

## AR TARGET SHEET

The following document was too large to scan as one unit, therefore it has been broken down into sections.

DOCUMENT # SD-WM-EV-100, Rev 0

EDMC # 42494

SECTION 2 OF 2

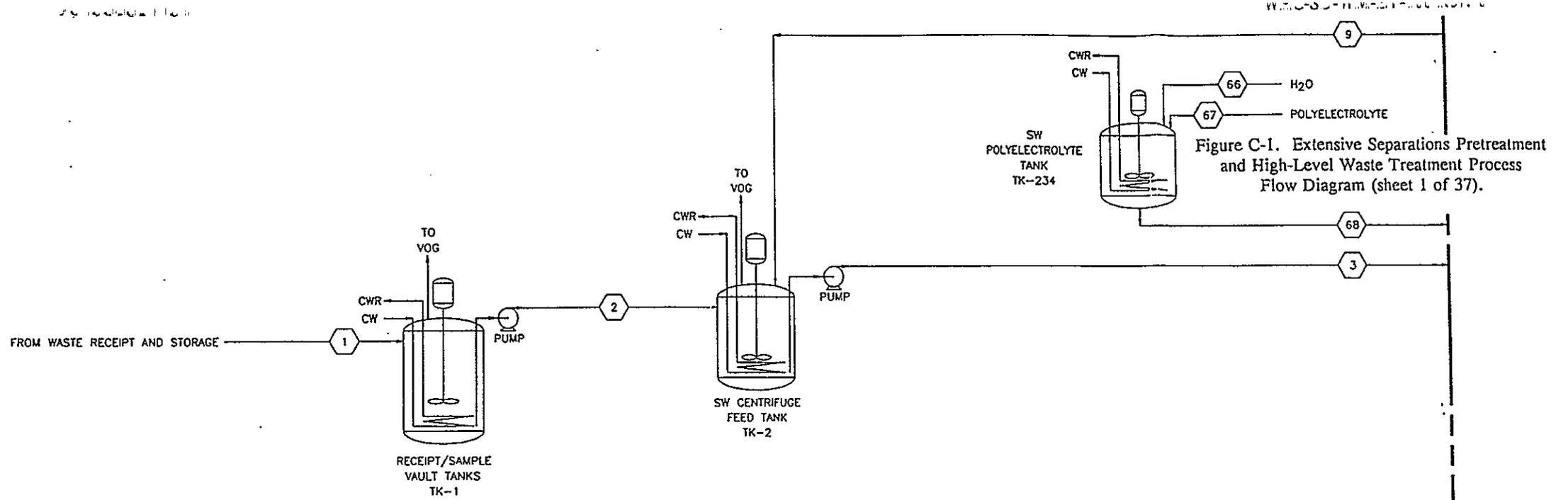


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 1 of 37).

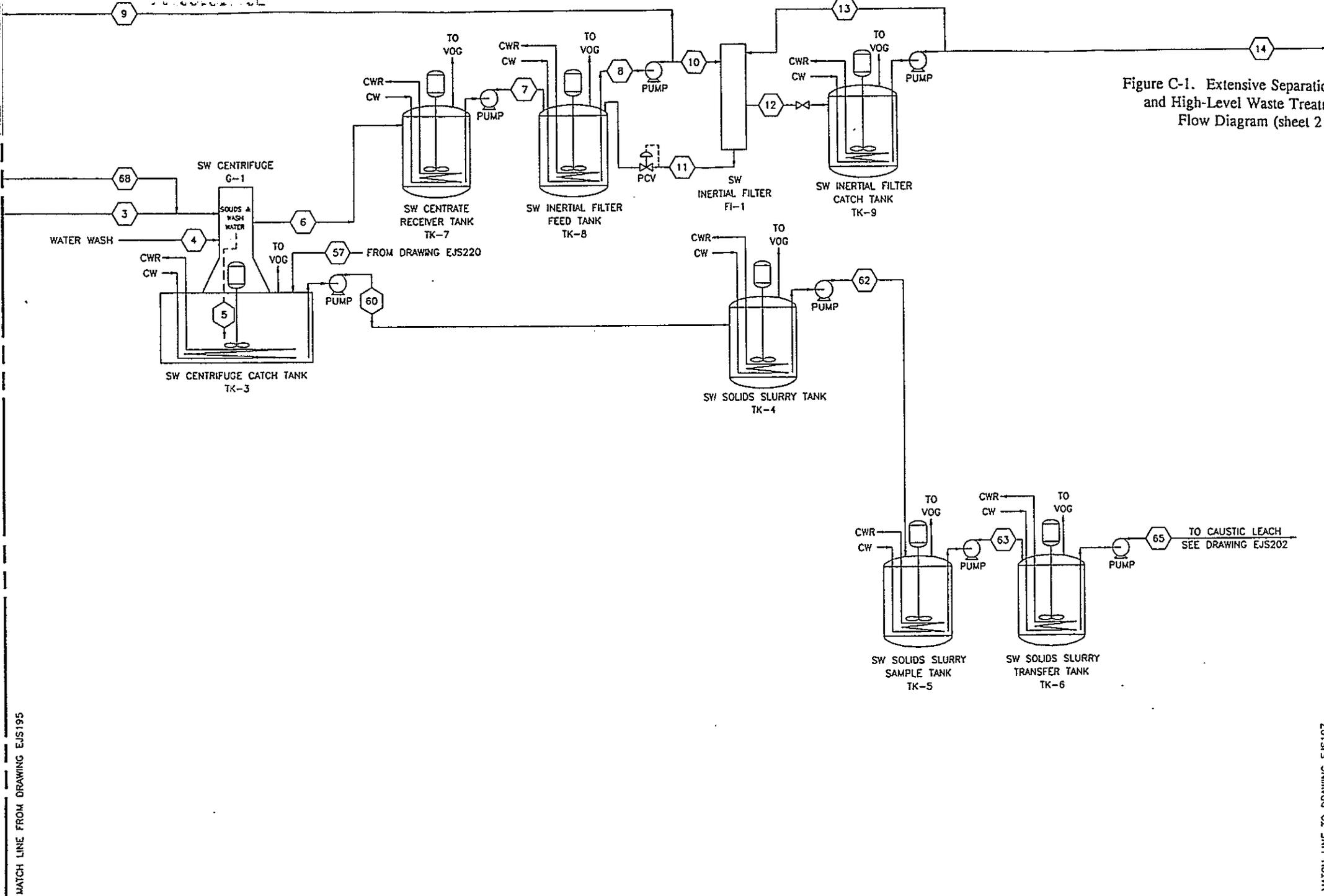


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 2 of 37).

MATCH LINE FROM DRAWING EJS195

MATCH LINE TO DRAWING EJS197

MATCH LINE FROM DRAWING EJS196

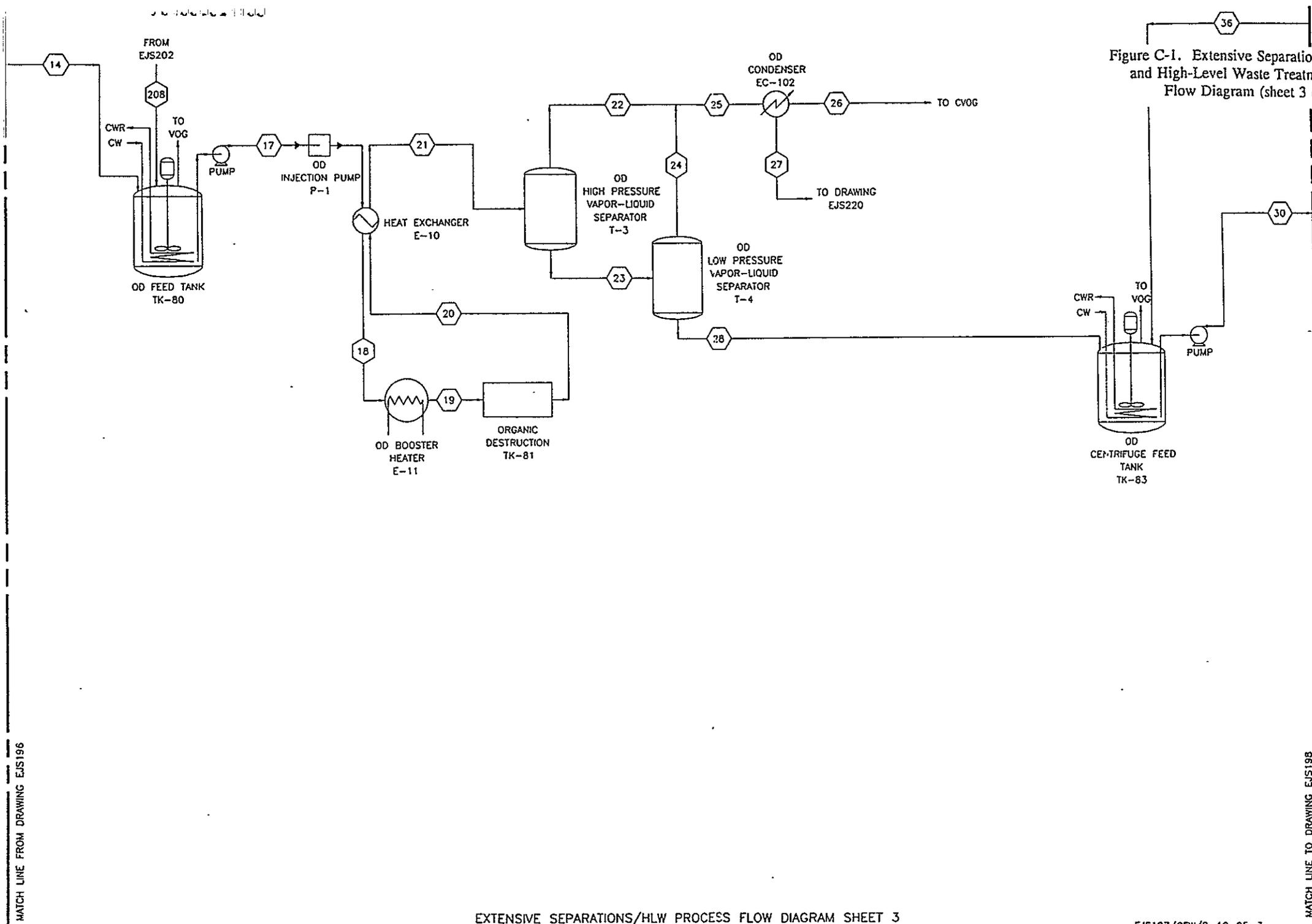


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 3 of 37).

