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PROCESS STUDY

GAS PURIFICATION FACILITIES AT PUREX

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PROCESS STUDY

GAS PURIFICATION FACILITIES AT PUREX

By

L. R. Michels

and

J. M. Gerhart

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December 31, 1958

Extraction Design and Development
 Facilities Engineering Operation
 CHEMICAL PROCESSING DEPARTMENT
 General Electric Company-Hanford Atomic Products Operation

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PROCESS STUDY
GAS PURIFICATION FACILITIES AT PUREX

INTRODUCTION

This report provides a summary of the results of a process study, requested by the Atomic Energy Commission in Reference 1, on the recovery of krypton and xenon from irradiated uranium at the Hanford Purex Plant. This request was prompted by original Commission forecasts, Reference 2, of the expanded requirements for Krypton-85 for commercial phosphorescent signal lights and markers and for xenon isotopes of low neutron cross-section for use in liquid xenon scintillation counters, in connection with D.M.A., government and university-sponsored work. Reference 1 requested that both Hanford and Savannah River submit order-of magnitude cost estimates for recovery facilities at the respective sites for three separate design cases. The cost information developed, along with market survey information obtained through the A. D. Little Company and Department of Defense market surveys, would serve as the basis for scheduling of the Hanford and Savannah River participation in the Commission's overall fission rare gas recovery program.

The three design cases defined by the Commission in Reference 1 were stated as follows:

"Case I

<u>Hanford and Savannah River</u>	- Bulk recovery of krypton-xenon mixture.
<u>Oak Ridge National Laboratory</u>	- Partitioning of krypton and xenon, purification and packaging of the isolated products.
<u>Mound Laboratory</u>	- Isotopic separation of the xenon isotopes.

Case II

<u>Hanford and Savannah River</u>	- Separate recovery of krypton and xenon.
<u>Oak Ridge National Laboratory</u>	- Final purification and packaging of the isolated products.
<u>Mound Laboratory</u>	- Isotopic separation of the xenon isotopes.

Case III

- Same as Case I and II except Savannah River and Hanford should consider the isotopic separation of the xenon isotopes."

As applied to Hanford, Case I entails bulk recovery of krypton and xenon. Case II entails Case I plus the subsequent partitioning of krypton and xenon. Case III entails Case II plus the subsequent isotopic rearrangement of the Case II xenon products.

1. HAN-68425, "Krypton-Xenon Recovery," by A. T. Gifford, Hanford Operations Office, AEC, dated February 27, 1958.
2. HAN-67394-1, "Krypton and Xenon-Uses, Requirements and Availability - FY-1959 Budget," by Hanford Operations Office, AEC, (Undated).

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The major portions of the Case I and II process studies were performed by Air Products, Incorporated, of Trexlertown, Pennsylvania, (References 3 and 4), subcontractor to the General Electric Company, under Design, Development, and Research Contract DDR-45. Specifications prepared by HAPO (Reference 5), which were incorporated into Contract DDR-45, served as the design basis for the Air Products' study. Case I and II architectural work was performed by the General Electric Company, based upon equipment arrangement studies developed jointly by the General Electric Company and Air Products, Incorporated.

The Case III xenon isotopic separation study was performed by the General Electric Company following initial contacts with Mound and Oak Ridge National Laboratory. The Case III design reported herewith is based upon the separation of a Case II xenon product of 99 per cent purity (balance krypton, with traces of O_2 and N_2) into a "heavy" product, largely the $Xe-134$ and $Xe-136$ isotopes, with a composite neutron cross-section of less than 1 barn and a raffinate fraction containing the lighter xenon isotopes with about 2 per cent krypton contaminant. Recently, interest was expressed (Reference 6) in the separate recovery of a "light" xenon product, virtually free of krypton, from the above raffinate. This "light" product presumably would find use in neutron scintillation devices. An addendum to the Case III study will be submitted at a later date when firm specifications, requested in Reference 6, become available on the "light" fraction and when additional thermal diffusion process studies have been completed.

SUMMARY AND DISCUSSION

Cases I and II

At the Hanford Purex Plant, the dissolution of uranium, irradiated at 8 MW/T and 650 MWD/T, at a 3.2 Purex capacity factor (582 Tons U/Mo) results in the liberation of about 4,600 grams per year of Krypton-85(1,870,000 curies) mixed with 90,000 grams of inert krypton and about 700,000 grams per year of mixed xenon isotopes(124,000 liters xenon gas at 14.7 psia and 60°F).

The various steps in the proposed process for the recovery of fission krypton and xenon from Purex dissolver off-gas, described in detail in References 3 and 4 are entirely feasible and well proven. The process employs standard low temperature distillation techniques, which have been

3. HW-API-1, "Report on Process Design Study for a Rare Gases Recovery and Purification Facility," by Air Products, Incorporated, dated November 1, 1958.
4. HW-API-2, "Supplementary Information on Process Design Study for Gas Recovery and Purification Facility," by Air Products, Incorporated, dated December 12, 1958.
5. HWS-6402, "Process Development Specifications for Rare Gas Recovery Facility - Phase I Contract," by L. R. Michels, dated May 29, 1958.
6. HW-58323, Letter, J. H. Warren to A. T. Gifford, "HAPO Gas Purification Study - Revised Schedule," dated December 2, 1958.

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used commercially both in this country and abroad for many years (Reference 7). All preliminary operations in which dissolver off-gas is treated prior to distillation for the removal of oxides of nitrogen and traces of acetylene are well proven by commercial practice and have been further checked in the Air Products' Laboratory for the specific gas concentrations anticipated at Purex. Further, recent shop testing of the new rare gas recovery unit constructed for Arco at the Murray Hill Laboratories of the Air Reduction Company, using commercial krypton-xenon feed, provide added confidence in the Hanford design (Reference 8).

The Case I process, summarized in this report and References 3 and 4, will provide a gas product containing about 99.85 volume per cent mixed krypton and xenon. The efficiency of rare gas recovery would be in excess of 90 per cent. On this basis, a minimum recovery of about 1,680,000 curies of Kr-85 and 112,000 liters xenon per year, as a Case I mixed rare gas, would be possible at Purex. All processing steps downstream of the batch dissolver operations would be performed in continuous equipment.

Shipment from Hanford to ORNL of the Case I mixed rare gas, containing 99.85 per cent combined krypton and xenon for subsequent partitioning into the individual Case II gases does not appear attractive. Incremental Hanford capital and operating costs for the partitioning step above Case I are relatively small, since the partitioning could be performed readily in relatively small equipment by low-temperature batch distillation methods within the Case I equipment cold box, using refrigeration available from the Case I equipment. Case I economics are discussed following the presentation of Case II costs in this section of the report.

In Case II, the mixed krypton-xenon product of Case I would be partitioned by low-temperature batch distillation techniques into separate krypton and xenon streams, with essentially complete recovery of the Case I components. Under irradiation conditions listed above, about 1,680,000 curies per year of a Case II krypton product containing about 5 volume per cent Krypton-85, 94 per cent inert krypton, 1 per cent xenon and less than 50 ppm of combined other impurities would be recovered at Purex. About 112,000 liters per year of a Case II xenon product of the following approximate volume composition would be recovered: Xe-131-11.9%, 132-17.5%, 134-27.3%, 136-42.2%, 128-129-130-0.1% total, Kr-1.0% total, and less than 50 ppm of other impurities.

The total Case II capital cost for equipment and facilities required to process the entire quantity of rare gas liberated from the Purex dissolver systems, including design and a 20 per cent contingency is about \$2,130,000.

7. "The Separation of Gases", by M. Ruhemann, Oxford Clarendon Press, 1949, pages 228 to 236.

8. HAN-70538-1, "Final Report Airco No. C-214-6, Development of Dissolver Off-Gas Separation Unit," by Air Reduction Company, dated October 1, 1958.

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These costs are based on the procurement of specialized gas processing equipment and instrumentation on a "design and fabricate" contract with the necessary building, process, service, and instrumentation tie-ins to be provided under a separate lump-sum contract. From a scheduling standpoint, beneficial use of Case II facilities could be obtained within 26 months following project authorization on a normal construction schedule.

As shown in Table I, the total Case II yearly operating expense, including depreciation on a 10-year basis, is about \$385,000. Half of this expense has been pro-rated to krypton and half to xenon, giving respective unit costs for krypton of about 11.5 cents per curie and for xenon about \$1.72 per liter, F.O.B., Hanford.

TABLE I
Case II - Summarized Economics
Basis: Facilities to Process Entire Purex Dissolver Off-Gas

<u>Case II Product</u>	<u>Krypton</u>		<u>Xenon</u>	
	<u>Isotope</u>	<u>Vol. Per Cent</u>	<u>Isotope</u>	<u>Vol. Per Cent</u>
	Kr-85	5.0	Xe-128	0.1
	Kr-Stable	94.0	Xe-129	
	Xe-Total	1.0	Xe-130	
			Xe-131	11.9
			Xe-132	17.5
			Xe-134	27.3
			Xe-136	42.2
			Kr-Total	1.0
Total		100.0	Total	100.0

Recovery, Annual Basis

Krypton, Curies

1,680,000

Xenon, Total Liters

112,000

Shipping Containers

Total Number Per Year(1)

52

52

Gross Weight Per Container, Pounds

3,300

1,800

Gas Pressure at 125°F, PSIA

130

620

**Total Capital Cost of all Facilities
required to recover Case II Products,
including Case I bulk recovery but
excluding shipping containers**

\$2,130,000(2)

- (1). Arbitrary shipping basis: 1 cylinder of Krypton and 1 cylinder of Xenon per week. Increased pressure would result in the shipment of fewer cylinders.
- (2). Includes cost of acetylene removal equipment.

Yearly Operating Costs(3)

Direct Labor and Supervision	33,000
Materials and Supplies	28,000
General and Process Utilities	20,000
General and Plant Maintenance	25,000
Radiation Monitoring	1,000
Analytical Control	12,000
Engineering Assistance & Process Tech.	15,000
→ Plant Overheads	<u>38,000</u>
Semi-Total	\$172,000
Depreciation, 10-year basis	<u>213,000</u>
Total Expense, Both Products	\$385,000

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	<u>Krypton</u>	<u>Xenon</u>
Total Pro-Rated Expense	\$ 192,500	\$ 192,500
Total Krypton Unit Cost, per curie	\$ 0.115	
Total Xenon Unit Cost, per liter		\$ 1.72

(3)
Exclusive of shipping costs.

