with ΔEO. Conversely, recycle gas flow decreases along with its associated capital and energy costs, leading to an economic optimum ΔEO.

As the catalyst ages, it loses activity and selectivity. The normal practice is to increase reactor temperature to maintain production as activity decreases. In most cases, the catalyst is replaced at or near the end of its economic service life before reaching the reactor temperature limit. The economic optimum life occurs when the sum of average operating and catalyst costs reaches a minimum, although other factors such as periodic maintenance or inspection schedules can frequently affect the timing. The main operating costs are due to ethylene and oxygen consumption, ethylene cost being typically about 70% of the total operating cost. Note that catalyst cost includes not only the cost of buying catalyst but also direct costs of changing catalyst, silver leasing and silver-handling losses. The economic catalyst life depends on the ratio of catalyst to ethylene cost. Fig. 1 shows the results of a typical study on economic catalyst life.

Reactor size limits. As plant sizes have continued to increase, EO reactor sizes have increased dramatically over the last 10 years. The maximum reactor diameter is usually determined less by fabrication limits than by practical considerations of handling, transporting and erecting at the plant site. Although site conditions vary, the diameter limit is approximately 28 ft (8.6 m) in a general case. Such

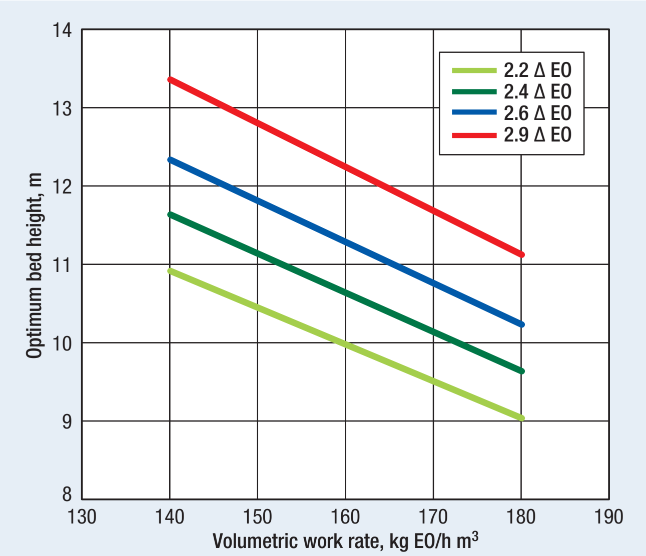

FIG. 4 Optimum catalyst bed height based volumetric work rate.

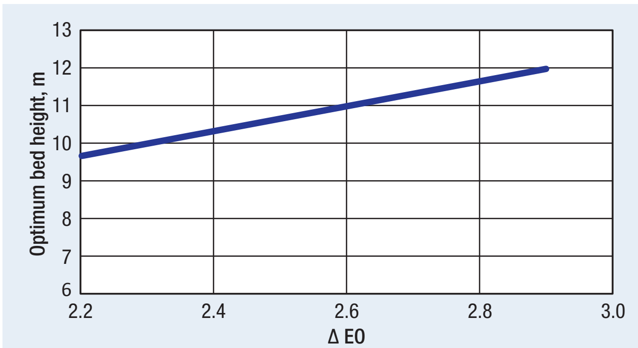

FIG. 6 Optimum catalyst bed height vs. ΔEO.

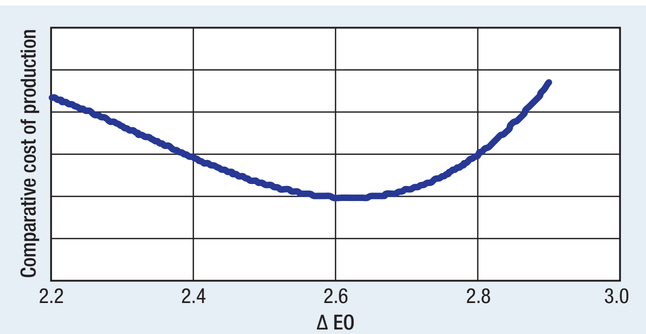

FIG. 5 Cost of production vs. ΔEO.

