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Economic Selection of the Components of an Air Separation Process

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This paper presents an availability analysis of one type of oxygen production cycle centering around separation of oxygen and nitrogen in a fractionating tower. The plant is driven by work inputs to compressors and blowers. The analysis shows the irreversible entropy production in the various units and, in turn, the added work inputs required as a consequence thereof. Furthermore, a comparison is made with an ideal process of the same type, wherein all irreversibilities are reduced to the minimum possible, subject to the constraints imposed by (a) the use of a tower, and (b) the properties of the flowing streams.

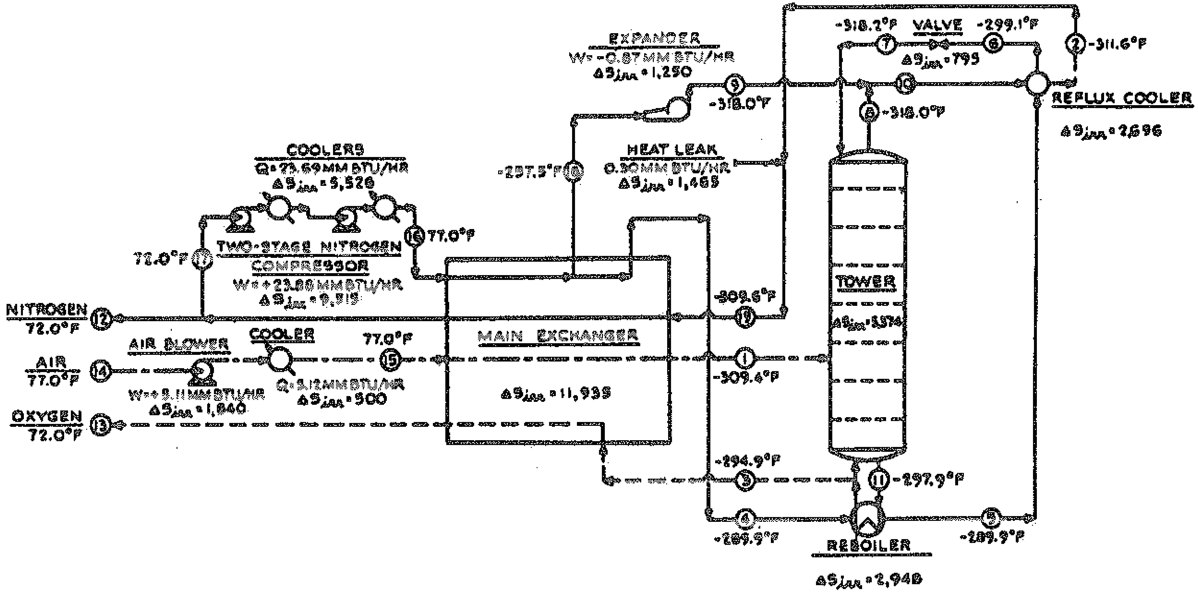
In addition, it is shown how the total cost of oxygen production varies with the irreversibility in the main heat exchanger. By relating this cost to the design parameters of the heat exchanger (pressure drop and warm-end temperature difference), the parameter values that minimize the total cost are determined. Thus, with the main heat exchanger as an example, it is shown how availability methods can be employed in design for "optimal" selection of components.

This work was first presented in an unpublished lecture given at MIT in 1949, along with similar analyses of several other chemical processes.

Oxygen Separation Process

This paper is concerned with the amount of work needed to separate air into an oxygen-rich fraction containing 90 mole percent oxygen and a nitrogen-rich fraction containing 99 mole percent nitrogen. The minimum work required in a thermodynamically reversible process conducted in an environment at one atmosphere and 77°F with feed and product gases at these conditions is 421 Btu per pound mole of air fed.