The above Case II costs are valid only within the framework of the above bases and assumptions. For example, on a 20-year depreciation basis, the unit costs would be reduced to \$.083 per curie of Krypton-85 and \$ 1.24 per liter of xenon.

Various other alternates should also be considered in evaluating Hanford's role in rare gas processing. Should the Commission be interested in permanent facilities for the recovery of only 500,000 curies of Krypton-85 and 33,000 liters of Xenon per year or less, considerable capital and operating cost savings would be obtained, although somewhat higher unit costs would result. Detailed estimates of these alternates have not been developed because of uncertainties in cost scaling factors. Also, equipment designed for processing of less than the total quantity of Purex dissolver off-gas would not be readily expandable at a future date to take advantage of the large rare gas recovery potentialities of Purex. Relatively high capital costs would be required at a future date for the subsequent expansion.

A Case II xenon product containing less than 1 per cent total krypton might prove advantageous from the standpoint of meeting rigid krypton specifications on a possible "light" Case III xenon product. Modification of the Case II batch distillation equipment and method of operation, delineated in Reference 3, is considered feasible to provide a xenon fraction containing in excess of 99.9 per cent xenon, at little added cost.

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The added total Hanford capital costs above Case I for the rare gas partitioning step are only about \$110,000. On this basis, if a Case I gas product only were recovered at Hanford, the total capital cost of the Case I facilities would be about \$2,020,000. As with Case II, there is possibility that the Case I capital cost can be reduced to \$1,920,000 by deletion of primary acetylene oxidation equipment(Reference 9). Case I direct operating costs would be essentially the same as for Case II. Construction schedules for Case I and II would be the same. On this basis, it is recommended that the Case II partitioning equipment be provided at Hanford as part of the overall rare gas recovery facility.

The need for special Case I facilities to remove acetylene from the gas stream prior to distillation has not been proven conclusively. (The acetylene problem is discussed in detail in a latter section of this report). For reasons of safety, a conservative approach to the acetylene problem dictated the inclusion of complete acetylene removal equipment, consisting of a reactor, heat exchangers and controls. Should further Purex sampling tests and other studies prove conclusively that acetylene removal equipment is unnecessary, the Case I and II costs shown above would be reduced as indicated.

Case III

As mentioned previously, the Hanford Case III process study had as its objective the separation of the Case II xenon product into a heavy molecular weight fraction (largely Xe-136 and -134), with a total krypton content of 10^{-14} per cent or less and a composite neutron cross-section of less than 1 barn. As a result of the above split, a raffinate xenon product, containing mainly the lighter xenon isotopes along with 2 volume per cent total krypton, would be formed. The raffinate product would either be shipped to ORNL or stored.

An early review of the available processes for achieving the desired degree of separation at a new Hanford facility indicated that only the thermal diffusion approach was feasible from an overall economic standpoint. Gaseous diffusion, simple distillation and gas centrifugation appeared too costly for consideration at a new Hanford facility. (Handwritten: 10.1.59 sqw/)

The design of the Case III thermal diffusion facilities is described in a later section of this report. Briefly, a total of 60 thermal diffusion columns, installed in pairs, would be required. Each column would be about 12-inches I.D. by 24 feet high and would require an estimated power consumption of about 40 KW. The Case III facility would be located immediately adjacent to the Case II facility, to the south and east of Purex Building.

The total additional capital cost required for the Case III facility (over Case II) is about \$3,700,000, including design, a 20 per cent contingency and all project additives. The estimated additional operating expense for the thermal diffusion plant, over Case II, would be about \$744,000 per year(Table II), including depreciation on a 10-year basis. If all thermal diffusion costs were charged to the Case III product, the added costs, over Case II, for the Case III product would be about \$13.25 per liter.

9. Letter, W.C. Schmidt to L.R. Michels, "Acetylene Content of Purex Off-Gas", dated December 16, 1958.

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TABLE

Case III - Summarized Economics - Additional Costs Over Case II
Basis: Facilities to Process Entire Purex Case II Xenon

<u>Composition of Case III Streams, Expressed as Weight Per Cent</u>	<u>Feed (Case II Product)</u>	<u>Product</u>	<u>Raffinate</u>
<u>Isotope</u>			
Xe-128)			
-129)	0.1	>.01	~0.2
-130)			
-131	11.5	.05	22.7
-132	17.2	3.3	31.2
-134	27.3	24.3	30.0
-136	43.0	71.9	13.9
Kr-Total	<u>1.0(3)</u>	<u>10-14</u>	<u>2.0</u>
Total	100.0	100.0	100.0
Recovery, Annual Basis	112,000	56,000	56,000
Liters at 14.7 PSIA and 60°F	(feed)		
<u>Additional Total Capital Cost of Thermal Diffusion Facilities over Case II</u> (1),(2)		\$3,720,000	
<u>Additional Yearly Operating Costs</u> (1)			
Direct Labor and Supervision		67,000	
Materials and Supplies		57,000	
General and Process Utilities		89,000	
General and Plant Maintenance		25,000	
Radiation Monitoring		2,000	
Analytical Control		36,000	
Engineering Assistance and Process Tech.		20,000	
Plant Overheads		<u>76,000</u>	
	Semi-Total(round off)	\$372,000	
Depreciation, 10-year basis		<u>372,000</u>	
<u>Total Expense</u>		\$744,000	
<u>Total Thermal Diffusion Costs,</u>			
Per liter of Case III Product			
(56,000 liters)		\$ 13.25	
(or cost per gram of Case III Product)		2.21	

- (1) Exclusive of shipping
- (2) Excluding Research and Development Costs.
- (3) Calculations for Case III were based on weight per cent. The 1 weight per cent krypton in the feed provided a more stringent design condition than would be required for the 1 volume per cent material obtained from Case II.

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The design of the Case III 12-inch thermal diffusion columns was derived from theoretical relationships; information on columns of this size has not been reported in the literature. The present 1-inch diameter hot wire columns at Mound Laboratory, which have been used in xenon separation work, appeared to have too low a capacity for the quantity of xenon to be processed at Hanford.

A Research and Development Program, which would cost about \$275,000 and require 24 months for completion would be required. This program would have as its objective the optimization of the design, operation and control of the 12-inch column. A prototype column would be procured and tested to verify the xenon separation performance, the power consumption (which can vary markedly with surface emissivities), and to develop the optimum method of heating. Design and construction of the Case III Facility itself would require an additional 23 months following completion of the Research and Development program.

As a result of contacts between General Electric Company and Oak Ridge National Laboratory personnel, ORNL prepared a report, (Reference 10), which was an independent analysis of the Case III Hanford xenon separation problem. The ORNL report concurred with the General Electric Company thermal diffusion approach, in the event new capital facilities are required. Further, the design of the ORNL-suggested thermal diffusion plant would be similar to the design described in this report, requiring about the same amount of separative capacity as the Case III facilities. The design philosophy varied in that the General Electric Case III plant was a two-stage, square cascade, using identical columns, while the ORNL approach used a tapered cascade, the taper being achieved by varying the column diameter of a number of stages.

However, Reference 10 mentioned that units of the Oak Ridge gaseous diffusion cascade, which presently are not in use, might possibly be made available for xenon isotopic separation work. According to Reference 10, the entire quantity of Case II xenon produced at Hanford in one year could be processed through the Oak Ridge cascade in a 38 to 86 day period, depending upon the power level at which the cascade is operated, to give the above Case III products. Obviously, the feasibility and overall economics of using the Oak Ridge Gaseous Diffusion Plant cascade for the isotopic separation of the xenon produced at Hanford should be explored further.

In any event, it is recommended that research and development work be initiated at Hanford in the near future on large thermal diffusion columns. This work should have as its objective the correlation of thermal diffusion column operation against the theoretical relationships and the optimizing of column design from mechanical and heat transfer stand-points. Such a research and development program would permit a reduction of about 2 years in the overall construction schedule of a large thermal diffusion plant, not only for xenon separation, but for other separations as well.

10. KOA-459, "Separation of the Isotopes of Xenon", by S. Blumkin et. al., (Undated).

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Air Products, Incorporated, performed the Case I and II study on the specialized process equipment. The following Air Products' personnel contributed in the specific areas indicated:

G. Fedorko, Process Manager	- Overall process study coordination.
J. F. Tompkins, Process Engineer	- Process calculations.
T. J. Collins, Proposals Engineer	- Report preparation
A. C. Diehl, Chief Estimator	- Cost estimates on equipment.
B. M. Anderson, Chief Instrument Engineer	- Instrument requirements.
C. McKinley, Director, Research & Development	- R. & D. Information.
L. L. Phannensteil, Project Engineer	- Preliminary equipment layout.

Assistance of the following off-site personnel in the Case III study is acknowledged: G. R. Grove of Mound Laboratory, S. A. Levin, E. Von Halle, and D. A. Hayford of the Oak Ridge Gaseous Diffusion Plant.

The writers wish to thank E. T. Merrill for technical assistance, J. C. Willi and W. R. Felts for design assistance, W. C. Schmidt for Purex gas analysis data, R. W. McKee for calculations of rare gas production and L. B. Bradley for contract administration.

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APPENDIX

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GENERAL BACKGROUND

As a result of discussions held at Washington, D. C. during July, 1957 between Messrs. F. P. Baranowski of the A.E.C. Division of Production, J. H. Warren and V. R. Cooper of Hanford Atomic Products Operation, with regard to the recovery of fission krypton and xenon at Hanford, preliminary exploratory studies were initiated by the General Electric Company on fission rare gas recovery, in support of the Commissions' Civilian Application Program. During these discussions, it was indicated that the predicted market for the Krypton-85 isotope would be about 200,000 curies per year through 1959 and about 500,00 curies per year or greater thereafter. Keen interest was also expressed in the recovery of xenon, for use in scintillation counting.

General Electric Company exploratory studies (Reference 11) determined that the recovery of both fission krypton and xenon at the Hanford Purex Plant was technically feasible and the new facilities could be on stream by the end of calendar year, 1960. The preliminary estimated capital cost of the facilities, based on an adaptation of the ORNL "RAGS" process (Reference 12) was within the \$5,000,000 and \$10,000,000 range.

On December 11, 1957 HAPO was requested to initiate a development program aimed at a low-cost rare gas recovery process adapted to Purex (Reference 13), and to submit program plans as soon as possible. In Reference 13, the capacity criteria for the facilities were established to be 500,000 curies per year with a potential for increasing the rate to 1,000,000 curies per year. Preliminary specifications were also provided for the xenon products.