MATCH LINE TO DRAWING EJS198

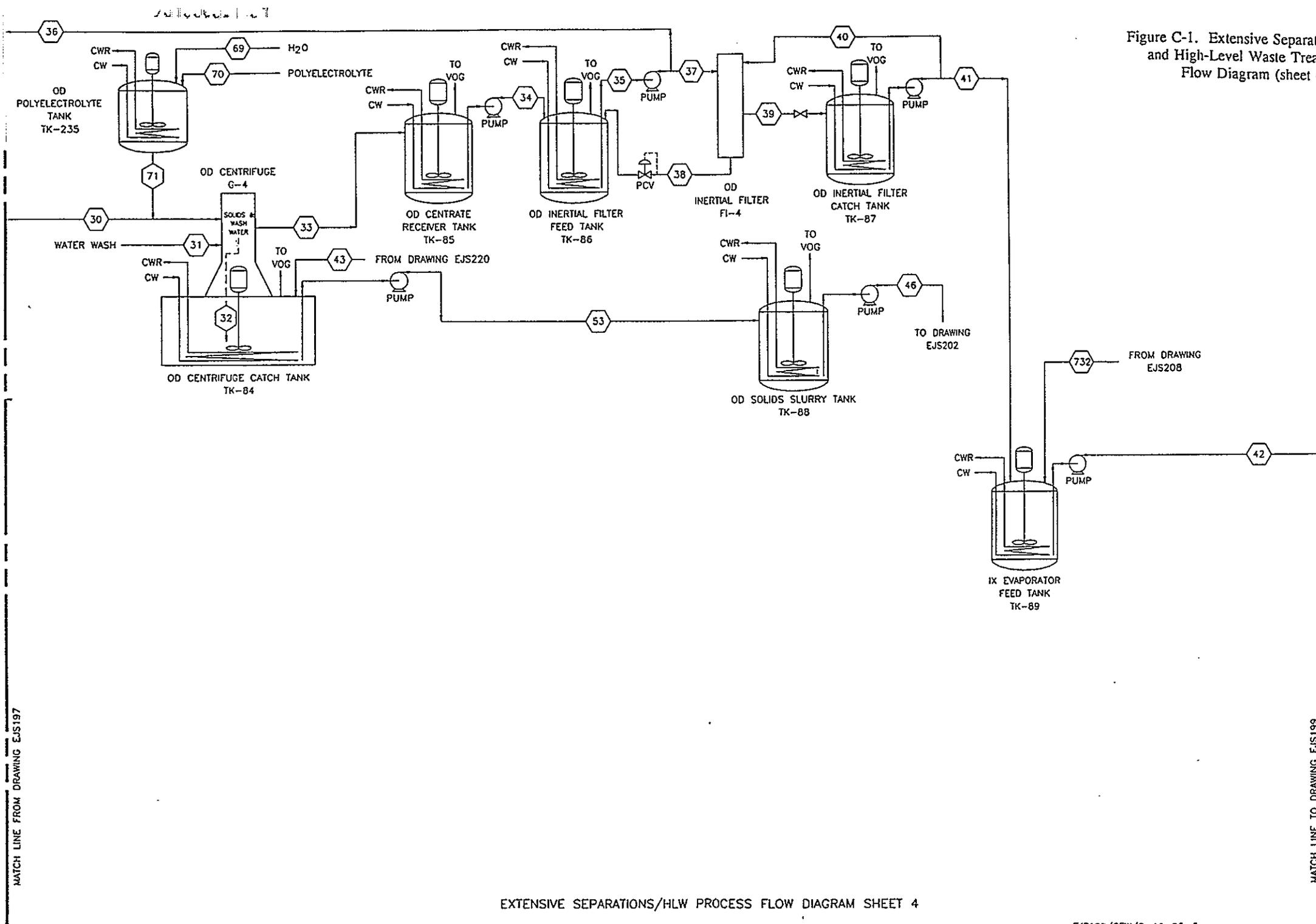
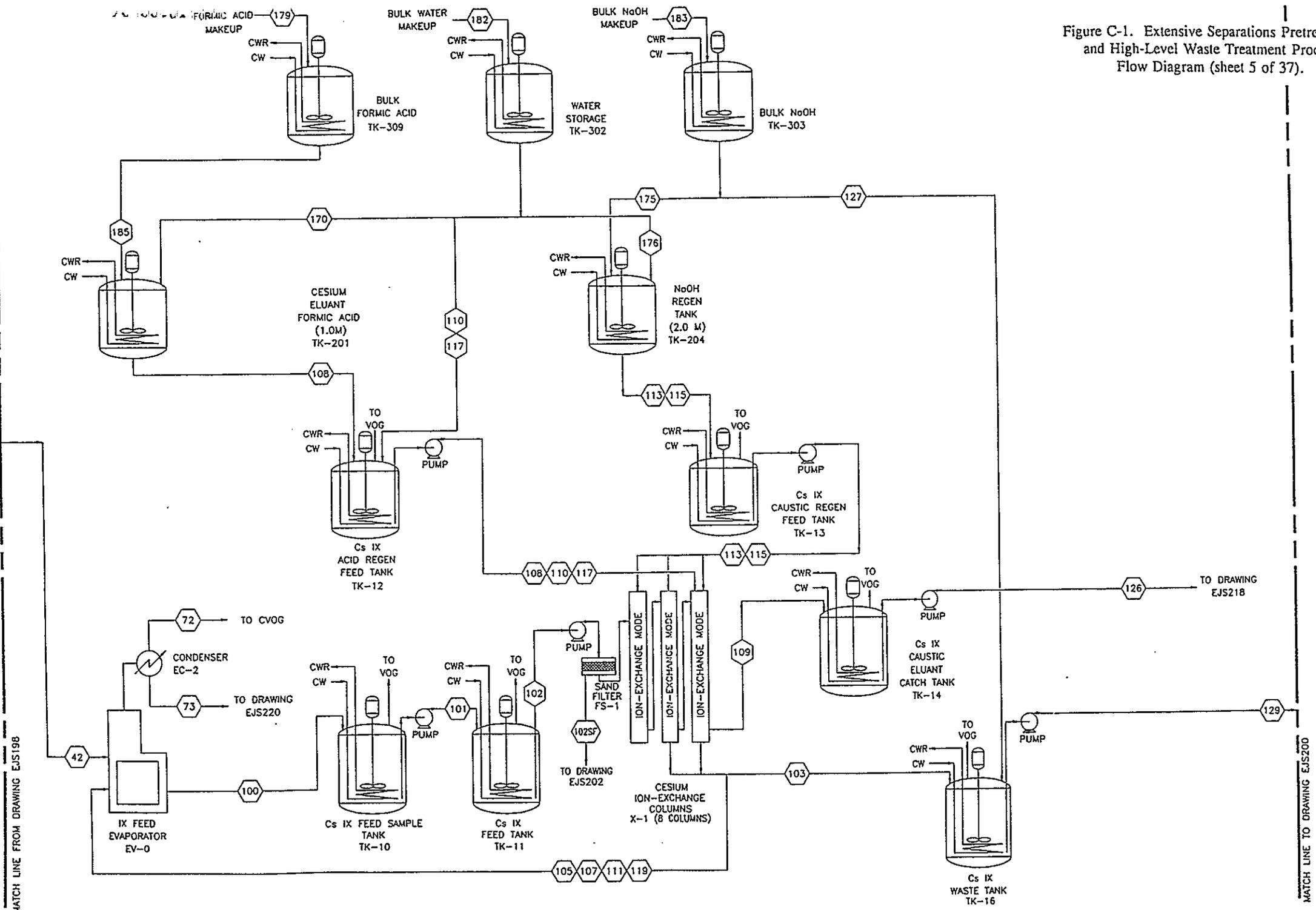


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 4 of 37).

MATCH LINE FROM DRAWING EJS197

MATCH LINE TO DRAWING EJS199

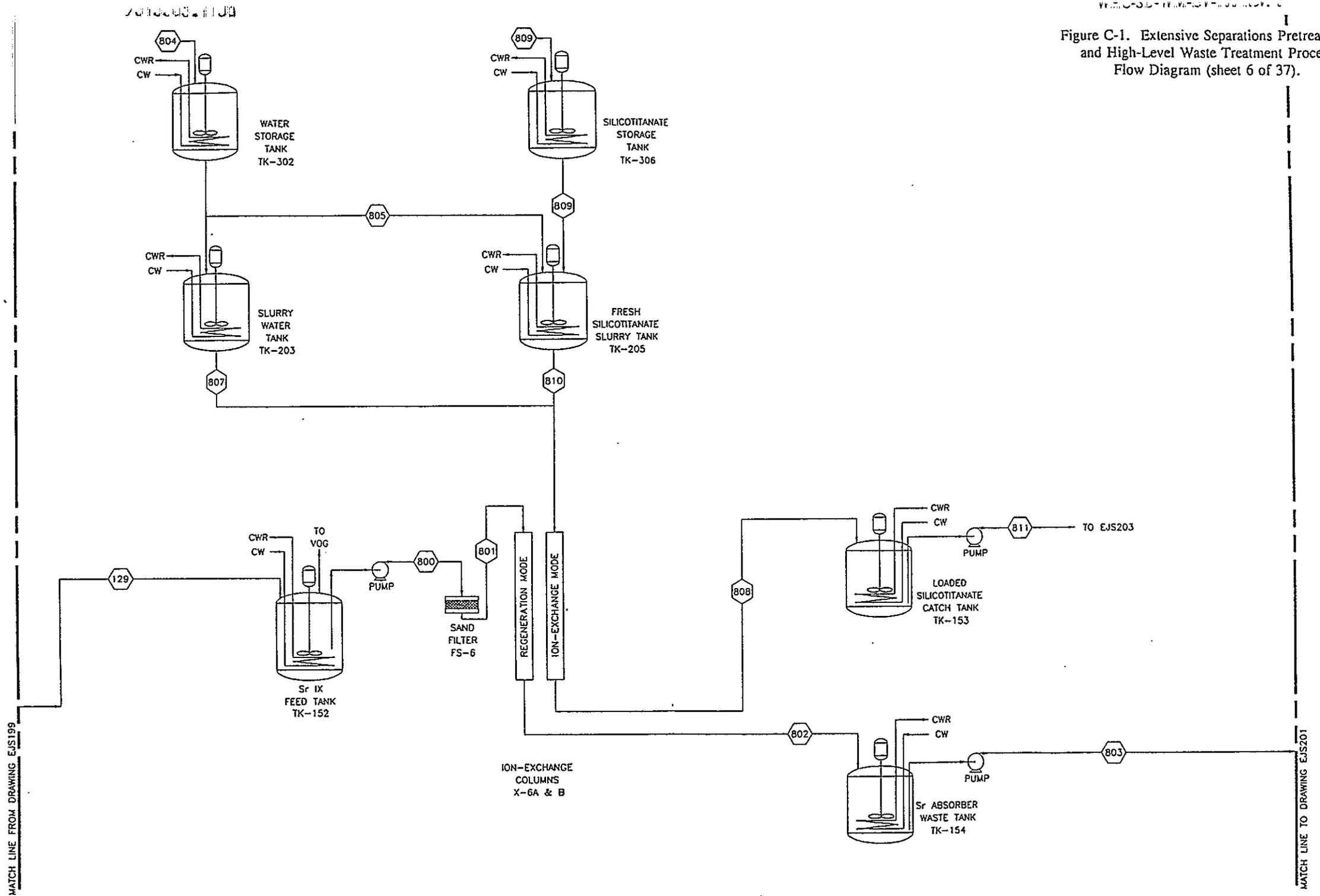
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 5 of 37).