The amount of work used in a practical oxygen separation process is much greater than this minimum because of irreversibility. Figure 1 illustrates the process flow for one type of oxygen production cycle. The process quantities refer to separation of



STREAM	①	②	③	④	⑤	⑥	⑦	⑧	⑨	⑩	⑪	⑫	⑬	⑭	⑮	⑯	⑰	⑱	⑲
	COLUMN FEED	HEATED OVERHD.	COLUMN BOTTOMS	REMADE NITROGEN	REFLUX	COOLED REFLUX	EXPANDED REFLUX	COLUMN OVERHD.	EXPANDED NITROGEN	⑧ + ⑨	LIQUID FROM TOWER	PRODUCT NITROGEN	PRODUCT OXYGEN	INITIAL FEED	FEED TO EXCH'R	COMPR'SD NITROGEN	RECYCLE NITROGEN	FEED TO EXPANDER	OVERHD. + HEAT LEAK
% NITROGEN	79	99	10	99	99	99	99	99	99	99	10	99	10	79	79	99	99	99	99
MOLS/HR	10,000	16,727	2247	6638	6638	6638	6638	14,391	2336	16,727	4813	7753	2247	10,000	10,000	8974	8974	2336	16,727
PRESSURE, PSIA	18.7	16.7	18.7	79	74	73	17.7	17.7	17.7	17.7	19.7	14.7	16.7	14.7	20.7	77	14.7	76	15.7
ENTHALPY, BTU/M	998	987	1090	1026	-1092	-1183	-1183	935	935	935	-1772	3692	3687	3727	3726	3716	3692	1306	1005
ENTROPY, BTU/M/°R	37.932	36.704	40.439	34.115	21.861	21.055	21.175	36.023	36.023	36.023	22.926	45.843	49.004	46.480	46.791	47.997	45.843	35.488	36.625

Figure 1. Practical oxygen process

10,000 pound moles of air per hour into a waste fraction containing 99 percent nitrogen and a product fraction containing 90 percent oxygen. The plant produces 380 tons of oxygen per day with a recovery of 96 percent.

Air at 77°F is compressed by a blower to a pressure of 20.7 psi, which is sufficient to force the air through the exchangers and towers of the separation plant. In the main exchanger, the air is cooled by means of the outgoing product oxygen and nitrogen, and emerges at its dewpoint at -309.4°F. In the tower, the air is fractionated into the nitrogen overhead and oxygen bottom streams. These streams are returned through the main exchanger and discharged from the plant at 72°F. The nitrogen is also used to subcool tower reflux and is finally discharged at atmospheric pressure. The oxygen, which emerges from the tower at a somewhat higher pressure than the nitrogen, is discharged from the plant at 16.7 psi.

To reboil the tower, to provide reflux, and to satisfy the refrigeration requirements of the plant, an auxiliary nitrogen stream amounting to 90 percent of the air fed to the plant is compressed in a two-stage compressor to 77 psi and is also cooled to its dewpoint of -289.9°F in the main exchanger. This nitrogen is liquefied in the reboiler, where its latent heat reboils the tower, is subcooled against outgoing product nitrogen in the reflux cooler, and finally is flashed through a valve into the top of the tower where it provides reflux at -318.2°F. In the tower, the nitrogen reflux stream is vaporized into the product nitrogen and flows with it through the nitrogen pass of the main exchanger where it gives up its heat to the incoming compressed nitrogen. To compensate for the heat leak to the plant and the enthalpy difference between the outgoing nitrogen and oxygen streams and incoming air, a portion of the compressed nitrogen is made to do work in the expander in order to lower its temperature from -275.5°F to -318.1°F.

In setting up the conditions for this process, the following assumptions have been made: (a) pressure drops of 2 psi through the main exchanger and 1 psi through the other exchangers and from top to bottom of the tower; (b) a temperature difference of 5°F for each of the gas-to-gas exchangers; (c) a net heat upflow in the tower equal to 3.5 percent greater than the minimum flow needed to carry out the separation with an infinite number of plates; (d) a heat leak of 30 Btu/mole of incoming air or 300,000 Btu/hr; and, (e) efficiencies for the compressors, the blower, and the expander equal to 75 percent. Thus, a total of 29 x 10^6 Btu/hr of work is expended in the nitrogen compressors and the blower, and 0.87 x 10^6 Btu/hr of work is recovered in the expander. The net work input is 28 x 10^6 Btu/hr or 10.8 kWh per 1000 ft^3 of oxygen. The theoretical work of carrying out this separation, evaluated from the change in enthalpy and entropy of the feed and products, is only 4.6 million Btu/hr or 1.7 kWh per 1000 ft^3 of oxygen.