The HAPO development program on rare gas recovery, as outlined in Reference 14, called for engaging the services of a qualified gas processing equipment firm for the process design of the major separation and final purification equipment, with the design of the necessary buildings, services and tie-ins to be made by G.E. personnel. This course of action was deemed preferable to a complete process design by G.E. personnel in that 1) HAPO would capitalize on the wealth of information available mainly to firms engaged in low-temperature gas processing, 2) the process specified could be optimized better with respect to a balance between capital and operating costs, 3) a savings in time in the overall future construction program would be possible, and 4) an elaborate basic development program would not be required.

11. HW-52423, "Krypton-85 Recovery at HAPO", by J. H. Warren, dated September 10, 1957.
12. ORNL-1300, "Removal of Nitrous Oxide and Krypton from Uranium Dissolver Off-Gas", dated December 20, 1952.
13. HAN-67734, "Krypton-Xenon Recovery - Request for Program Study", by A. T. Gifford, Hanford Operations Office, AEC, dated December 11, 1957.
14. HW-54271, "HAPO Krypton-Xenon Recovery - Preliminary Development Program", by J. H. Warren, dated December 30, 1957.

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Following contract negotiations, Design, Development and Research Contract DDR-45(Reference 15) was executed with Air Products, Incorporated. This contract was approved by Air Products on July 18, 1958, and approved by the General Electric Company on July 28, 1958. The Contract called for a Phase I process study only for Cases I and II.

The results of the Air Products' study, summarized in documents HW-API-1 and 2(Reference 3 and 4) along with the supporting General Electric Company process design on the Case III isotopic separation, form the basis for this over-all study.

Case I and II Design Bases

The design maximum and minimum flow conditions(Reference 5) used for the Case I study were as follows:

	Maximum Flow	Minimum Flow
Flow Rate, SCFM	1,400	350
Delivery Pressure, PSIA	10	10
Delivery Temperature, Degrees F	105	70
Composition, Percent by Volume		
H ₂ O	5.0 ✓	5.0
NO	4.6 ✓	1.0
Total NO ₂ (Including N ₂ O ₄)	0.4	0.1
N ₂ O	0.1	0.1
A	0.9	0.9
O ₂	10.0	10.0
CO ₂	0.03	0.03
H ₂	0.6	0.6
He	5 ppm	5 ppm
Ne	15 ppm	15 ppm
I ₂	Trace	Trace
NH ₃	Trace	Trace
Hydrocarbons	Trace	Trace
Total Krypton	4.1 ppm	2.0 ppm
Total Xenon	21.4 ppm	6.2 ppm
N ₂	78.4	82.3

The maximum flow condition represents treated off-gas from two dissolvers in parallel at peak off-gas evolution rates.

Reference 5 also stipulated that the recovery facility should be capable of idling at thermal equilibrium conditions without inbleed of atmospheric air when dissolver feed gas was not available for processing, to avoid undue dilution of Krypton-85 with atmospheric krypton.

15. "Design, Development and Research Contract DDR-45 - Prime Contract W-31-109-ENG-52 - Contract for Purification System".

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Although Reference 5 proposed that Case I equipment should be capable of producing a mixed krypton-xenon product within the range of 0.1 to 5.0 volume percent total krypton plus xenon, the Air Products' study indicated at an early date that an essentially pure rare gas mixture could be obtained, following the primary low temperature stripping step, through the use of relatively small-sized simple equipment operated continuously, thus leading to lower shipping costs than for the shipment of a very dilute product. The 99.85 per cent product purity (mixed rare gases) specification for Case I was derived by Air Products as a result of investigations of the capabilities of the equipment involved.

Reference 5 specified that Case II should be capable of producing a krypton gas product containing a minimum of 99 volume per cent krypton (mixed krypton isotopes) and a xenon gas product containing a minimum of 99 volume per cent xenon (mixed xenon isotopes). The recovery of krypton and xenon from the feed gas was to be 90 per cent or greater for each element. Further, the equipment specified should represent an optimum balance between capital expenditures and operating costs without sacrifice of safety. Incremental capital for improved operability or reduced operating cost should have an economic payout period of no greater than two years.

Case I and II Process Description

Complete descriptions of the Case I and II processes appear in References 3 and 4. For convenience of the reader, however, highlights of the processes are summarized herewith.

Drawing No. SK-2-17998, along with the accompanying material and heat balance sheets, Tables III-A, III-B and III-C, comprise the Case I and II Process Flow Diagram. Feed gas for the Case I equipment is withdrawn from the 10-inch off-gas line discharging the T-XB Auxiliary Nitric Acid Absorber of the Purex 293-A Acid and Iodine Recovery Facility (Drawing No. SK-2-2959). Following the stripping of krypton and xenon from the gas in the Case I equipment, the lean off-gas is returned to the same 10-inch line in the 293-A Facility, downstream of the take-off point, for discharge through the existing jets to the Purex Stack. In this manner, the present Purex dissolver vacuum controls will be maintained. Purex dissolving operations within the 202-A Building can be carried out, as at present, irrespective of whether the Case I equipment is operating or not.

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The process description for the Case I and II processes, from Reference 3, is as follows:

"A - Feed Gas Preparation

Feed gas from the TX-B tower is delivered to the suction of the feed gas compressor (1.10) at 10 psia where it is compressed to 800 psia. The process cycle is designed to operate at the maximum flow conditions, but it will also operate satisfactorily at the minimum. When the flow rate is below the maximum, waste gas from the regenerators is recycled to the suction of the compressor to make up the difference.

"At 800 psia pressure, 98% of the water and oxides of nitrogen in the feed gas condense to nitric acid in the compressor and are removed in the entrainment separators following the feed gas compressor (1.10). Entrained lubricant oil is eliminated in an oil separator (37.10).

"The 800 psia gas stream next passes through an acetylene removal system. This consists of an acetylene reactor heat exchanger (2.61) at 500°F, where acetylene is oxidized to carbon dioxide and water. A reactor gas cooler (2.64), acetylene reactor heat exchanger (2.62), and reactor pre-heater (2.63) heat the gas to 500°F and cool it again to ambient temperature. If the feed gas contains less than 0.05 ppm acetylene, the removal system will be by-passed.

"Facilities for a three(3) minute feed gas hold-up is provided in the nitric oxide converter I (2.71) for the reaction $\text{NO} + \frac{1}{2} \text{O}_2 \longrightarrow \text{NO}_2$ to take place. This holdup is provided to reduce the nitric oxide concentration to 1 ppm. It is necessary, at this point in the process cycle, to completely convert nitric oxide to nitrogen dioxide as it cannot be removed to a level of 1 ppm by freeze-out in the regenerators.

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"B- Feed Gas Separation

Before the feed gas enters the feed gas regenerators (5.10), it is expanded to 45 psia, the operating pressure of the feed gas separation circuit. The feed gas regenerators (5.10) consist of two stainless steel vessels filled with quartz pebbles. Warm feed gas is cooled by downward flow over the cold pebbles while cold waste gas flows upward in the other bed to cool down the pebbles. Periodically, the flows will be switched from one vessel to the other.

"The carbon dioxide, water, nitrous oxide (N_2O) and nitrogen dioxide (NO_2) impurities in the feed gas will freeze-out on the regenerator pebbles when the warm feed gas flows downward through the bed. On switching, the cold waste gas vaporizes the impurities, removing them from the system.

"The feed gas leaving the feed gas regenerators (5.10) and entering the cold box contains less than 3 ppm CO_2 and N_2O , 1 ppm NO , and 0.01 ppm NO_2 and water.

"The feed gas is cooled in the regenerator preheater (5.41) to near its saturation temperature before entering the feed gas stripping column (7.10). In this column, the gas passes upward contacting liquid air being refluxed at the top of the column. Essentially all of the krypton and xenon in the feed gas stream are adsorbed into the liquid. Waste gas leaving the top of the feed gas stripping column (7.10) subcools the liquid reflux in the reflux subcooler (5.42). Most of this waste gas is then warmed in the feed gas regenerators (5.10) and passes to the jets and then out the stack. The remainder of the gas is circulated to the recycle air heat exchanger (5.11) where it is warmed against incoming high pressure recycle air.

"The liquid from the upper section of the feed gas stripping column flows to the lower section of the feed gas stripping column (7.20). In this section, krypton and xenon are concentrated, and nitrogen is stripped from the liquid, leaving an oxygen-rich product stream. The design krypton-xenon concentration in the crude krypton-xenon product stream is 5000 ppm.

"The more volatile impurities in the feed gas (hydrogen, neon, helium), leave with the waste gas at the top of the feed gas stripping column.

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"The impurities which are less volatile than the crude krypton-xenon product stream (nitric oxide, nitrogen dioxide, acetylene, carbon monoxide, water and nitrous oxide) leave the bottom of the feed gas stripping column in the crude krypton-xenon products stream.

"The feed gas stripping column (7.10 and 7.20) is operated at 45 psia to raise the temperature so that the crude krypton-xenon product stream will have a maximum solubility for impurities which tend to plug the column. For this reason, two feed gas distillation column reboilers (7.30) are furnished for alternate operation.

"C - Recycle Air Refrigeration

The low temperature refrigeration for this process is provided by the recycle air system. A portion of the waste gas from the top of the feed gas stripping column is warmed in the recycle air exchanger (5.11) and fed to the recycle air compressor (1.20). Make-up air, to supply regenerator unbalance, is compressed in the make-up recycle air compressor (1.30), caustic scrubbed in the recycle air purifier (2.10) and combined with the waste gas from the recycle air exchanger (5.11). This gas is compressed to 2500 psia in the recycle air compressor (1.20), dried in the recycle air drier assembly (3.10) and returned to the recycle air heat exchanger (5.11). The high pressure gas is cooled against the incoming low pressure gas. At an intermediate point, additional cooling of the high pressure gas is provided in the Freon-air cooler (4.10) by vaporizing Freon at -80°F and returned to the recycle air exchanger (5.11) for final cooling.

"On leaving the recycle air exchanger (5.11), the gas is expanded to 130 psia forming liquid air which is collected in recycle air separator (37.40). The collected liquid is used for plant refrigeration and condensing service.

"Vapor from this phase separator (37.40) is condensed to liquid in feed gas distillation column reboiler 7.30 furnishing boil-up for the feed gas stripping column (7.10, 7.20). The condensed liquid is subcooled in reflux subcooler 5.42 and expanded into the top of the feed gas stripping column (7.10).