MATCH LINE FROM DRAWING EJS198

MATCH LINE TO DRAWING EJS200

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 6 of 37).



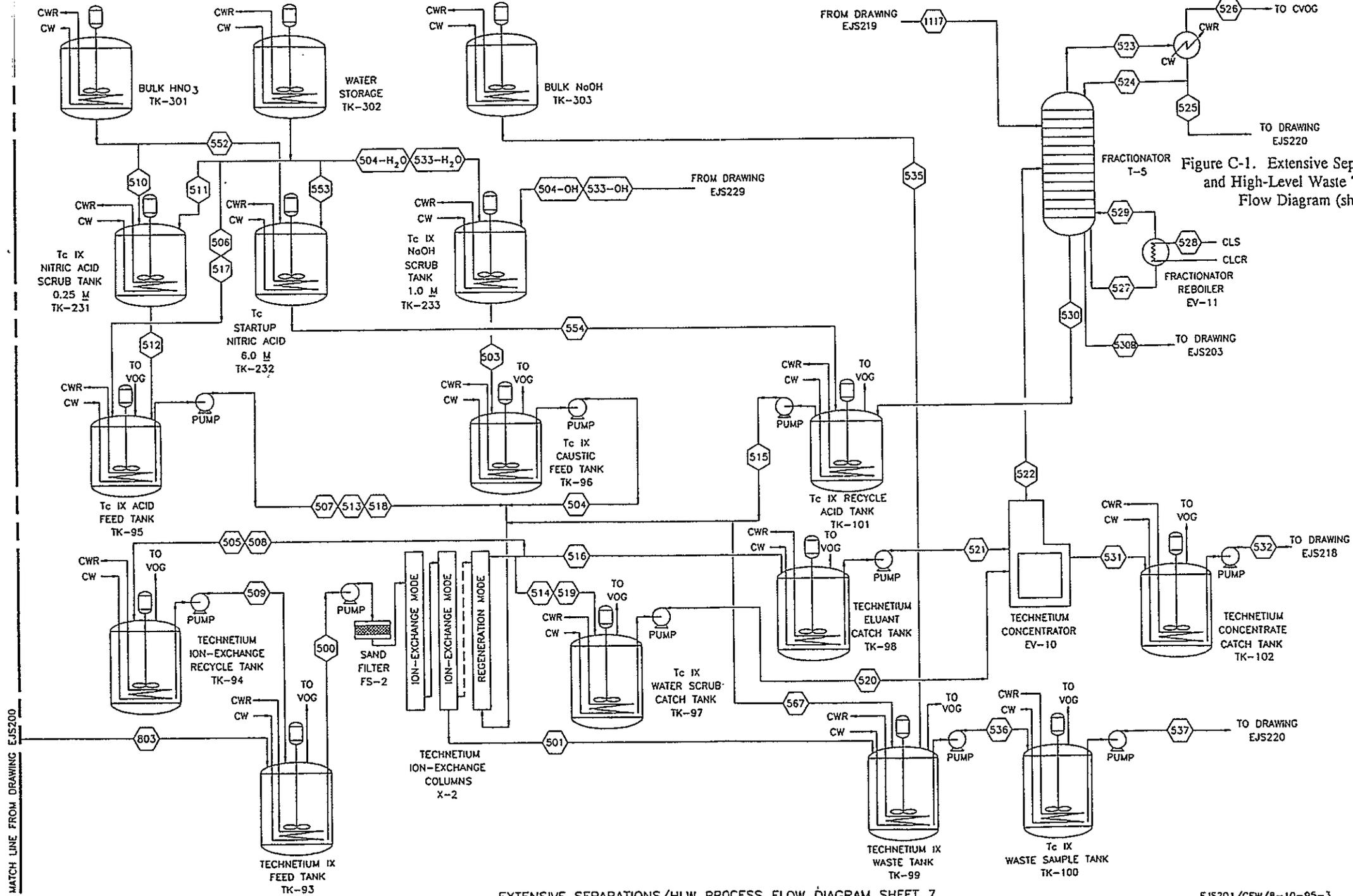
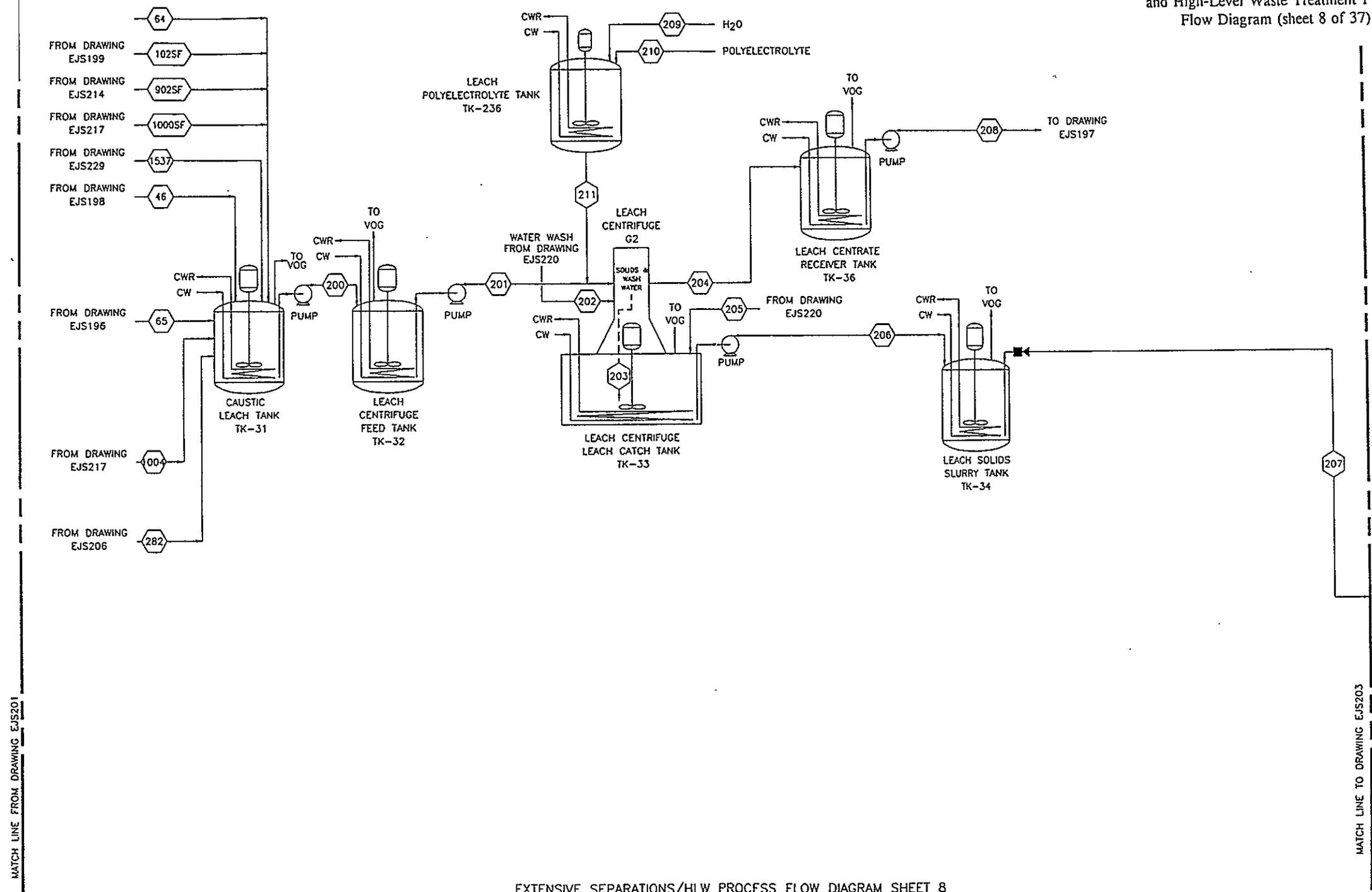


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 7 of 37).

MATCH LINE FROM DRAWING EJS200

MATCH LINE TO DRAWING EJS202

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 8 of 37).