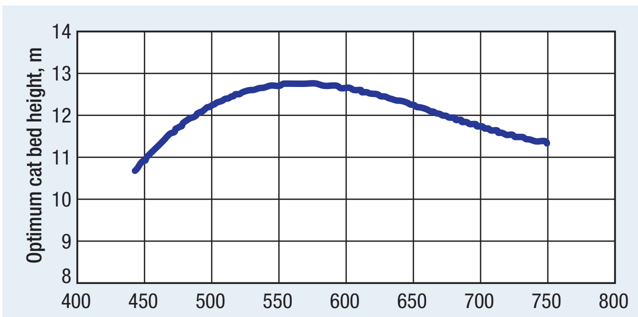

FIG. 7 Optimum catalyst-bed height vs. single reactor capacity.

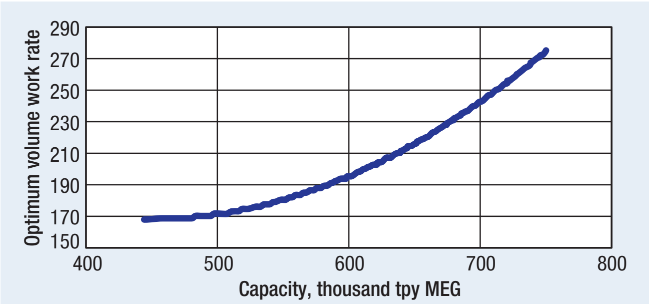

FIG. 8 Optimum work rate vs. single rector capacity.

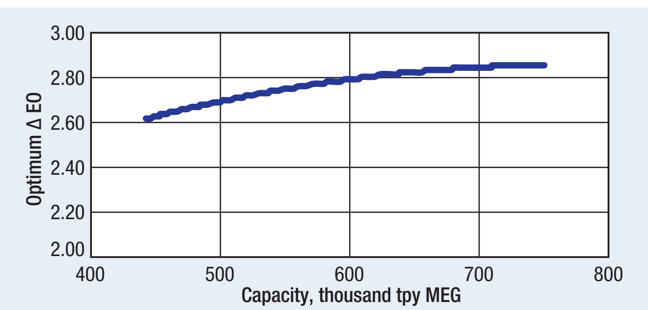

FIG. 9 Optimum Δ EO vs. reactor capacity.

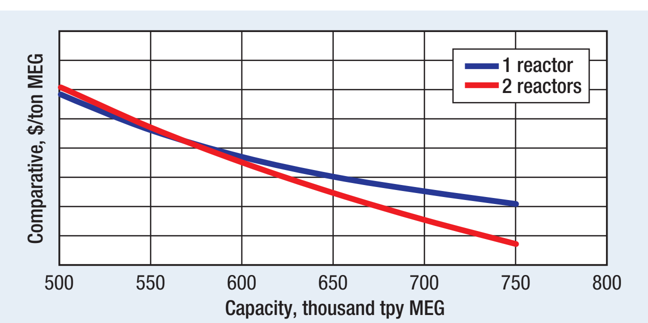

FIG. 10 MEG cost vs. production.

a reactor contains about 24 m^3 of catalyst for each meter of catalyst-bed height. For example, if the catalyst-bed height were 10 m, the empty reactor would weigh more than 1,000 metric tons and would contain well over 200 m^3 of catalyst. At a volumetric work rate of 200, this reactor would produce about 50 tph of EO or 400,000 tpy (400 Mtpy) of EO based on 8,000 operating hours per year. In most cases, large sites are integrated EO/EG processing units that convert most or all of the EO to EG. Plant capacity is usually expressed in terms of monoethylene glycol (MEG), which is produced by the hydrolysis of EO. In a conventional noncatalytic hydrolysis reactor, the EO requirement is approximately 0.8 tons of EO per ton of MEG. So 400 Mtpy of EO produced in a large EO reactor is typically equivalent to 500 Mtpy of MEG. Reactor and plant capacity and costs will later be described in terms of MEG equivalent. Fig. 2 shows a typical large modern EO reactor.