"D - Crude Products Preparation and Storage

The liquid crude krypton-xenon products stream from the feed gas distillation column reboiler (7.30), at 45 psia, is totally vaporized in the crude products vaporizer (40.00) to 100°F . This crude product stream gas then enters the hydrocarbon furnace (2.81) where the hydrocarbons are converted to carbon dioxide and water, and the nitrous oxide converted to nitrogen

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plus oxygen. After cooling to 90° F, the gas is compressed in a diaphragm-type crude products compressor (1.40) to 400 psia. At the compressor suction, a 500 cubic foot crude product gas holder (16.30) is provided to collect vaporized liquid concentrates of krypton and xenon which would occur during plant shutdown. At the discharge side, a crude products storage tank (16.10) is provided, equivalent in capacity to all the liquid hold-up in the plant. Retention time is provided in nitric oxide converter II (2.91) for reduction of the nitric oxide concentration to 0.05 ppm at 90° F. The gas is next caustic-scrubbed at 130 psia in the crude products purifier assembly (2.20) to reduce the carbon dioxide concentration to less than 1 ppm, and nitrogen dioxide concentration to less than 0.05 ppm.

"Water is removed from the crude krypton-xenon product gas in a set of regenerative alumina crude product driers (3.10). The gas is then cooled in crude products column feed cooler 5.30 and totally condensed in the crude products column liquefier 5.40. The saturated liquid is fed to the packed crude products column (8.10) where a separation is made between oxygen, and the krypton-xenon stream. The oxygen is taken overhead, partially refluxed in crude products column condenser (8.20), warmed in the crude products column feed cooler (5.30) and recycled to feed gas compressor 1.10. Liquid air from recycle air phase separator 37.40 is used as coolant for the crude products column condenser (8.20) and crude products column feed liquefier (5.40). Liquid crude krypton-xenon product containing less than 1000 ppm oxygen collects in crude products column reboiler 8.30. Periodically, this liquid is drawn from crude products column reboiler 8.30 and transferred to liquid crude product storage tank (16.20) maintained at -160° F by liquid air from recycle air phase separator 37.40. The storage tank holds 8½ days supply of crude krypton-xenon product at the monthly average flow rate.

"By pressurizing the products concentrate tank (16.20) through the use of electric heat, a shipping container can be charged by equalization.

"E - Crude Products Concentration

Case II process involves the separation of krypton and xenon into gases of 99% purity. This is accomplished by batch distillation in a products concentrate column (8.40). Approximately nine (9) days average supply of crude krypton-xenon product is transferred from crude products concentrate liquid storage tank (16.2) to the products concentrate column reboiler (8.60) at 45 psia. Electric heat is used to furnish vapor boil-up for products concentrate distillation column (8.40) and 2500 psia air, at -160° F, is used as reflux for the products concentrate condenser (8.50).

DECLASSIFIED"F - Pure Product Separation

The products concentrate column (8.40) is started at total reflux. Successive gas fractions are taken off in the overhead stream in the following order - oxygen, krypton, xenon, and bottom. Specification grade xenon and krypton are condensed in cold traps 16.40 and 16.50 respectively. Trace impurities in the krypton and xenon streams are removed by passing the gases through a titanium furnace (2.92) at 1800° F. The oxygen, bottoms gas and intermediate impure fractions are recycled to crude products compressor 1.40. The cold traps (16.40 and 16.50) are periodically heated to pressurize the gases into the shipping containers."

Case I and II Facilities

Facilities to house the Case I and II equipment are shown in drawings SK-2-2966, SK-2-2967 and SK-2-2968, the plot plan, architectural plan and architectural elevation drawings respectively. The new facility would be located to the southeast of the Purex 202-A Building and east of the 293-A Facility and the railroad tunnel. Relocation of an overhead three-inch steam line, an 8-inch buried raw water line, and an electrical encasement would be required. A small crib located to the south of the facility would be required for disposal of spent caustic, compressor lubricant and nitric acid solution condensed as a result of the gas compression.

The following Case I and II facilities would be required:

1. A 40' x 60' x 21' high Butler-type steel building to house the compressors, acetylene removal equipment, and air purifiers.
 2. Four cells of concrete construction within an over-all dimension of 21' x 40' to house the Case I and II cold box (11.00) and process equipment, where high radiation levels are anticipated. The cell housing the cold box would be 29' high; the other cells would be 20' high.
 3. A gas analysis laboratory to house a new mass spectrometer for use in krypton-xenon analysis and process control.
 4. A platform at the northeast corner of the new facility for filling of shielded cylinders with product gases from either Case I or Case II and for cylinder load-out and storage.
 5. Two 10-inch stainless steel lines extending from the T-XB tower of the 293-A Building to the new facility, for supply and return of process gas.
 6. Electrical facilities, as delineated on drawing number SK-2-2964.
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DECLASSIFIEDCase I and II Operating and Radiological Control

The Case I and II equipment would be operated remotely from the Purex 202-A Building Head-End Control Room, to permit optimum use of Purex operating, supervisory and Technology personnel. Routine inspection and minor maintenance of gas compression equipment would be performed locally under SWP procedures. The present SWP change facilities in the new Purex Crane Maintenance Facilities would be used. Major scheduled maintenance would be performed on about a monthly basis, during periods of shut-down of Purex dissolver operations. Where high radiation levels would otherwise be anticipated, process equipment within the concrete cells would be purged to remove gaseous krypton-85 activity prior to initiating maintenance.

Shielding criteria and specifications for the new facility are presented in References 5 and 3 respectively. Briefly, the thickness of concrete specified in drawings SK-2-2967 and 2968, would be adequate to reduce radiation to 10 mr/hr or less on contact.

Product gases from Case I and Case II would be stored and shipped in 10-inch diameter Type ICC-3AA carbon dioxide cylinders. The Case I (mixed rare gas) and the Case II krypton gas cylinders would be shielded with about 2-inches of lead (equivalent) on the sides and 3-inches of lead at the ends of the cylinders. The Case II xenon gas cylinders would be shielded with 1-inch of lead on the sides and 1½-inches at the ends. The above shielding is based upon the shipment (or storage) of one cylinder per week of Case I gas or one cylinder per week of Case II krypton plus one cylinder per week of Case II xenon. Radiation levels would be 10 mr/hr or less at one meter from the cylinder shields. Shipment of the product gases would presumably be either by rail or motor carrier, with AEC escort and under a Bureau of Explosives permit.

As shown in drawings SK-2-2966 and SK-2-2962, a small crib has been provided directly to the south of the new facility. Spent caustic, compressor lubricant and condensed nitric acid solution from the process equipment, discharged from a catch-mix tank, would overflow to the crib. Since it is suspected that compressor lubricant would be present in the acid stream, the return of this acid to the Purex process is not recommended. It is anticipated that about 0.001 curie per day of mixed fission products, along with minor traces of Iodine -131 would be cribbed. The strontium-90 component, the isotope of greatest concern from the standpoint of cribbing, would be of the order of 0.00001 curie per day. Controlled neutralization of wastes in the catch-tank might be necessary to reduce the Sr-90 concentration in the liquid waste to less than 8×10^{-7} microcuries per cc for safe disposal.

Case I Acetylene Removal

Acetylene removal equipment has been provided in the flowsheet for the Case I gas pretreatment. If not removed to less than 1/20 part per

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million in the feed gas, acetylene could collect as a solid suspended in liquid oxygen in the reboilers of the low temperature distillation column and present a possible explosion hazard (Reference 16). To date, analytical procedures which have been used to sample the Purex T-XB off-gas have not revealed the presence of acetylene above 1/20 ppm (Reference 9).

Uranium metal (ingot uranium) contains 200 to 300 parts per million of carbon. Under current Purex dissolving conditions, assuming all the carbon to be present as uranium carbide, off-gas liberated from the dissolver could contain about 200 to 400 parts per million of acetylene gas. Undoubtedly, a portion of this acetylene would become nitrated in the dissolver. Reference 3 cites literature references which indicate that the Purex silver reactors, operated at 370°F, should serve as catalyst chambers for the oxidation of acetylene into CO₂ and H₂O. The retention time of off-gas in the silver reactors is about 4 seconds at peak off-gas evolution.

Further gas sampling tests and other studies are contemplated to define the acetylene content of the off-gas as a function of the over-all Purex dissolver operation, and specifically as a function of the state of regeneration of the silver nitrate bed. The cost estimates presented in this report include acetylene removal equipment and auxiliaries and may therefore be conservative.

Removal of NO and NO₂ from the Feed Gas

The major problem encountered in the design of the rare gas recovery equipment was the removal of NO and NO₂ from the feed stream to one part per million or less, to avoid subsequent plugging difficulties in the low temperature recovery equipment. The following process alternates, reviewed in detail in Reference 3, were considered:

1. Homogeneous gas phase oxidation at 800 psia and ambient temperature, according to the reaction $\text{NO} + \frac{1}{2} \text{O}_2 \rightleftharpoons \text{NO}_2$, with subsequent NO₂ removal in the regenerators.
2. Catalytic reaction of NO and NO₂ with hydrogen.
3. Catalytic oxidation of NO over a silica gel bed at 40°F.
4. Oxidation of NO within the regenerators.
5. Absorption of NO in water or caustic solutions at various pressures.

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16. Kerwat, E., "Some Aspects of Hydrocarbons in Air Separation Plants," Chemical Engineering Progress, Vol. 54, No. 10, page 96, 1958.

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Alternate 2, the catalytic reaction of NO and NO₂ with hydrogen, was not practical owing to the large quantities of hydrogen needed, not only to react with the oxides of nitrogen, but with the accompanying oxygen in the gas. Control of hydrogen addition with varying feed gas rate and composition would be difficult.

Alternate 3, oxidation of NO to NO₂ at 40° F over silica gel would involve a prior gas drying step to a -60° F dew point and a rather complex over-all system.

Alternate 4, oxidation of NO to NO₂ within the feed gas regenerators, was not feasible owing to the low rate of oxidation at the reduced temperatures. Complete plugging of the regenerators with solid N₂O₃ and other sub-oxides would occur.

Alternate 5, the absorption of NO and NO₂ in water under pressure was not considered attractive owing to the higher preferential solubility of the rare gases in water and the reliberation of NO from the solution according to the reaction: $3\text{NO}_2 + \text{H}_2\text{O} \rightarrow 2\text{HNO}_3 + \text{NO}$.

Alternate 1, the homogenous gas phase oxidation of NO to NO₂ under pressure was considered the most reliable method to provide the assured NO - NO₂ removal to 1 ppm or less. The recommended pressure of 800 psia for the oxidation was derived from economic studies of capital and operating costs as a function of compressor discharge pressure. At 800 psia about a 3-minute retention period would be required for the degree of oxidation required.