MATCH LINE FROM DRAWING EJS201

MATCH LINE TO DRAWING EJS203

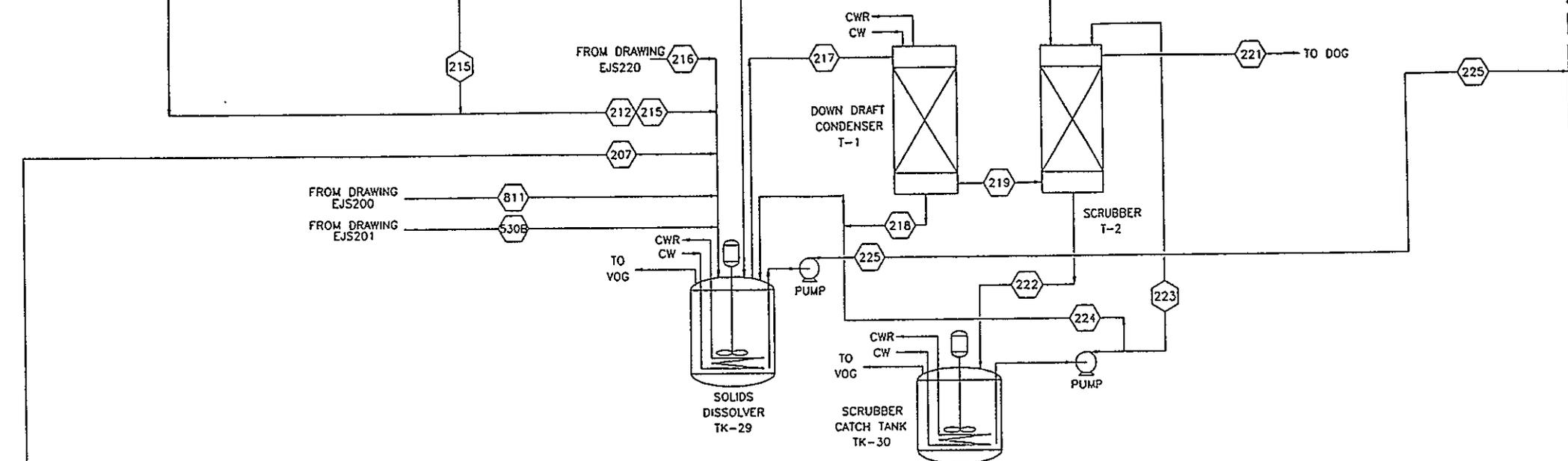
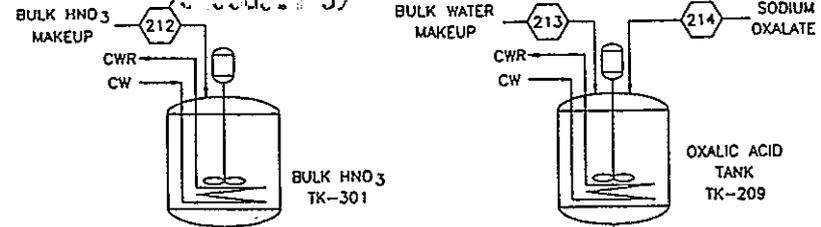
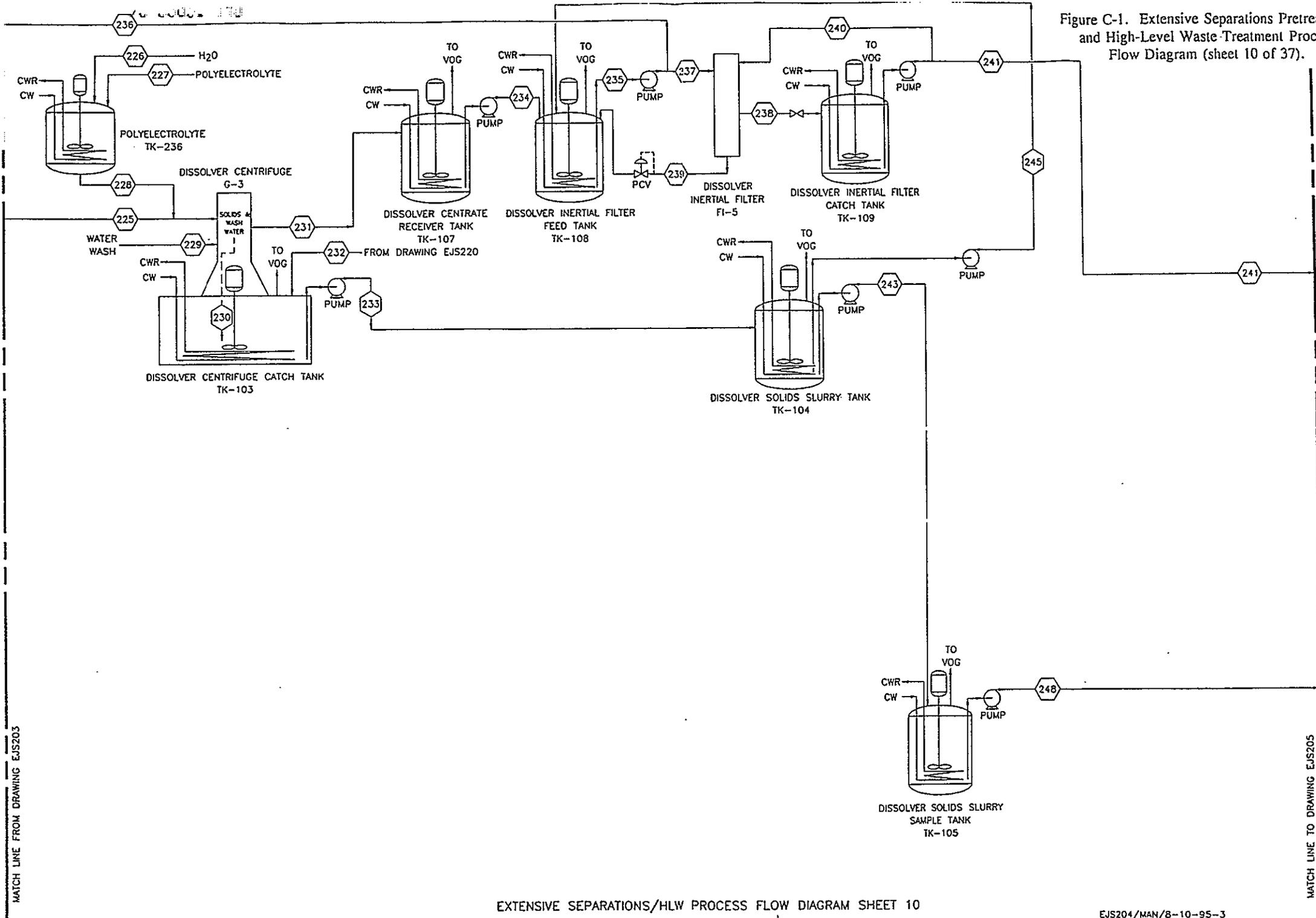


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 9 of 37).

MATCH LINE FROM DRAWING EJS202

MATCH LINE TO DRAWING EJS204

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 10 of 37).



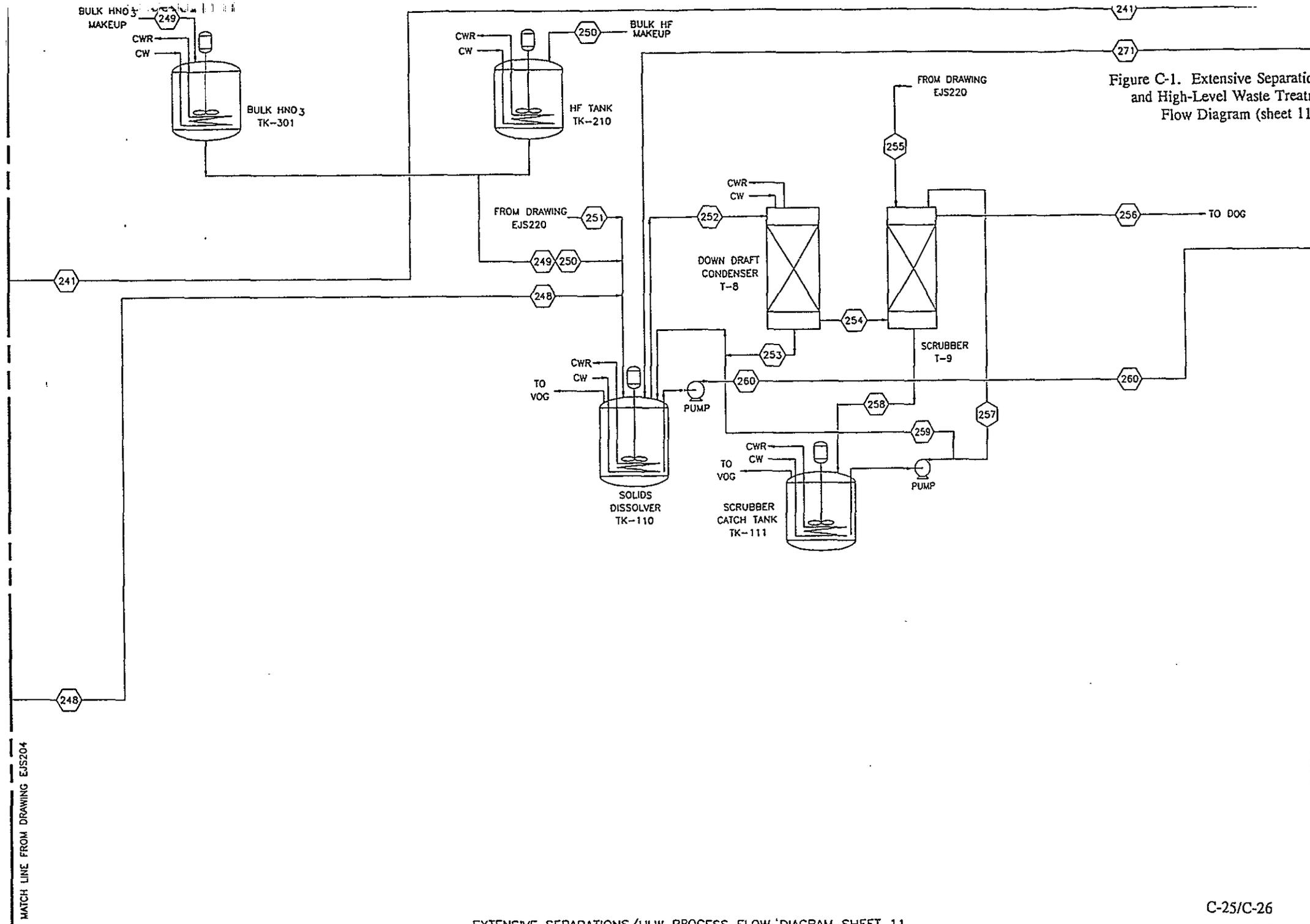
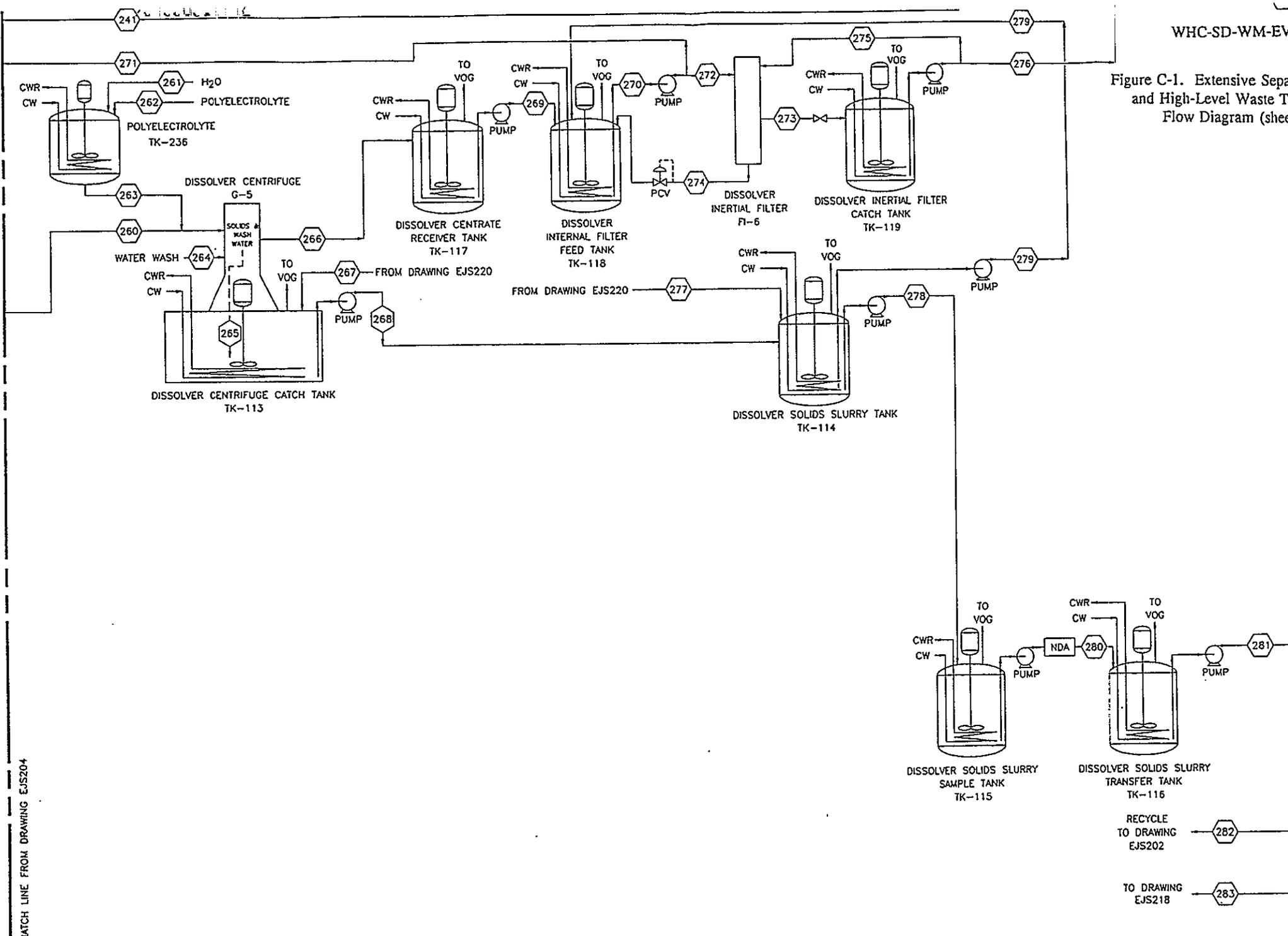