GENERAL OPTIMIZATION

The first step is determining the catalyst life. The economic catalyst life minimizes the sum of raw materials and catalyst costs and depends on the rate of selectivity loss over time. Actual catalyst characteristics depend on catalyst type; specifics are considered proprietary but general trends are reasonably universal. Selectivity and activity decline are both accelerated by increasing conversion and productivity (ΔEO and work rate). Although the catalyst life is occasionally limited by operating temperature, it is more common to replace catalyst at or near its economic service life.

For a new project, a target catalyst life might also be defined as the result of a preferred shutdown schedule based on other considerations such as required inspection intervals or coordination with other plants within a complex. Targeting a specific catalyst life for such reasons generally leads to a constrained optimum. The catalyst life for a given set of operating conditions sets the average cost of raw materials and catalyst per unit of product. Catalyst-bed height can also be varied, and this essentially involves trading capital vs. power costs.

The capital cost of adding reactor volume depends on whether this is accomplished by increasing bed height (tube length) or by adding more tubes. From a capital viewpoint, it is more economical to increase bed height than tube count. However, the reactor pressure drop is proportional to the cube of the bed height for fixed-reactor volume and has a strong effect on recycle compressor power. Catalyst particle size and shape also affect pressure drop. The generally used shape has evolved from spherical to the present hollow cylinder and continues to be subject to further investigation. For simplicity, we will assume that the catalyst characteristics are already fixed and that they focus only on the process variables that can be optimized. Fig. 3 shows an example of the effect of catalyst-bed height on production costs for one set of conditions.

The general trend when reactor size is not constrained, as illustrated in Fig. 4, is that the economic bed height tends to decrease with an increasing work rate (smaller reactor) and increase with an increasing ΔEO (lower gas flow). Each set of operating conditions (ΔEO and work rate) defines the cycle gas flow. Reactor pressure drop and, therefore, recycle compressor power are then defined by the catalyst-bed height. All other operating and capital costs associated with cycle gas flow can be calculated.

Finally, the cost of production can be calculated over a range of conditions for a particular set of raw material and utilities costs. With four variables being adjusted (ΔEO, work rate, catalyst life and bed height), the resulting cost curve is relatively flat and a modest deviation in any of the variables has only a small effect on the cost of production. Fig. 5 shows production cost vs. ΔEO with the three other variables optimized, indicating a minimum cost at approximately 2.6 ΔEO. The corresponding volumetric work rate is 170, and the single-reactor capacity, for this example, is estimated at 450 Mtpy of MEG at the typical diameter limit. The optimum catalyst-bed height for this case approximates a straight line when plotted against ΔEO and the trend line, as shown in Fig. 6.

CONSTRAINED OPTIMIZATION REACTOR SIZE LIMITS

Economies of scale mean that the unit cost of production decreases as plant capacity increases. As limiting reactor diameter is reached, there is an additional constraint and a choice between either adding a reactor or increasing the individual reactor capacity. Increasing the number of reactors adds capital cost and plot area, while increasing the individual reactor capacity forces an increase in work rate and/or bed height that increases operating costs.

The economic optimum bed height is affected by power cost, as expensive power tends to reduce the economic bed height. In one example, using 12¢/kWh, raising the reactor capacity above 450 Mtpy is most economically achieved by increasing bed height to accommodate 550 Mtpy–600 Mtpy, at which point the optimum bed height has reached a maximum and more capacity increase is achieved by using work rate. Further, the economic ΔEO tends to increase with capacity, as this limits the increase in cycle gas flow. These trends are illustrated in Figs. 7–9.

Increasing individual reactor capacity, therefore, saves capital but increases operating costs. This is worthwhile up to a breakeven point, beyond which it becomes more economical to add a reactor. In this example, the breakeven point occurs at an individual reactor capacity of 570 Mtpy of MEG, as illustrated in Fig. 10.

The process engineer has several variables to adjust to minimize cost of production for a specific case. ΔEO, work rate, catalyst-bed height and life are the main limiting size factors for EO reactors. As illustrated here, two EO reactors would be the economic choice for processing capacities over 570 Mtpy. **HP**

NOTE

Scientific Design currently offers a single EO reactor for MEG capacities up to 700 Mtpy, constrained by capabilities of the approved fabricators and the usual logistical concerns.

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