Under Alternate 1, those parts of the compressor exposed to the process gas would be fabricated largely of stainless steel and other corrosion resistant alloys. Probably a somewhat lower service factor will be obtained than from a standard compressor handling air alone. In Reference 4, an alternate compression system, involving a prior gas drying step, has been proposed to permit the use of a standard gas compressor. The Case I and II estimates reflect the use of the stainless steel compressor and are therefore conservative.

Case I and II Optimum Refrigeration Cycle

The refrigeration system described previously, entailing the use of a 2500 psia compressor and a supplementary Freon refrigeration unit, represents the optimum refrigeration cycle. Other thermodynamic cycles are discussed in Reference 3. The use of purchased liquid nitrogen for refrigeration is technically feasible although the cost of the liquid nitrogen would be in excess of \$200,000 per year, when processing the entire quantity of Purex dissolver off-gas.

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CASE III FACILITIES

INTRODUCTION

The objective of the Hanford Case III facilities, described in this section, is to alter the isotopic composition of the xenon produced in the Case II facilities. In this study, the product desired was considered to be an isotopic mixture of xenon having a composite cross-section of less than one barn with a maximum of 10^{-14} per cent Krypton-85. The cross-section of each xenon isotope of concern is as follows:

Xe 131	120 barns
Xe 132	5 barns
Xe 134	.2 barns
Xe 136	.15 barns

The regular variation of cross-section with atomic weight is fortunate since the desired composition could be obtained by "splitting" the mixture between the 132 and 134 isotopes. Several processes for Case III were investigated. These were: 1) Simple distillation, 2) Gas centrifugation, 3) Gaseous diffusion and 4) Thermal diffusion.

Visits were made to A.E.C. sites employing several of the above processes and consultations were held with workers in these fields. As a result of a meeting with personnel at Oak Ridge, a detailed economic evaluation of the various processes was made by the Operations Analysis Division at Oak Ridge (Reference 10). The Reference 10 study, in conjunction with other consultations indicated that thermal diffusion was the most economical process for the construction of new facilities at HAPO to handle the Case III design flow rate (1,700 grams/day). This section of the report presents the results of the design study of a thermal diffusion plant adjacent to the Hanford Case I and II facilities. The design presented is based entirely on theoretical calculations, since a thermal diffusion plant of the size and configuration described has not, to the writers' knowledge, been built as yet. Because of possible deviations between theory and practice, it is recommended that a research and development program be undertaken before Case III detailed designs are completed.

CASE III PROCESS DESCRIPTION

For a detailed description of the thermal diffusion process, the reader is referred to References 17, 18, and 19. However a brief description of the thermal diffusion process is as follows:

When a temperature gradient exists in a gaseous mixture, small diffusion currents are established with one component being transported away

17. "The Separation of Isotopes by Thermal Diffusion", R. C. Jones & W. H. Furry, Reviews of Modern Physics, Vol. 18, page 151.
18. "The Theory of Isotope Separation", Karl Cohen, McGraw-Hill Book Co., Inc., 1951.
19. "Nuclear Chemical Engineering", Benedict & Pigford, McGraw-Hill Book Co., Inc., 1957.

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from the heat source and the other being transported toward the heat source. This is known as the thermal diffusion effect, and this effect makes possible the thermal diffusion column.

The mixture to be separated is confined in a narrow, annular space between two concentric, vertical tubes. The inner tube is heated and the outer tube is cooled. The temperature difference between the inner and outer walls establishes a small difference in isotopic content at the walls due to the thermal diffusion effect. The concentration of the lighter isotope is increased adjacent to the hot wall, while the isotopic concentration of the heavier component is increased adjacent to the cold wall. At the same time, convection currents are established in the annular space with the gas moving upward in the zone adjacent to the hot wall and moving downward in the zone adjacent to the cold wall. This counterflow multiplies the concentration difference and makes possible substantial separations in a practical length of column.

For the Case III study, the composition of the feed gas, expressed as weight percent, was as follows:

<u>Isotope</u>	<u>Percent By Weight</u>
Krypton (all isotopes)	1.0
Xenon 128, 129, 130	0.15
131	11.45
132	17.16
134	27.25
136	43.00

The controlling physical property of the product gas was the neutron cross-section of the gas. Since the end use of the xenon was for liquid xenon gamma scintillation counters, it was necessary to produce a gas with a low neutron cross-section. On this basis, the composition of the product gas and raffinate were derived as follows:

<u>Product Gas</u>		<u>Raffinate Gas</u>	
<u>Isotope</u>	<u>Percent By Weight</u>	<u>Isotope</u>	<u>Percent By Weight</u>
Xe ¹³⁶	71.9	Xe-128	0.1
Xe ¹³⁴	24.3	Xe-129	
Xe ¹³²	3.3	Xe-130	
Xe ¹³¹	0.5	Xe-131	22.7
		Xe-132	31.3
		Xe-134	30.1
		Xe-136	13.9
Kr (all isotopes)	10 ⁻¹⁴	Kr - Total	2.00

This product composition will result in a composition neutron cross-section for the gas of about 0.9 barns.

Feed gas for the thermal diffusion plant is obtained from the gas storage facilities of Case II. As shown on the Engineering Flow Diagram, SK-2-17892, the gas pressure is reduced through a pressure reducing valve and fed to the individual columns via a gas header. The plant consists of 30 thermal diffusion columns operated in parallel. Since each thermal diffusion column consists of two adjacent physical columns, a total

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of 60 physical columns is required. These columns will hereafter be called the "enriching" and the "stripping" columns. In a typical enriching column, the concentration of Xe^{136} is increased and the concentration of Xe^{131} is decreased; in the stripping column, the reverse holds true. Figure 1 shows the variation of the gas composition as a function of position in the column. The point 0 corresponds to either the bottom of the stripping column or the top of the enriching column. By the same token, a positive distance is measured up the stripper and a negative distance is measured down the enricher.

A mixing pump between the columns constantly interchanges gas from the top of one column and the bottom of the adjacent column to permit both columns to function as a single long column. Feed gas is introduced into the middle of the overall column (i.e. the base of the stripping column); however, other inlets are provided so that the feed point can be adjusted to satisfy varying process demands. The arrangement of columns from a process standpoint is one of the variables that would have to be explored further in an R. & D. Program. In the Reference 10 study, a tapered cascade arrangement was used in which 16 to 24 columns of various diameters were connected in a series cascade and 14 to 28 cascades in parallel comprised the entire plant. For this arrangement, the total number of columns required for the design flow rate (1,700 grams/day) would be about 450 columns with the diameter of the columns varying from 12-inches to 1-inch.

The design approach in the study presented herewith was to provide identical columns which could be arranged as desired to conform to process demands. The arrangement presented comprises 30 pairs of columns operating in parallel; however, many other configurations could be obtained to accommodate changing demands for product composition, percent recovery, etc., merely by rerouting the interconnecting piping.

For example, if a demand arose for high purity Xe^{136} , it would be relatively simple to connect sufficient columns in series to obtain the product desired. It would also be possible to produce several different products simultaneously, merely by adding additional product compressors or cold traps. Some other advantages of this arrangement are:

1. Lower fabrication costs for fewer units of identical design.
2. Maintenance procedures for the columns and their auxiliaries are simplified and spare parts inventory is reduced.

CASE III EQUIPMENT DESCRIPTION

A. Thermal Diffusion Column

The column itself consists of three concentric tubes: an inner tube, an outer tube, a water jacket with lead shielding as shown on SK-2-2960. The inner tube, which will be referred to as the hot wall, is 11.485 inches in diameter and 24 feet long. The inner tube which is fastened to the outer tube at the bottom of the column is provided with spacer blocks at 6-foot intervals along the length of the column.

Figure 1.

ISOTOPIC COMPOSITION
vs
COLUMN POSITION

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DISTANCE FROM FEED POINT (FT.)

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The purpose of the spacer blocks is to align the inner tube such that the hot and cold wall are concentric throughout the length of the column. Thermal diffusion occurs within the 0.188-inch wide gap between the hot and cold walls. Both the outer surface of the inner tube and the inner surface of the outer tube are machined and polished since the concentricity of the tubes has a significant effect upon the efficiency of the column and a high polish is needed to reduce the radiant heat loss. The water jacket is made of 14-inch pipe with a .625-inch wide annulus between the outer wall of the column and the inner wall of the water jacket. The estimated power consumption per column during normal operating is 41 KW. The corresponding water flow rate is 9.25 gallons per minute per column for a 30° F temperature rise. Thus, the total cooling water consumption for the entire plant is 555 gallons per minute.

B. Column Heaters

Heat is supplied by means of three U-bend resistance heaters inside the inner tube. Since the operating temperature of the outer surface of the inner tube is 1600° F, the heater elements will operate at 1800 to 2000° F depending on the emissivities of the tube and heater surfaces. To attain these high temperatures, Kanthal heating elements, capable of operation at temperatures up to 2450° F are used. The optimum method for heating the columns is another item that would be subject to investigation in a R. & D. program. Among other possible heating methods are:

1. Using the inner tube itself as a heater through the use of a low voltage, high current system, with several columns connected in series from a power supply standpoint.
2. Use of a condensing liquid, such as sodium, inside the tube with one or more external reboilers supplying the heat for the entire plant.

The above methods were considered in this process study, but were not adopted because of the uncertainties involved. For this reason the resistance heating method using commercially available heaters was used.

The entire column shell is wrapped with lead shielding $\frac{1}{4}$ -inch thick. This reduces the radiation level (at the midpoint between two columns), to 6 mr/hr to permit personnel to work an entire shift within the column room if necessary.

C. Gas Pumps

Each pair of columns requires two gas pumps. These pumps are positive displacement, duplex-action pumps. The mixing pump as shown on SK-2-17892 intermixes the gas from the bottom of the stripping column and the top of the enriching column, thereby tying the two columns together process-wise. The flow through both sides of the pump is approximately 1,000 liters (STP) per day. The pump has a local speed control so that the flow can be adjusted from 50 to 150 per cent of the nominal flow.

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The product and raffinate withdrawal pump is also a duplex-action pump with the product being pumped in one cylinder and the raffinate in the other cylinder. This pump has a local adjustment for stroke length in each cylinder and a remote control for pumping speed. The total flow through both sides of this pump is 9.6 liters (MTP) per day. The product withdrawal pump also acts as seal between the parallel stages and prevents parasitic recycling of the gas.