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 11 of 37).

MATCH LINE FROM DRAWING EJS204

MATCH LINE TO DRAWING EJS206

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 12 of 37).



MATCH LINE FROM DRAWING EJS204

MATCH LINE TO DRAWING EJS207

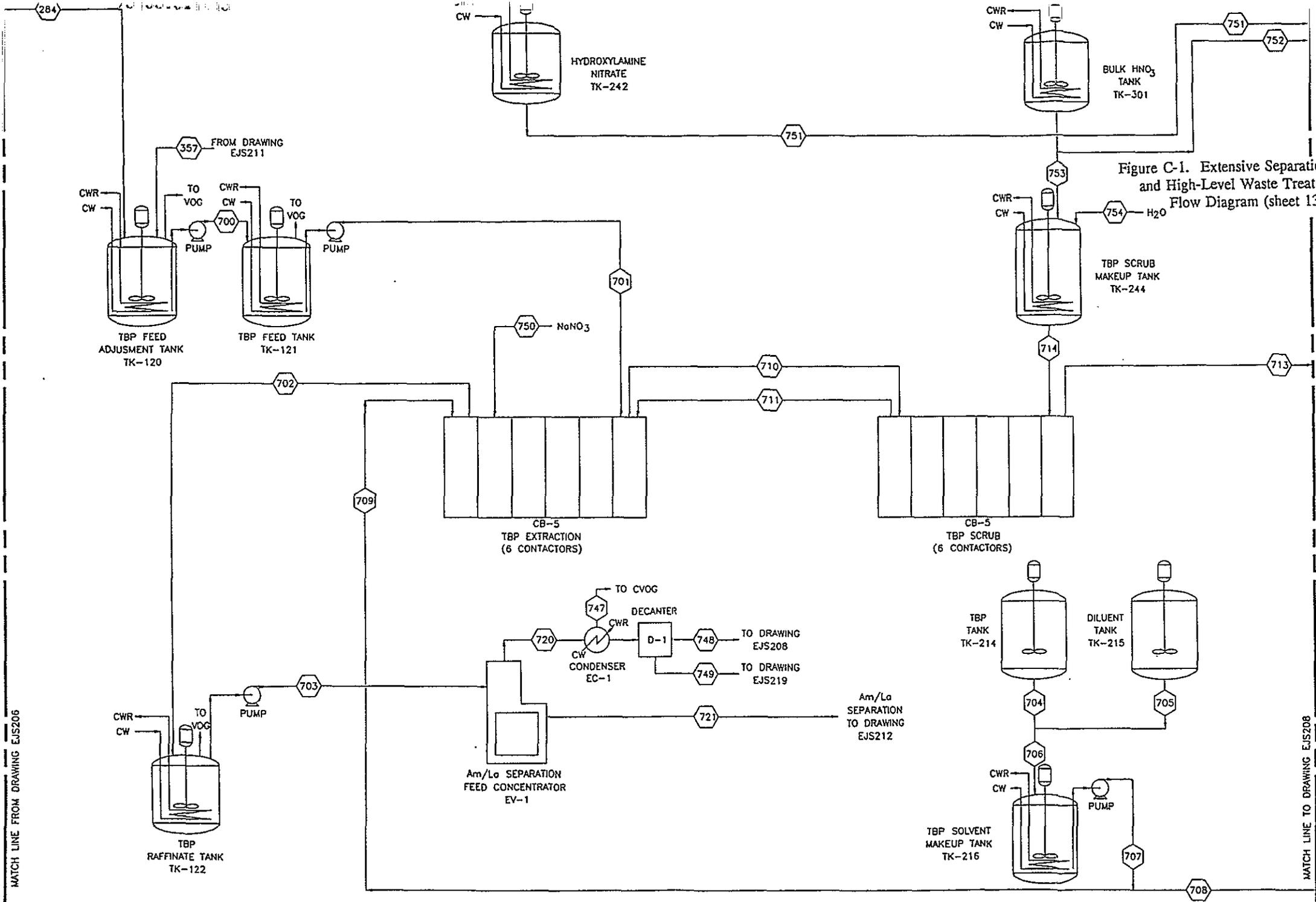


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 13 of 37).

MATCH LINE FROM DRAWING EJS206

MATCH LINE TO DRAWING EJS208

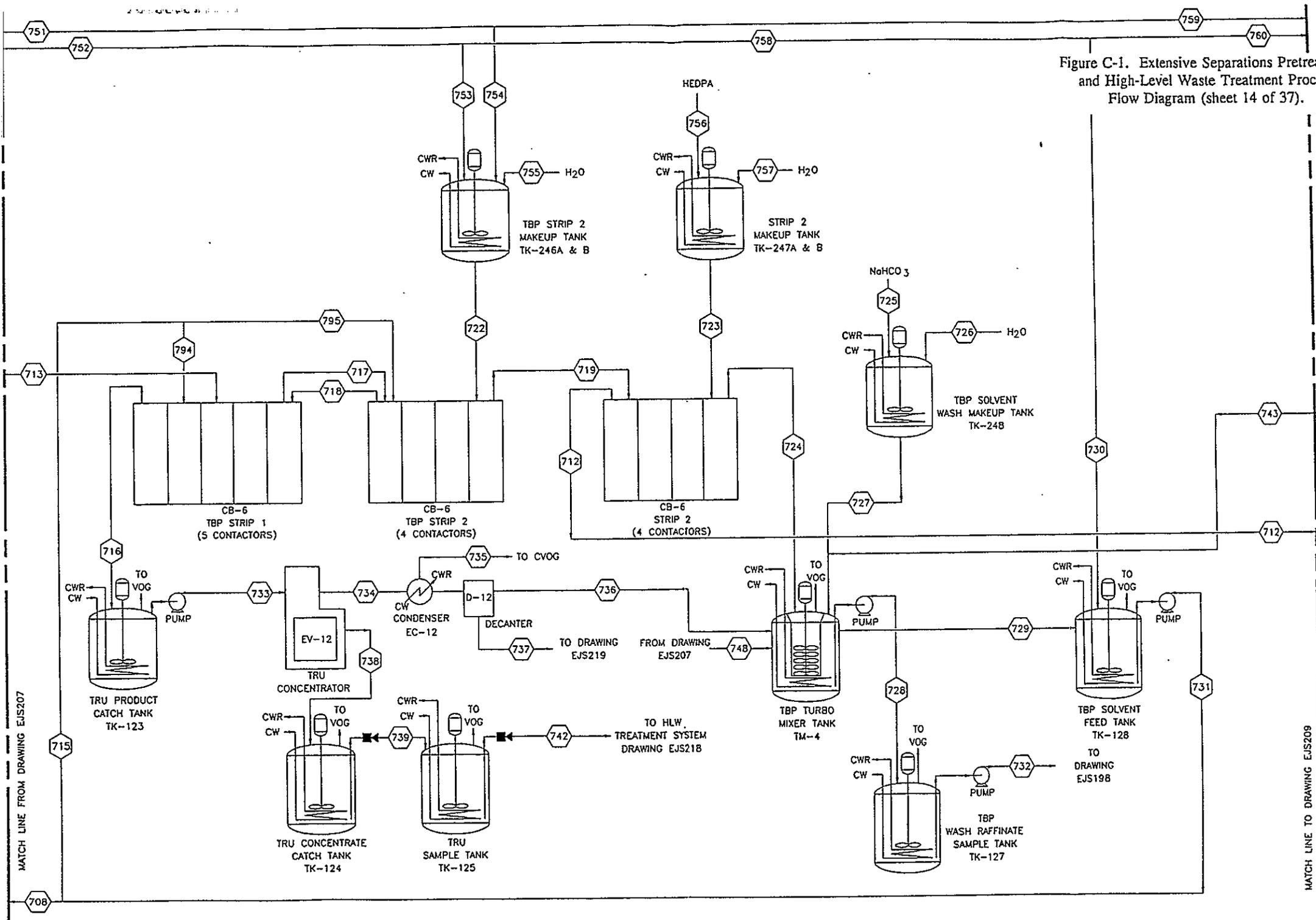


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 14 of 37).