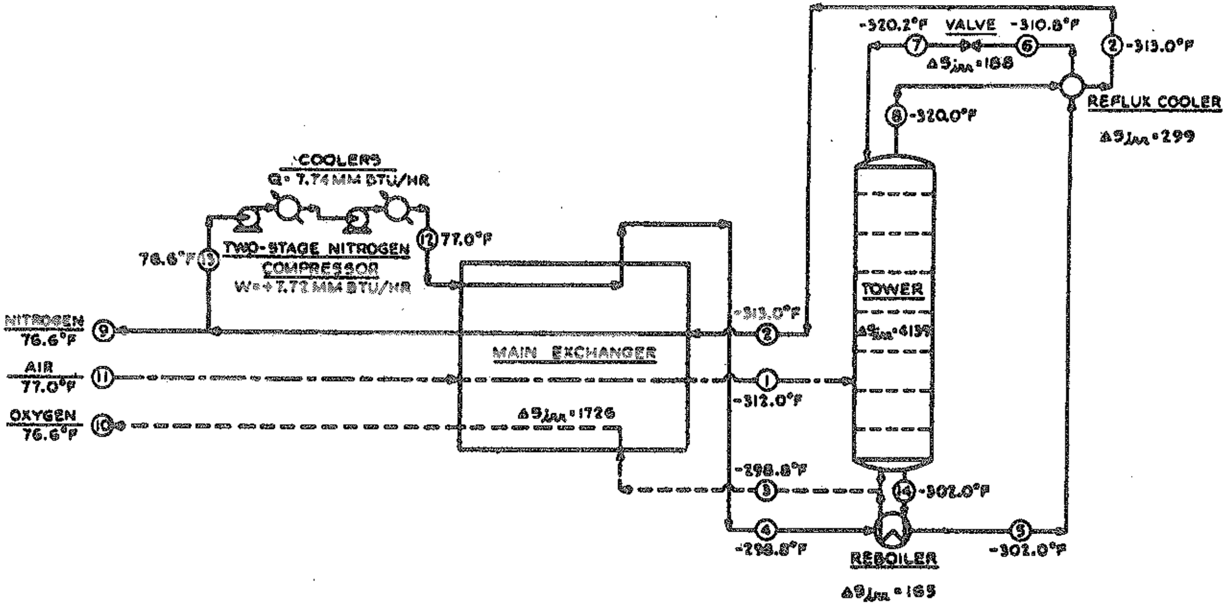
The process inefficiencies that are responsible for the increased work expenditure can be identified by evaluating the entropy production of each element of the process, namely, by evaluating the irreversibilities. The largest portions of this entropy production, shown in Figure 1, are the main exchanger, the nitrogen compressors, the nitrogen coolers, and the tower.

To improve the effectiveness of the process, it is necessary to determine the extent to which the entropy production is characteristic of the process and the extent to which it is a consequence of equipment inefficiencies that could be reduced by increased capital expenditure for either larger or more effective equipment. To do this, we consider an idealized process design for a plant in which the exchangers are so large that the pressure drop is zero and the temperature difference is the minimum consistent with process heat balance. In addition, we assume that the heat leak may be made negligible by elaborate insulation, that only the minimum heat needed to reboil the tower will be used, and that the nitrogen compressor may be made 100 percent efficient and isothermal.

Figure 2 shows the flow sheet for such an idealized process. The work input is reduced from 28 x 10^6 Btu/hr to 7.7 x 10^6 Btu/hr, and the irreversible entropy production has been made practically negligible in all pieces of equipment except the tower. The entropy production here is still high because of the fact that fractionation occurs away from equilibrium conditions at all points of the tower except at the feed point.