The product and raffinate gases from all the columns are compressed to 500 psig using two gas compressors. Each compressor has a 200 liter per day capacity. Gas holders in the product and raffinate header systems provide the necessary surge capacity between the gas withdrawal pumps and the compressors. It might be feasible to replace the product and raffinate compressors and gas holder with cold-traps using refrigeration from Case I.

D. Column Control and Instrumentation

The most important operating variables requiring control in a thermal diffusion process are the gas withdrawal rate, the temperature drop across the gas film, and the gas pressure. Product and raffinate are withdrawn from the column by means of a duplex positive displacement gas pump designed such that the ratio of product to raffinate can be controlled by adjusting the stroke length of the two pistons. In addition, the total gas withdrawal rate is controlled by a variable speed coupling between the drive motor and the pump.

Temperature control is effected by varying the power input to the individual columns. A reactrol system as shown on SK-2-2965 is provided for each column. Since the heater elements are continuous for the full height of the column, the possibility of local hot spots is avoided. Thus, temperature is sensed only at the midpoint of the column. The column wall temperature which controls the reactrol would be recorded and would activate a high and low-temperature alarm as shown on SK-2-2965.

The cooling water flow rate is controlled by the temperature of the cooling water discharge. In addition, high temperature control-alarms on the cooling water open circuit breakers on the column heater systems so that in the event of low flow, or water failure, the columns are shut down.

Process control is affected by periodic sampling of both the product and raffinate streams. For this purpose the analytical laboratory with mass spectrographic equipment which has been provided for Case I and II will be used. Based on the composition of the samples, the product and raffinate withdrawal rates and/or the flow ratio between the two streams would be adjusted for process control. The pressure within the columns would be controlled by a single throttling valve on the main feed gas header. The feed gas header was sized such that the pressure drop to the furthest column would be less than .01 inches of H₂O to assure equal pressure in all columns.

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CASE III ARCHITECTURAL

The case III thermal diffusion facilities would be housed in a concrete structure as shown on SK-2-2963. A control room and laboratory, located adjacent to the column building would be provided, with access via an air lock, between the buildings. When Case III thermal diffusion facilities are built, control of the Case I and II equipment would be transferred to this control room.

The column building would contain the thermal diffusion columns, pumps, switchgear, compressors, and interconnecting piping. Access to the pumps, piping, and electrical connections would be from the floor level while the piping connections at the top of the columns would be reached from a platform at the 21' level.

The rear portion of the building would be used as a maintenance area. The maintenance area would be reached through a labyrinth as shown on SK-2-2963. The walls comprising the labyrinth would extend upward to the top of the columns with space above for crane passage. Thus, columns could be brought from the column room to the maintenance area. A monorail in the maintenance area would be used for disassembly of the columns and for moving the component parts within the maintenance area.

The end of the building would contain the compressor room, gas bottling room and stairways to the upper walkway.

SUGGESTED RESEARCH AND DEVELOPMENT PROGRAM

As mentioned previously, the Case III thermal diffusion column design was derived from theoretical relationships. Design and operating information was not available on large thermal diffusion columns, of the size necessary to handle the quantity of xenon produced at the Hanford Purex Plant. A research and development program to optimize the design of the proposed 12-inch column would require about 2 years for completion, of which roughly one year would be required for procurement of the column. Any future construction program entailing the use of large columns might be delayed by this two-year R. & D. period, unless a non-optimum design were to be adopted with resulting increased capital cost to provide high operating flexibility.

Research and development data on large columns would be valuable not only for the above specific xenon application but for possible future isotopic separations in other fields. On this basis, it is suggested that a research and development program be initiated as soon as possible. This program would be directed specifically to the separation of xenon isotopes in large columns. Operating parameters would be correlated against theory. The specific objectives, related to the xenon separation would be:

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1. Optimize geometrical relationships of the column.
2. Optimize the method of heating.
3. Determine radiation emissivity factors governing power input.
4. Optimize the mechanical design of the column.
5. Determination of the method(s) of stream composition measurement and column control.

The above R. & D. Program is estimated to require about two years for execution. The first year would be primarily concerned with defining, designing, and installing the experimental facilities, while the second year would be devoted to experimental measurements and verification of the selected column design, in relation to control and operating characteristics. The cost of the program is anticipated to be equally divided between direct liquidations for scientific and engineering personnel and direct materials and services associated with the fabrication of special equipment and supporting facilities. The total direct costs over the two-year period are estimated to be \$275,000. A detailed review of the R. & D. Program is being prepared by H.L.O. and will be issued about February, 1959.

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GENERAL ELECTRIC COMPANY

HANFORD ATOMIC PRODUCTS OPERATION

PROJECT COST ESTIMATE SUMMARY

TITLE GAS PURIFICATION PLANT CASE I STUDY STAGE	PROJECT NO.
	DATE 12-11-58
	PREPARED BY Estimating/RLL
	CHECKED BY
PREPARED FOR CHEMICAL PROCESSING DEPARTMENT	

INSTALLATION & ERECTION COST

I. CONSTRUCTION OPERATION 30,000

II. G. E. PLANT FORCES 10,000

III. FIXED PRICE CONTRACTOR 1,320,000

IV. SUPERVISION OF CONSTRUCTION (TITLE III)..... 40,000

V. PROJECT START - UP 20,000

VI. CONTINGENCY 270,000

VII. ESCALATION

SUB-TOTAL 1,690,000

VIII. GENERAL OVERHEAD (____ % OF SUBTOTAL) 85,000

TOTAL INSTALLATION & ERECTION COST 1,775,000

MANAGEMENT SERVICES 51,000

DESIGN SERVICES (TITLE I & II) A-E 98,000
G.E. 96,000

TOTAL FUNDS REQUESTED 2,020,000

TRANSFERRED CAPITAL PROPERTY

TOTAL PROJECT COST 2,020,000

* INCLUDES : _____ FOR EQUIPMENT PROCURED BY G.E.	APPROVED BY	DATE
	(ESTIMATING)	
	(ENGINEERING)	
	(FINANCIAL)	

GENERAL ELECTRIC COMPANY

HANFORD ATOMIC PRODUCTS OPERATION

PROJECT COST ESTIMATE SUMMARY

TITLE GAS PURIFICATION PLANT CASE I & II STUDY STAGE (CUMULATIVE COST)	PROJECT NO.
	DATE 12-11-58
	PREPARED BY Estimating/RLL
	CHECKED BY
PREPARED FOR CHEMICAL PROCESSING DEPARTMENT	

INSTALLATION & ERECTION COST

I. CONSTRUCTION OPERATION 30,000

II. G. E. PLANT FORCES 10,000

III. FIXED PRICE CONTRACTOR 1,385,000

IV. SUPERVISION OF CONSTRUCTION (TITLE III) 45,000

V. PROJECT START - UP 20,000

VI. CONTINGENCY 285,000

VII. ESCALATION

SUB-TOTAL 1,775,000

VIII. GENERAL OVERHEAD (5 % OF SUBTOTAL) 90,000

TOTAL INSTALLATION & ERECTION COST 1,865,000

MANAGEMENT SERVICES 57,000

DESIGN SERVICES (TITLE I & II) A-E 105,000
G.E. 103,000

TOTAL FUNDS REQUESTED 2,130,000

TRANSFERRED CAPITAL PROPERTY

TOTAL PROJECT COST 2,130,000

# INCLUDES \$ _____ FOR EQUIPMENT PROCURED BY G.E.	APPROVED BY	DATE
	(ESTIMATING)	
	(ENGINEERING)	
	(FINANCIAL)	

PROJECT COST ESTIMATE SUMMARY

TITLE GAS PURIFICATION PLANT CASES I, II, & III (CUMULATIVE COST) STUDY STAGE	PROJECT NO.
	DATE 12-11-58
	PREPARED BY Estimating/RLL
	CHECKED BY
PREPARED FOR CHEMICAL PROCESSING DEPARTMENT	

INSTALLATION & ERECTION COST

I. CONSTRUCTION OPERATION	60,000	DECLASSIFIED
II. G. E. PLANT FORCES	20,000	
III. FIXED PRICE CONTRACTOR	4,030,000	
IV. SUPERVISION OF CONSTRUCTION (TITLE III).....	125,000	
V. PROJECT START - UP	50,000	
VI. CONTINGENCY	825,000	
VII. ESCALATION		
SUB-TOTAL	5,110,000	
VIII. GENERAL OVERHEAD (<u>5</u> % OF SUBTOTAL)	255,000	
TOTAL INSTALLATION & ERECTION COST	5,365,000	
MANAGEMENT SERVICES	160,000	
DESIGN SERVICES (TITLE I & II)	A-E G.E.	105,000 220,000
TOTAL FUNDS REQUESTED		5,850,000
TRANSFERRED CAPITAL PROPERTY		
TOTAL PROJECT COST		5,850,000

* INCLUDES \$ _____ FOR EQUIPMENT PROCURED BY G.E.

APPROVED BY	DATE
(ESTIMATING)	
(ENGINEERING)	
(FINANCIAL)	

TABLE III-A

HEAT AND MATERIAL BALANCE

Composition (Volume Percent)