MATCH LINE FROM DRAWING EJS207

MATCH LINE TO DRAWING EJS209

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 15 of 37).

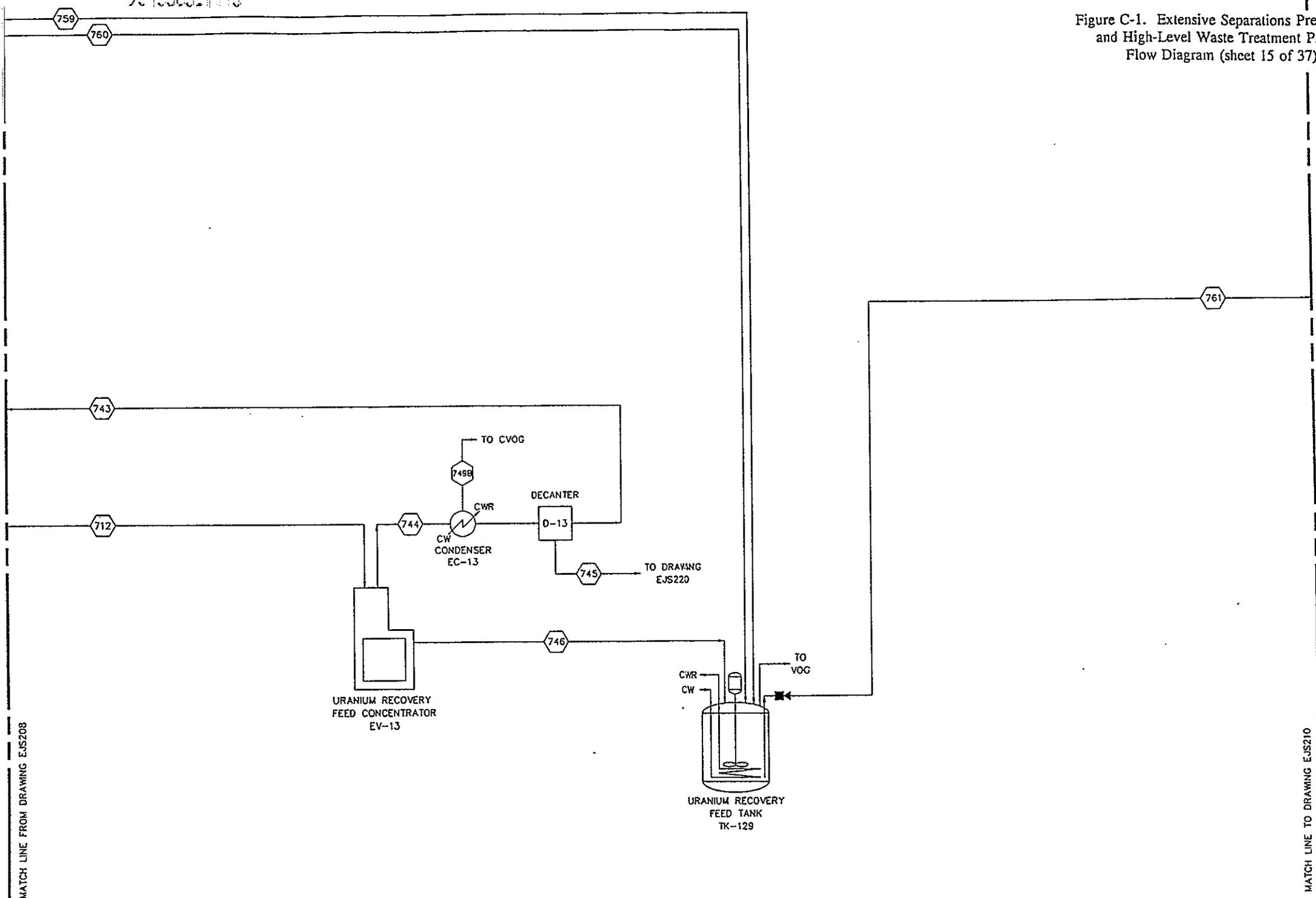


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 16 of 37)

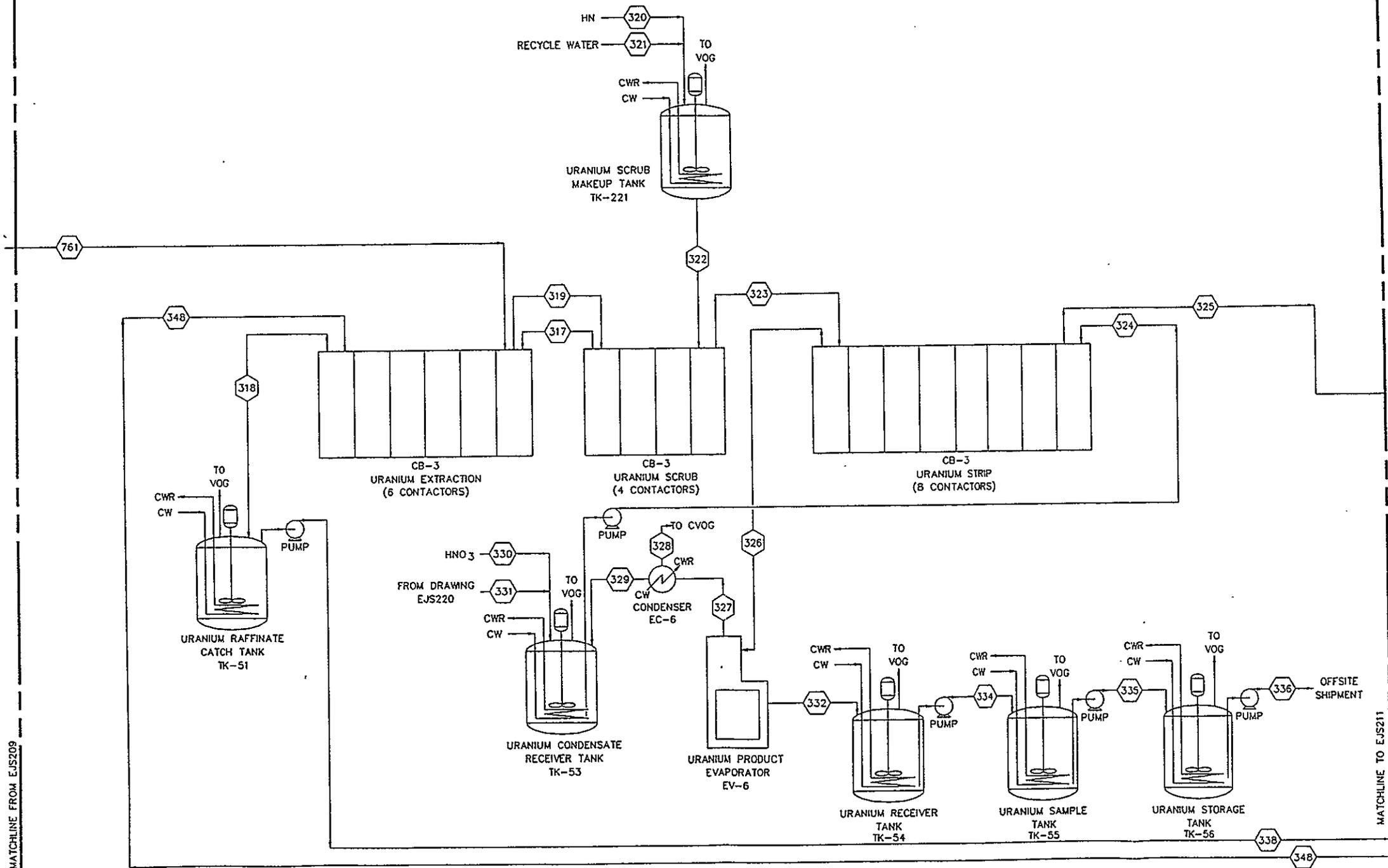
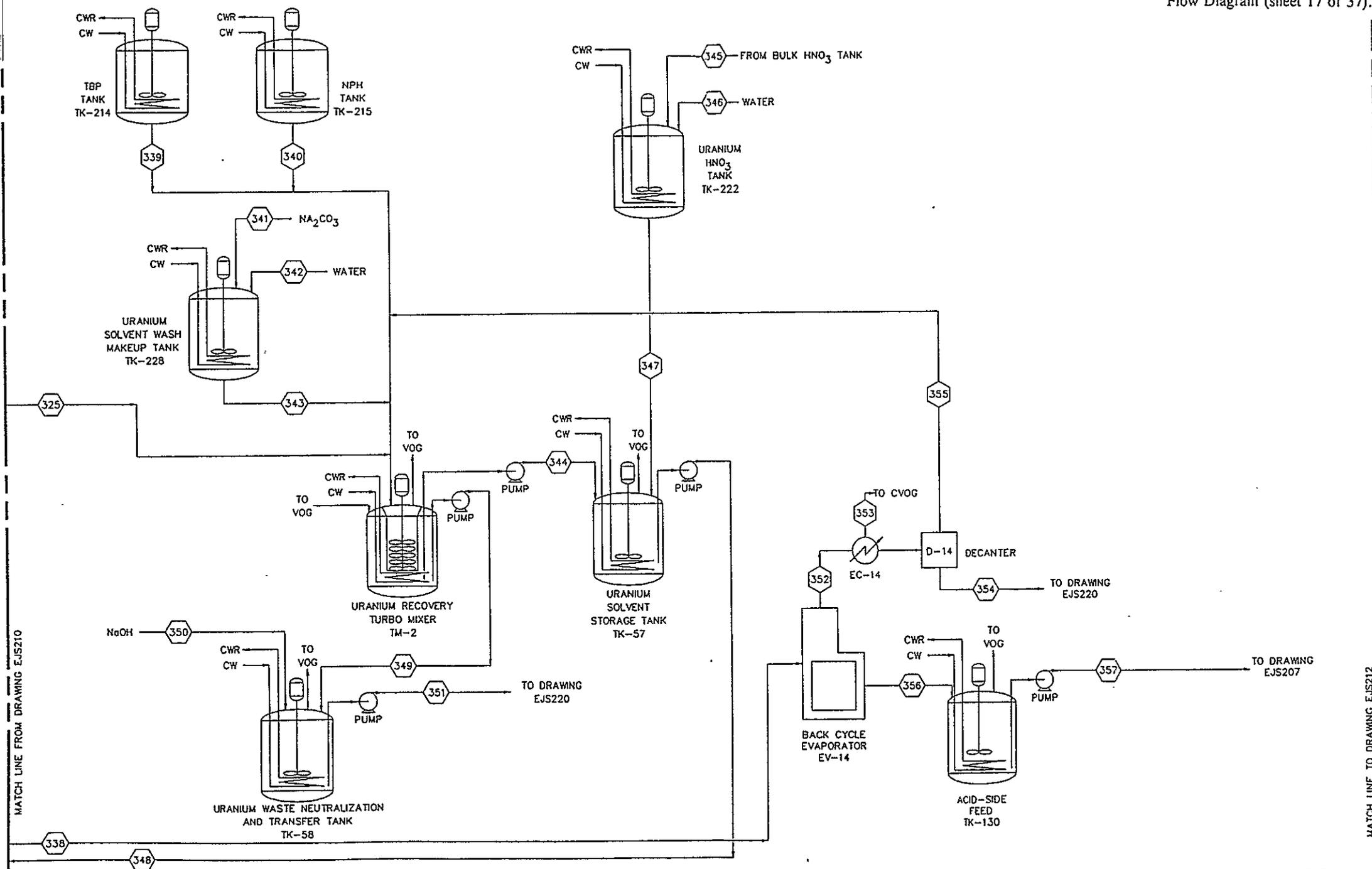


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 17 of 37).