Table I compares the work input and rate of entropy production in the practical process with corresponding quantities in the ideal process. This brings out the fact that the single place where the greatest reduction in power can be effected is in the main exchanger, and also shows the relative importance of the individual pieces of equipment as contributors of the total work input. It also shows that the total work input is the sum of (a) the theoretical work, and (b) the work equivalent of irreversible entropy production, given by the product of ΔS_{irr} and ambient temperature T_0.

To illustrate how the entropy production may be used to select approximate optimum design conditions for individual units of process equipment, we calculate the effect of varying the parameters of the main exchanger upon the total cost of the exchanger, the compressors, and the power required to run the plant for 5 years. The results are given in Figure 3. The figure illustrates the cost balance on this exchanger. The pressure drop and the warm-end temperature difference for which the exchanger is designed contribute to the cost of the plant in two principal ways. First, the lower the pressure drop and temperature difference, the larger the exchanger and the greater its cost. Second, the lower the pressure drop and warm-end temperature difference, the lower the irreversible rate of entropy production in the main exchanger, the smaller the compressors needed for the plant and the lower the



STREAM	① COLUMN FEED	② HEATED OVERHD.	③ COLUMN BOTTOMS	④ RECYCLE NITROGEN	⑤ REFLUX	⑥ COOLED REFLUX	⑦ EXPANDED REFLUX	⑧ COLUMN OVERHD.	⑨ PRODUCT NITROGEN	⑩ PRODUCT OXYGEN	⑪ INITIAL FEED	⑫ COMPRESSED NITROGEN	⑬ RECYCLE NITROGEN	⑭ LIQUID FROM TOWER
% NITROGEN	79	99	10	99	99	99	99	99	99	10	79	99	99	10
MOLS/HR	10,000	13,833	2247	6080	6080	6080	6080	13,833	7753	2247	10,000	6080	6080	4690
PRESSURE, PSIA	14.7	14.7	14.7	48	48	48	14.7	14.7	14.7	14.7	14.7	48	14.7	14.7
ENTHALPY, BTU/M	982	978	1068	1009	-1226	-1349	-1349	924	3728	3721	3727	3731	3728	-1840
ENTROPY, BTU/M/°R	38.919	36.699	40.776	34.769	20.808	20.004	20.025	36.924	45.916	49.332	47.474	42.544	45.916	22.565

Figure 2. Ideal oxygen process

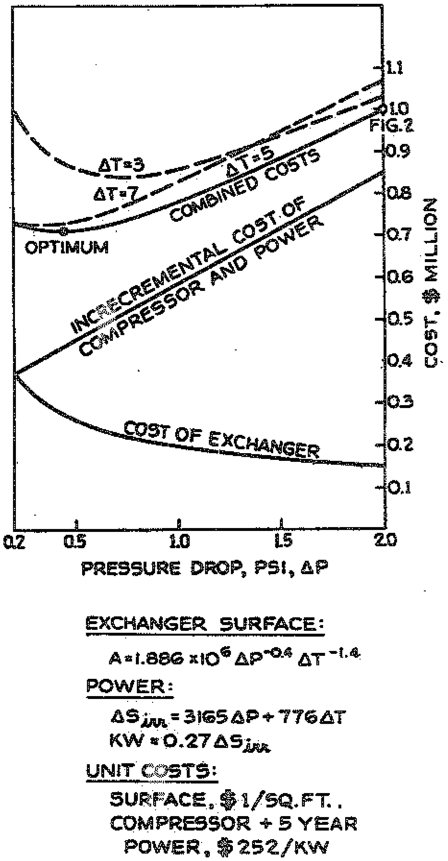


Figure 3. Cost balance on main exchanger

Table I.
Entropy Production Breakdown in Oxygen Processes
(Feed Rate: 10,000 Moles Air/hr)