Point	Temp. °F	Pressure PSIA	Total Flows Mols/Hour	H ₂ O	NO	NO ₂	N ₂ O	A	O ₂
1	100	10.7	221.673	5.0	4.6	0.4	0.1	0.9	10.0
2	90	10.7	1.123	---	---	---	---	---	99.9
3	100	10.7	222.796	4.968	4.569	0.398	0.100	0.895	10.4
4	100	800	199.839+H ₂ O	Satd.	0.511	1.202	0.111	1.000	9.3
5	450	800	199.839+H ₂ O	Satd.	0.511	1.202	0.111	1.000	9.3
6	500	800	199.839+H ₂ O	Satd.	0.511	1.202	0.111	1.000	9.3
7	150	800	199.839+H ₂ O	Satd.	0.511	1.202	0.111	1.000	9.3
8	100	800	199.839+H ₂ O	Satd.	0.511	1.202	0.111	1.000	9.3
9	100	800	198.879+H ₂ O	Satd.	1 ppm	1.717	0.111	1.003	9.1
10	100	45	198.879+H ₂ O	Satd.	1 ppm	1.717	0.111	1.003	9.1
11	-275	45	190.156	---	1 ppm	---	1 ppm	1.027	8.7
12	-297	45	190.156	---	1 ppm	---	1 ppm	1.027	8.7
13	-275	45	23.720	---	0.018	---	0.018	---	99.4
14	-275	45	22.590	---	0.018	---	0.018	---	99.4
15	-275	45	1.1298	---	0.018	---	0.018	---	99.1
16	100	45	1.1298	---	0.018	---	0.018	---	99.1
17	100	45	1.1298	---	0.017	---	---	---	99.1
18	100	400	1.1298+H ₂ O	Satd.	0.017	---	---	---	99.1
19	100	100	1.1297+H ₂ O	Satd.	---	0.017	---	---	99.1
20	100	45	1.1289+H ₂ O	Satd.	---	---	---	---	99.1
21	100	45	1.1289	---	---	---	---	---	99.1
22	-265	45	1.1289	---	---	---	---	---	99.1
23	-275	45	1.1289	---	---	---	---	---	99.1
24	-275	45	1.1663	---	---	---	---	---	99.8
25	-275	45	0.54	---	---	---	---	---	99.8
26	-275	45	0.54	---	---	---	---	---	99.8
27	-275	45	1.123	---	---	---	---	---	99.8
28	-164	40	1.0539	---	---	---	---	---	0.6
29	-164	40	1.048	---	---	---	---	---	---
*30	-164	40	0.4075	---	---	---	---	---	0.6
*31	100	14.7	0.1231	---	---	---	---	---	---
*32	100	14.7	0.6918	---	---	---	---	---	---
33	-297	45	219.040	---	---	---	---	0.892	7.
34	-293	45	219.040	---	---	---	---	0.892	7.
35	-203	45	198.713	---	---	---	---	0.892	7.
36	-300	20	198.713	---	---	---	---	0.892	7.
37	-289	18	198.713	---	---	---	---	0.892	7.
38	86.5	14.7	207,436	---	---	1.646	0.107	0.874	7.

* Batch Distillation

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	CO ₂	H ₂	He	Ne	Kr	Xe	N ₂	Total Flow lbs/hr.	Enthalpy BTU/HOUR
	0.03	0.6	5 ppm	15 ppm	4.1 ppm	21.4 ppm	78.4	6219.0	1,214,768
8	---	---	---	---	---	---	0.02	35.9	6,075
39	0.030	0.597	4.97 ppm	14.92 ppm	4.08 ppm	21.3 ppm	77.90	6254.9	1,221,316
37	0.033	0.667	5.56 ppm	16.67 ppm	4.56 ppm	23.8 ppm	87.13	5596.8+H ₂ O	1,068,725
37	0.033	0.667	5.56 ppm	16.67 ppm	4.56 ppm	23.8 ppm	87.13	5596.8+H ₂ O	1,564,287
37	0.033	0.667	5.56 ppm	16.67 ppm	4.56 ppm	23.8 ppm	87.13	5596.8+H ₂ O	1,634,990
37	0.033	0.667	5.56 ppm	16.67 ppm	4.56 ppm	23.8 ppm	87.13	5596.8+H ₂ O	1,139,508
37	0.033	0.667	5.56 ppm	16.67 ppm	4.56 ppm	23.8 ppm	87.13	5596.8+H ₂ O	1,068,725
05	0.033	0.669	5.56 ppm	16.67 ppm	4.56 ppm	23.8 ppm	87.36	5570.6+H ₂ O	1,065,991
05	0.033	0.669	5.56 ppm	16.67 ppm	4.56 ppm	23.8 ppm	87.36	5570.6+H ₂ O	1,065,991
90	3 ppm	0.685	6 ppm	17 ppm	5 ppm	24 ppm	89.49	5384.5	539,092
90	3 ppm	0.685	6 ppm	17 ppm	5 ppm	24 ppm	89.49	5384.5	509,285
11	0.053	---	---	---	0.081	0.419	---	770.0	12,859
11	0.053	---	---	---	0.081	0.419	---	733.4	74,321
11	0.053	---	---	---	0.081	0.419	---	36.6	638
11	0.053	---	---	---	0.081	0.419	---	36.6	6,690
13	0.053	---	---	---	0.081	0.419	0.017	36.6	6,690
13	0.053	---	---	---	0.081	0.419	0.017	36.6+H ₂ O	6,611
13	0.053	---	---	---	0.081	0.419	0.017	36.6+H ₂ O	6,678
82	---	---	---	---	0.081	0.419	0.017	36.6+H ₂ O	6,684
32	---	---	---	---	0.081	0.419	0.017	36.6	6,684
82	---	---	---	---	0.081	0.419	0.017	36.6	3,833
82	---	---	---	---	0.081	0.419	0.017	36.6	638
8	---	---	---	---	---	---	0.02	53.2	5,471
8	---	---	---	---	---	---	0.02	17.3	1,777
8	---	---	---	---	---	---	0.02	17.3	305
8	---	---	---	---	---	---	0.02	35.9	3,695
17	---	---	---	---	16.103	83.880	---	130.3	0(Base)
---	---	---	---	---	71.000	29.000	---	102.2	5,437
17	---	---	---	---	16.103	83.880	---	50.4	0
---	---	---	---	---	99.0	1.0	---	10.4	---
---	---	---	---	---	1.0	99.0	---	90.5	---
18	---	0.595	5 ppm	14.6 ppm	---	---	91.393	6188.9	551,981
18	---	0.595	5 ppm	14.6 ppm	---	---	91.393	6188.9	558,552
18	---	0.595	5 ppm	14.6 ppm	---	---	91.393	5614.6	506,718
13	---	0.595	5 ppm	14.6 ppm	---	---	91.393	5614.6	506,718
18	---	0.595	5 ppm	14.6 ppm	---	---	91.393	5614.6	536,525
9	0.032	0.583	4.8 ppm	14.5 ppm	---	---	89.266	5812.4	1,093,188

TABLE III-B
HEAT AND MATERIAL BALANCE
Recycle Air System

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<u>Point</u>	<u>Temp.</u> <u>°F</u>	<u>Pressure</u> <u>PSIA</u>	<u>Mols/Hour</u>	<u>Lbs/Hour</u>	<u>Enthalpy</u> <u>BTU/Hour</u>
101	100	14.7	9.66	279.9	52,937
102	100	40	9.66	279.9	52,937
103	100	40	61.00	1767.2	334,280
104	100	2500	61.00	1767.2	313,540
105	0	2500	61.00	1767.2	244,917
106	-75	2500	61.00	1767.2	211,060
107	-175	2500	61.00	1767.2	130,540
108	-175	2500	14.00	405.6	29,960
109	-165	2500	14.00	405.6	32,060
110	-135	2500	14.00	405.6	38,080
111	-270	130	68.28	1978.1	172,638
112	-270	130	.39	11.3	273
113	-270	130	.39	11.3	1,080
114	-270	130	14.39	416.9	39,160
115	-280	98	.70	20.3	490
116	-280	98	.70	20.3	1,932
117	-270	130	6.81	199.3	4,816
118	-270	130	30.00	869.1	83,100
119	-270	130	30.31	878.1	83,959
120	-285	45	0.70	20.3	1,932
121	-270	130	30.00	869.1	21,000
122	-292	130	30.00	869.1	14,459
123	-297	45	30.00	869.1	14,459
124	-270	130	7.97	230.9	5,579
125	-293	45	20.33	589.0	51,842
126	-290	45	51.34	1487.3	130,917
127	95	40	51.34	1487.3	280,060

TABLE III C
PROCESS EQUIPMENT LIST

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<u>Item No.</u>	<u>Description</u>	<u>Item No.</u>	<u>Description</u>
1.10	Feed Gas Compressor	4.00	Freon Refrigeration Unit
1.20	Recycle Air Compressor	4.10	Freon Air Cooler
1.30	Make-Up Recycle Air Compressor	5.10	Feed Gas Regenerators
1.31	Compressor Aftercooler	5.11	Recycle Air Heat Exchanger
1.40	Crude Products Compressor	5.30	Crude Products Column Feed Cooler
1.50	Vacuum Pump	5.40	Crude Products Column Feed Liquefier
2.10	Recycle Air Purifiers	5.41	Regenerator Preheater
2.11	Caustic Circulating Pumps	5.42	Reflux Subcooler
2.12	Caustic Mix Tank	5.50	Plant Defrost Heater
2.20	Crude Products Purifier	7.10	Feed Gas Stripping Column
2.21	Caustic Circulating Pumps	7.20	Feed Gas Distillation Column
2.22	Caustic Mix Tank	7.30	Feed Gas Distillation Column Reboiler
2.61	Acetylene Reactor	8.10	Crude Products Column
2.62	Acetylene Reactor Heat Exchanger	8.20	Crude Products Column Condenser
2.63	Reactor Preheater	8.30	Crude Products Column Reboiler
2.64	Reactor Gas Cooler	8.40	Products Concentrate Distillation Column
2.71	Nitric Oxide Converter I	8.50	Products Concentrate Column Condenser
2.81	Hydrocarbon Furnace	8.60	Products Concentrate Column Reboiler
2.82	Hydrocarbon Furnace Aftercooler	16.10	Crude Products Storage
2.91	Nitric Oxide Converter II	16.20	Products Concentrate Liquid Storage
2.92	Titanium Furnace	16.21	Products Concentrate Heater
2.94	Titanium Furnace Aftercooler	16.30	Crude Products Gas Holder
3.10	Recycle Air Drier	16.40	By Product Cold Trap
3.20	Crude Products Drier	16.50	Product Cold Trap
3.41	Drier Reactivation Blower	16.60	Bomb Dollies
3.42	Drier Reactivation Aftercooler	16.70	Bomb Dollies
3.43	Drier Reactivation Blower Precooler	37.10	Oil Separator
3.44	Drier Reactivation Heater	37.20	Caustic Entrainment Trap
3.45	Drier Reactivation Blower	37.40	Recycle Air Phase Separator
3.46	Drier Reactivation Blower Aftercooler	37.50	Caustic Entrainment Trap
3.47	Drier Reactivation Blower Precooler	37.70	Drier Phase Separator
3.48	Drier Reactivation Heater	40.10	Crude Products Vaporizer

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ELEVATION C.C. SK-2-2967

ELEVATION A-A SK-2-2967

[illegible]

HW-58400
PAGE -15

SCOPE

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FOR DESIGN
NOT FOR CONSTRUCTION

NO.		DESCRIPTION		REV.	BY	DATE	APP'D	FOR	DATE
REVISIONS									
DRAW. NO.		SK-2-2966		REV. NO.	—		DATE	—	
CLASSIFICATION		NONE		CLASSIFIED BY		DATE			
SCALE		1" = 20'-0"		BY		FOR		DATE	
DRAWN		DATE		CHECKED		DATE		ISSUED	
DES. ENG.		DATE		CHECKED		DATE		ISSUED	
INST. NO.		12128		PROJECT NO.		DO-0806		U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC	
GAS PURIFICATION PLANT PLOT PLAN CASE I & II									
202-A		SK-2-2966		REV. NO.		—		—	

DATE	DWG. NO.	DRAWING TITLE
		REFERENCE DRAWINGS
NEXT USED ON

HW-58400
PAGE

GENERAL NOTES
1. PIPING SHALL AS PER CODES IN SPEC. HW 5500

SCOPE

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SK-2-2462	MUREX OFF GAS FACILITY PLOT PLAN
N-2-57015	OVER HEAD PIPING 281A TO 202A
N-2-35060	PIPING GEN. ARRGT. STM. & AIR
N-2-33597	STM. & AIR LINES RELOCATION
N-2-57022	EQUIP. ARRGT. & PIPING
NO. 0000 NO.	DRAWING TITLE
NO.	REFERENCE DRAWINGS
NEXT USED ON	

NO.	DESCRIPTION	REV BY	DATE	APP'D BY	DATE	FOR	DATE
REVISIONS							
CLASSIFICATION		NONE					
CLASSIFIED BY		J. H. 12-1-51					
SK-2-2959							
SCALE: 1"=10'-0"							
DRAWN BY		J. H. 12-1-51					
CHECKED BY							
ISSUED BY							
DES. ENG. J. H. 12-1-51							
PROJECT NO. DO-0806							
U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC							
GAS PURIFICATION PLANT PIPING ARRGT. CASE I, II & III							
BLDG NO. 202-A		INSTR. NUMBER 8504					
SK-2-2959

QTY.