MATCH LINE FROM DRAWING EJS210

MATCH LINE TO DRAWING EJS212

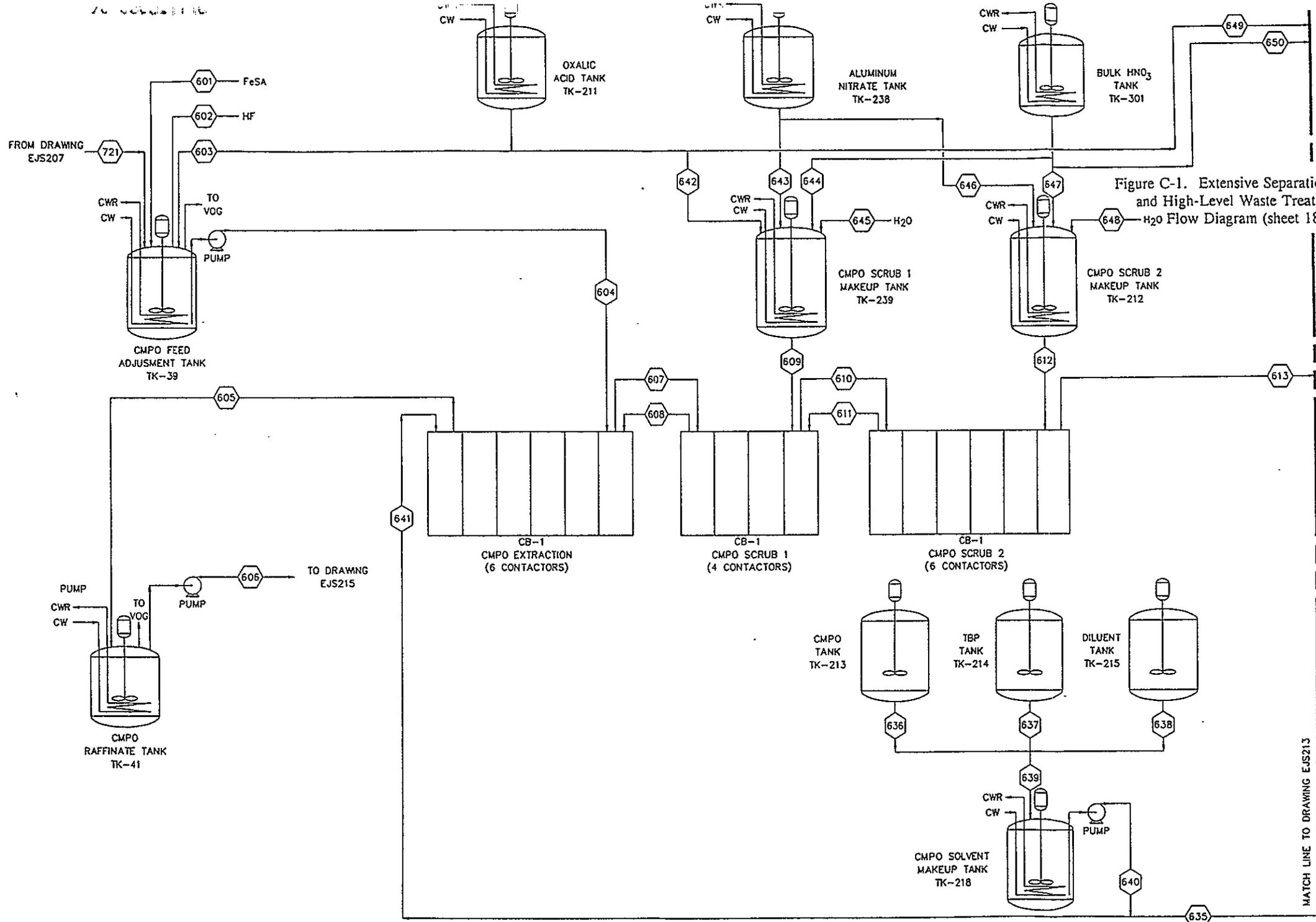
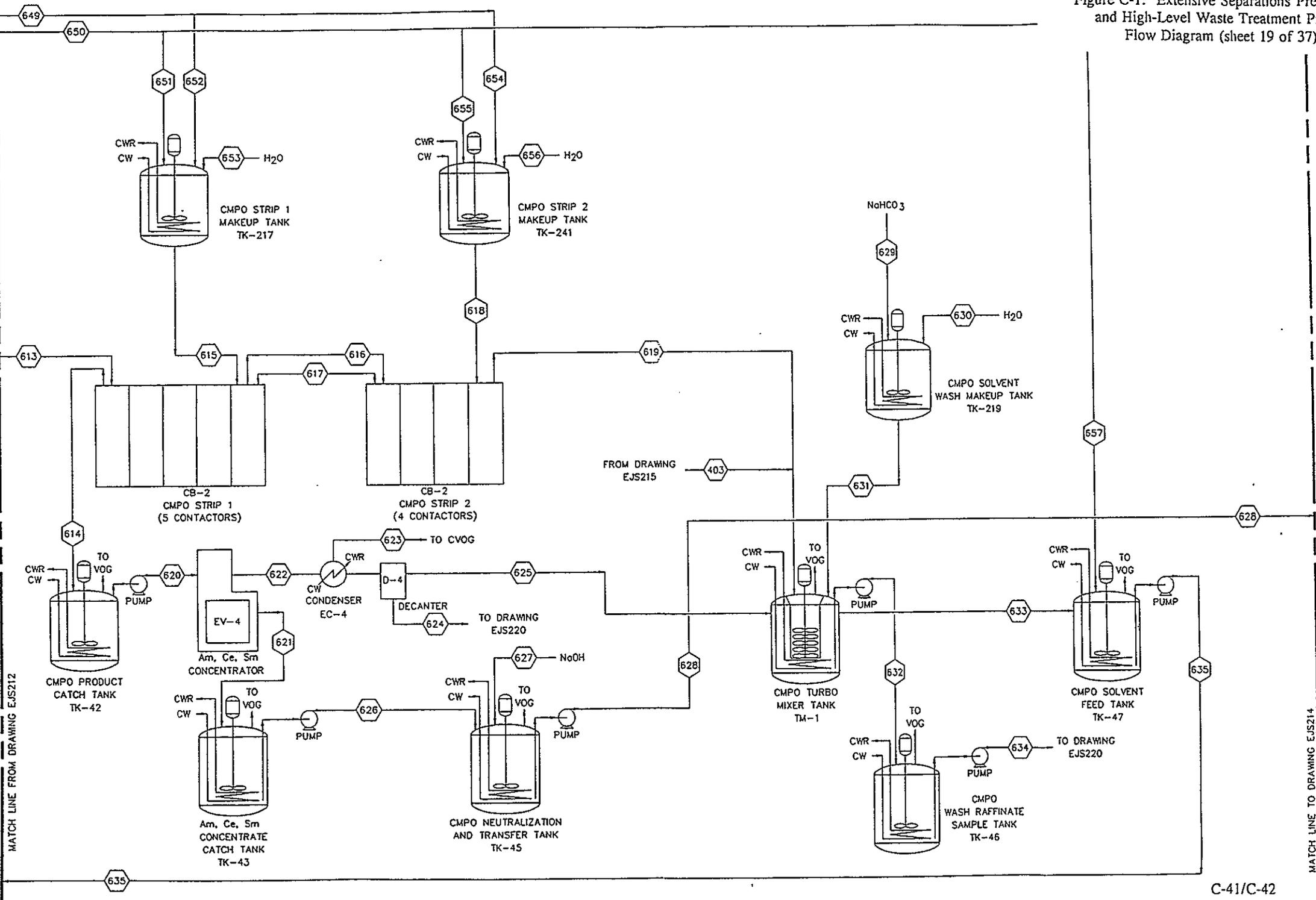


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process H<sub>2</sub>O Flow Diagram (sheet 18 of 37).