PROCESS	IDEAL		PRACTICAL	
	Work Input, 10^6 Btu/hr	$T_0\Delta S_{irr}$, 10^6 Btu/hr	Work Input, 10^6 Btu/hr	$T_0\Delta S_{irr}$, 10^6 Btu/hr
Air Blower		–	5.11	0.988
Air Cooler		–		0.268
N_2 Compressor	7.72	–	23.88	5.109
N_2 Coolers		–		2.967
Main Exchanger		0.927		6.409
Expander		–	–0.87	0.671
Tower		2.222		2.886
Reboiler		0.089		1.583
Reflux Cooler		0.161		1.448
Valve		0.101		0.427
Heat Leak		–		0.787
TOTAL	7.72	3.50	28.12	23.54
$T_0\Delta S_{irr}$, 10^6 Btu/hr		3.50		23.54
Theoretical Work=$\Delta H-T_0\Delta S$		4.21		4.61*
Predicted Work, 10^6 Btu/hr		7.71		28.15

*The 4.61 consists of the 4.21 units of ideal theoretical work plus 0.40 unit needed because the N_2 and O_2 leave at $T = 72^\circ F < 76.6^\circ F$; i.e., the 0.40 unit is the theoretical work done by cooling from 76.6°F to 72°F.

cost of power supplied to the plant. The equation given in the figure indicates how the exchanger surface (A) will vary with pressure drop and temperature difference. The assumed cost for exchanger surface is $\$1/\mathrm{ft}^2$.

The effect of pressure drop and temperature difference in increasing the rate of entropy production from the ideal case to the practical case is represented by the equation given in Figure 3. Were the availability efficiency 100 percent for the compressors, their coolers, and motors, then the additional power required because of irreversibility in the heat exchanger would simply be $\mathrm{kW} = T_0\Delta S_{irr} = 537^\circ\mathrm{R}\,[\Delta S_{irr}(\mathrm{Btu}/^\circ\mathrm{R\text{-}hr})]\,[\mathrm{Btu}/3412\ \mathrm{kWh}] = 0.157\,[\Delta S_{irr}(\mathrm{Btu}/^\circ\mathrm{R\text{-}hr})]$. Assuming a composite availability efficiency of 58 percent for compressor and auxiliaries, the actual power requirement would be $\mathrm{kW} = 0.27\ \Delta S_{irr}$. It is assumed that the initial cost of the compressors is increased by \$50 for every kilowatt increase in power input and that power is to be charged at the rate of 0.6¢/kWh for a period of 5 years. Assuming an interest rate of 5.5 percent and a 90-percent plant capacity factor, the present worth of the total cost of power is \$252 per kilowatt or \$68 per unit rate of entropy production. (As this economic analysis was made in 1949, costs and interest rate are no longer representative.

Electricity has increased by a factor of about 4, and capital equipment and interest rate by a factor of about 2.)

The solid curves in Figure 3 show the effect of pressure drop upon cost for a warm-end temperature difference of 5°F. The opposing tendencies of entropy production and exchange surface are indicated. Total cost is minimized at a pressure drop of about 0.5 psi rather than the 2 psi used in the practical process. Similar curves for the total cost at temperature difference 3°F and 7°F indicate that a warm-end temperature of 5°F is close to the optimum.

Thus, by consideration of the rate of entropy production in the main exchanger, the optimum pressure drop and temperature difference across this exchanger may be estimated without the necessity for a complete system design at each combination of conditions.

Closure

In a real application, dating back to 1949, this paper concisely illustrates the theme of the symposium: the use of availability (a) to calculate "work penalties" for the various irreversibilities in a process, and (b) to determine the total costs attributable to the irreversibilities and the use of these costs for "optimal" design. Another important result is the determination of the minimum work for a process with unavoidable irreversibilities, such as those associated with the fractionating tower here.

Editor's Note:

This paper illustrates the use of entropy balances in Second-Law analysis. See the editor's note at the end of the article by Cremer, in this volume.

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