PT.
NO.

DESCRIPTION

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SCOPE

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NO.	DESCRIPTION	REV. BY	DATE	APP'D BY	FOR	DATE
REVISIONS						
CLASSIFICATION		CLASSIFIED BY				
NONE		DATE 12-4-58				
DWG. NO. SK-2-2964		REV. NO. - - 0				
SCALE:		APPROVALS				
DRAWN BY R. RIVER DATE 12-3-58		BY R. R. Thichala FOR P.D.D. DATE 12-8-58				
CHECKED DATE						
ISSUED DATE						
DES. ENG. 3411-22 UNIV. 40-400						
INST. NO. 7062						
PROJECT NO. DO 0806						
U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC						
GAS PURIFICATION PLANT ELECT. ONE LINE DIAG. & PLOT PLAN CASE I & II						
DWG. NO. 202-A		REV. NO. 7201				
DWG. NO. SK-2-2964		REV. NO. 7201				

	1		2		3		4		5
	GRAMS/DAY	%	GRAMS/DAY	%	GRAMS/DAY	%	GRAMS/DAY	%	GRAMS/DAY
Xe 31	197.2	11.6	193.0	22.7	4.2	.5	4.2	.5	193.0
Xe 132	294.1	17.3	266.0	31.3	28.1	3.3	281.1	3.3	266.0
Xe 134	462.4	27.2	255.8	30.1	206.6	24.3	206.6	24.3	255.8
Xe 136	729.3	42.9	118.2	13.9	611.1	71.9	611.1	71.9	118.2
Kr (ALL ISOTOPES)	17.0	1.0	17.0	2.0	—	$\times 10^{-14}$	—	$\times 10^{-14}$	17.0
O ₂ N ₂ ETC.	75 PPM		75 PPM		—		—		75 PPM
H ₂ O									
TOTAL	1700		850		850				
VOLUME FT ³ /DAY	11		14.8		14.8		.81		.81
PRESSURE (PSIA.)	14.7		13.2		13.2		1000		1000
TEMP. (°C)	27		453		453		27		27

DECLASSIFIED

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NO.	DESCRIPTION	REV.	BY	DATE
REVISIONS		REV.	BY	DATE
CLASSIFICATION		CLASSIFIED BY		
NONE		DATE 11-10-58		
SK-2-17892		0		
SCALE:	DATE 11-10-58	APPROVALS	DATE	REV.
CHECKED	DATE	DATE	DATE	DATE
ISSUED	DATE	DATE	DATE	DATE
DES. ENG.	UNIT			
UNIT	1062			
PROJECT				
U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC				
GAS PURIFICATION PLANT ENGINEERING FLOW DIAG. THERMAL DIFFUSION COLUMN CASE III				
REV.	202-A	REV.	7001	
REFERENCE DRAWINGS		SK-2-17892		
NEXT USED ON				

DECLASSIFIED

CERAMIC GUIDE BLOCK

SECT. "D-D"

SCOPE

(6) 1" DIA HOLES ON 4" DIA. EQ. SPACED IN S.S.T. R. ONLY

CERAMIC GUIDE BLOCKS 3 @ 120° SPACING AT 6" INTERVALS

12" SCH. 160 S.S.T. PIPE MACHINE TO OD. SHOWN W/ POLISH SURFACE

B" UNIT

SECT. A-A

APPROVED
FOR DESIGN
NOT FOR CONSTRUCTION

DESCRIPTION		REV BY	APP BY	FOR	DATE
REVISIONS					
CLASSIFICATION		CLASSIFIED BY			
NONE		J. Smith 12-1-51			
DWG NO.	SK-2-2960	DESIGN NO.	00-0806	REV NO.	0
SCALE: NONE		APPROVALS			
DRAWN: (DATE) 11-7-50		CHECKED: P. Smith 12-1-51			
ISSUED: DATE		P. Smith 12-1-51			
SUB. ENG. J. Smith UNIT Design					
PROJECT NO. 00-0806					
U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC					
GAS PURIFICATION PLANT THERMAL DIFFUSION COLUMN CASE III					
DRAWING TITLE		BLDG NO.	202-A	INDEX NUMBER	2500
REFERENCE DRAWINGS		DWG NO. SK-2-2960			
NEXT USED ON

QTY

12

DESCRIPTION

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PAGE 51

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SCOPE

NO.	DESCRIPTION	REV.	BY	DATE	APPROVED BY	DATE
REVISIONS						
CLASSIFICATION		NONE				
DWS NO.		SK-2-2961				
SCALE: NONE		APPROVALS				
DRAWN BY: J. Smith		DATE: 12-1-57				
CHECKED BY: J. Smith		DATE: 12-1-57				
ISSUED BY: J. Smith		DATE: 12-1-57				
DES. ENGR. J. Smith		DATE: 12-1-57				
PROJECT NO. DO-0806		U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC				
GAS PURIFICATION PLANT PIPING ARR'G'T. CASE 3						
DWS NO. 202-A		INDEX NUMBER 8407				
DWS NO. SK-2-2961						

DWS NO. SK-2-2961
DRAWING TITLE
REFERENCE DRAWINGS

NEXT USED ON

DECLASSIFIED
DECLASSIFIED

PLAN of CELL, CONTROL ROOM & LABORATORY
SCALE $\frac{1}{4}" = 1'-0"$

SLAB
OVER

GRADE

HW-58400
PAGE 5.2

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FOR DESIGN
NOT FOR CONSTRUCTION

SCOPE

DESCRIPTION		REV	BY	DATE	APP'D	DATE
REVISIONS						
CLASSIFICATION		NONE				
PROJECT NO.		SK-2-2963				
SCALE AS SHOWN		APPROVALS				
DRAWN BY: [Signature]		CHECKED BY: [Signature]				
DATE: [Date]		DATE: [Date]				
REVISED BY: [Signature]		REVISED BY: [Signature]				
DATE: [Date]		DATE: [Date]				
DES. ENG. [Signature]		DATE: [Date]				
12128						
DO-0806						
U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC						
GAS PURIFICATION PLANT ARCH. PLAN & SECTIONS CASE III						
202 A		0800.01				
SK-2-2963

HW-58400
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SCOPE

APPROVED
FOR DESIGN
NOT FOR CONSTRUCTION

DESCRIPTION		REV BY	DATE	APP'D BY	FOR	DATE
REVISIONS						
SK-2-2962						
CLASSIFICATION		REVIEWED BY				
NONE		DATE 10/10/50				
SCALE: 1" = 20'-0"		APPROVALS				
DRAWN BY DATE 10-3-50		BY	FOR	DATE		
CHECKED DATE		BY	FOR	DATE		
ISSUED DATE		BY	FOR	DATE		
DES. ENG. DATE		BY	FOR	DATE		
REV. 12090		BY	FOR	DATE		
PROJECT NO. DO-0806		BY	FOR	DATE		
U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC &						
GAS PURIFICATION PLANT PLOT PLAN CASE I, II & III						
202-A		SK-2-2962		BY	FOR	DATE

DRAWING NO.	DRAWING TITLE
	REFERENCE DRAWINGS
NEXT USED ON	

DET. 3 H-2-57043
LEVEL REEL
6151 REV. 4 &
REV. 4

ARRICADE HWSTD. E-5-28 &
NON SIGNS

10 0110-01 11

HW-58400
PAGE 574

SERVICE
CENTER
E III

DUCT SCHEDULE
(ALL CONDUIT IS FIBRE DUCT)
A. - 4" - 3/4" 1/2" 15KV SHLD OR NEUT-GE SI-58040
B. DO
C. DO
D. DO
E. DO
F, G, H. - SPARE.
J. - 4" - 3/4" 1/2" 15KV SHLD. ETC.
K. - SPARE.

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FOR DESIGN
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SCOPE

SECT. B-B
N.T.S.

SECT. A-A
N.T.S.

NO.	DESCRIPTION	REV.	DATE	APP'D.	DATE	FOR	DATE
REVISIONS							
CLASSIFICATION		NONE					
DATE		2/24/59					
SK-2-2965							
SCALE: AS NOTED		APPROVALS					
DRAWN		DATE		BY		FOR	
CHECKED		DATE		BY		FOR	
ISSUED		DATE		BY		FOR	
DES. ENG.		DATE		BY		FOR	
UNIT		DATE		BY		FOR	
PROJECT		DATE		BY		FOR	
U. S. ATOMIC ENERGY COMMISSION HANFORD ATOMIC PRODUCTS OPERATION GENERAL ELECTRIC							
GAS PURIFICATION PLANT ELECT ONE LINE DIAG. & PLOT PLAN - CASE I, II & III							
REV. 202-A		REV. 7201					
SK-2-2965

AIIM

Association for Information and Image Management

1100 Wayne Avenue, Suite 1100
Silver Spring, Maryland 20910
301/587-8202

Centimeter

Inches

MANUFACTURED TO AIIM STANDARDS
BY APPLIED IMAGE, INC.

DATE

FILMED

12 / 14 / 94

END