MATCH LINE FROM DRAWING EJS211

MATCH LINE TO DRAWING EJS213

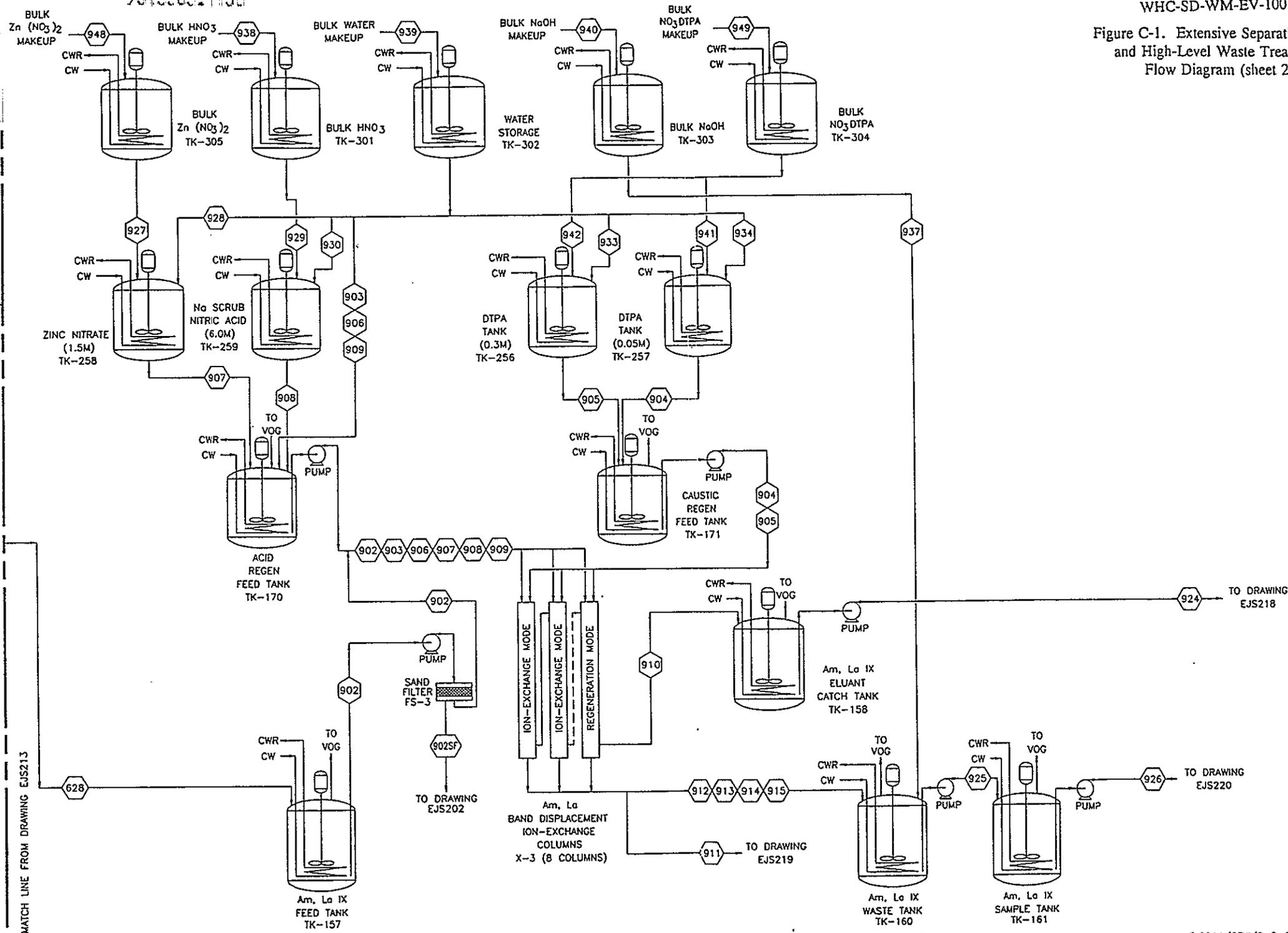
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 19 of 37).



MATCH LINE FROM DRAWING EJS212

MATCH LINE TO DRAWING EJS214

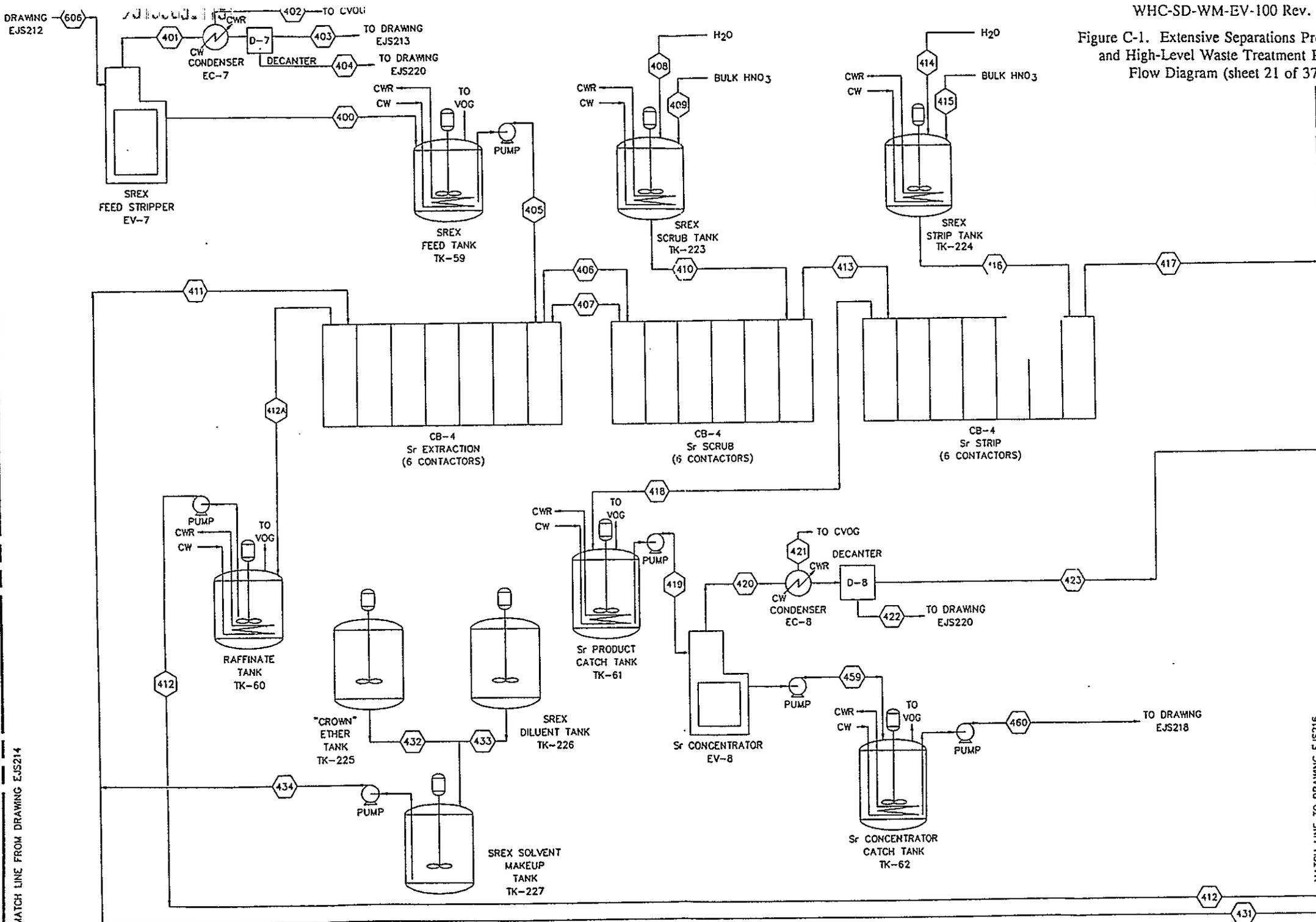
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 20 of 37).



MATCH LINE FROM DRAWING EJS213

CONTINUED ON DRAWING EJS215

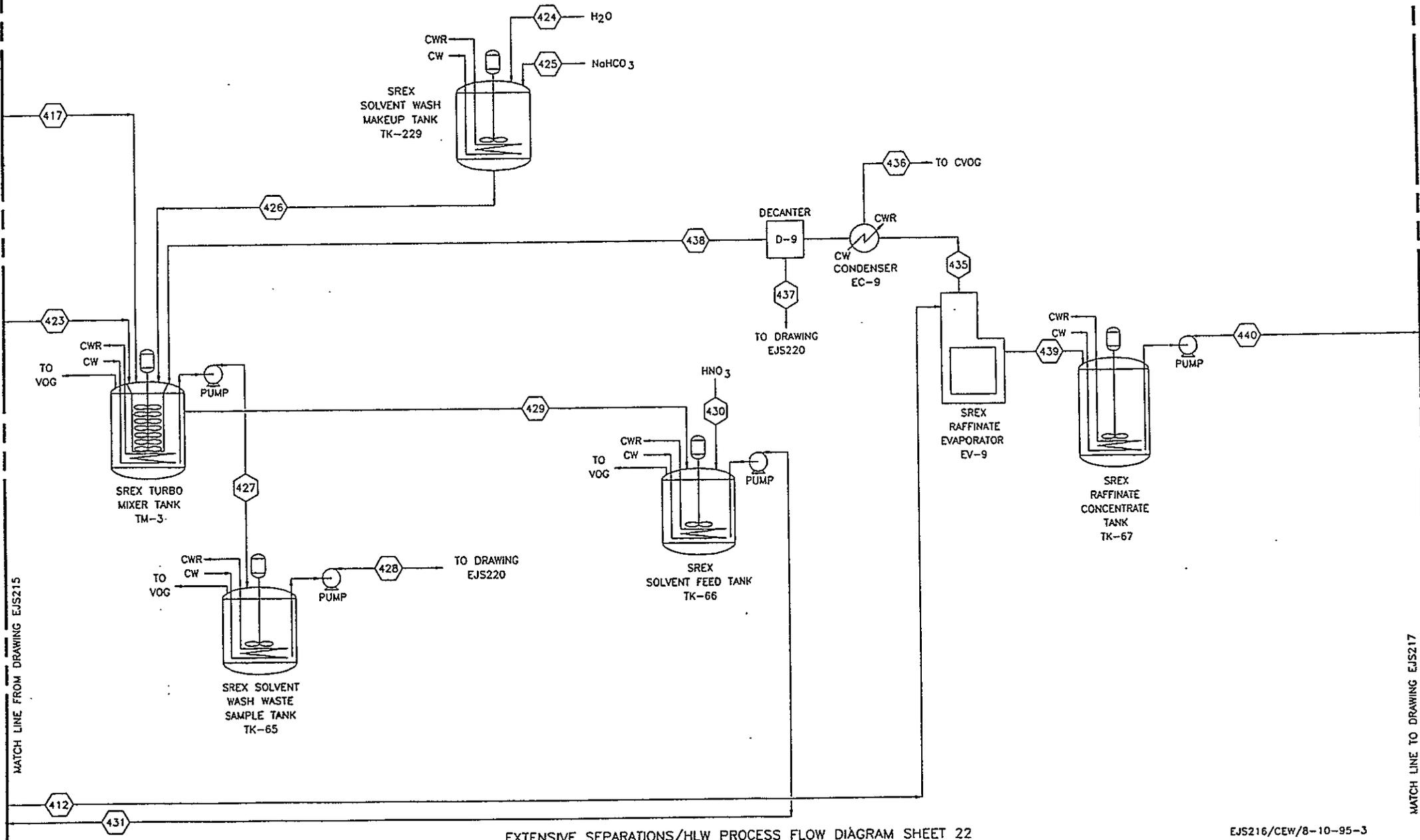
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 21 of 37).



MATCH LINE FROM DRAWING EJS214

MATCH LINE TO DRAWING EJS216

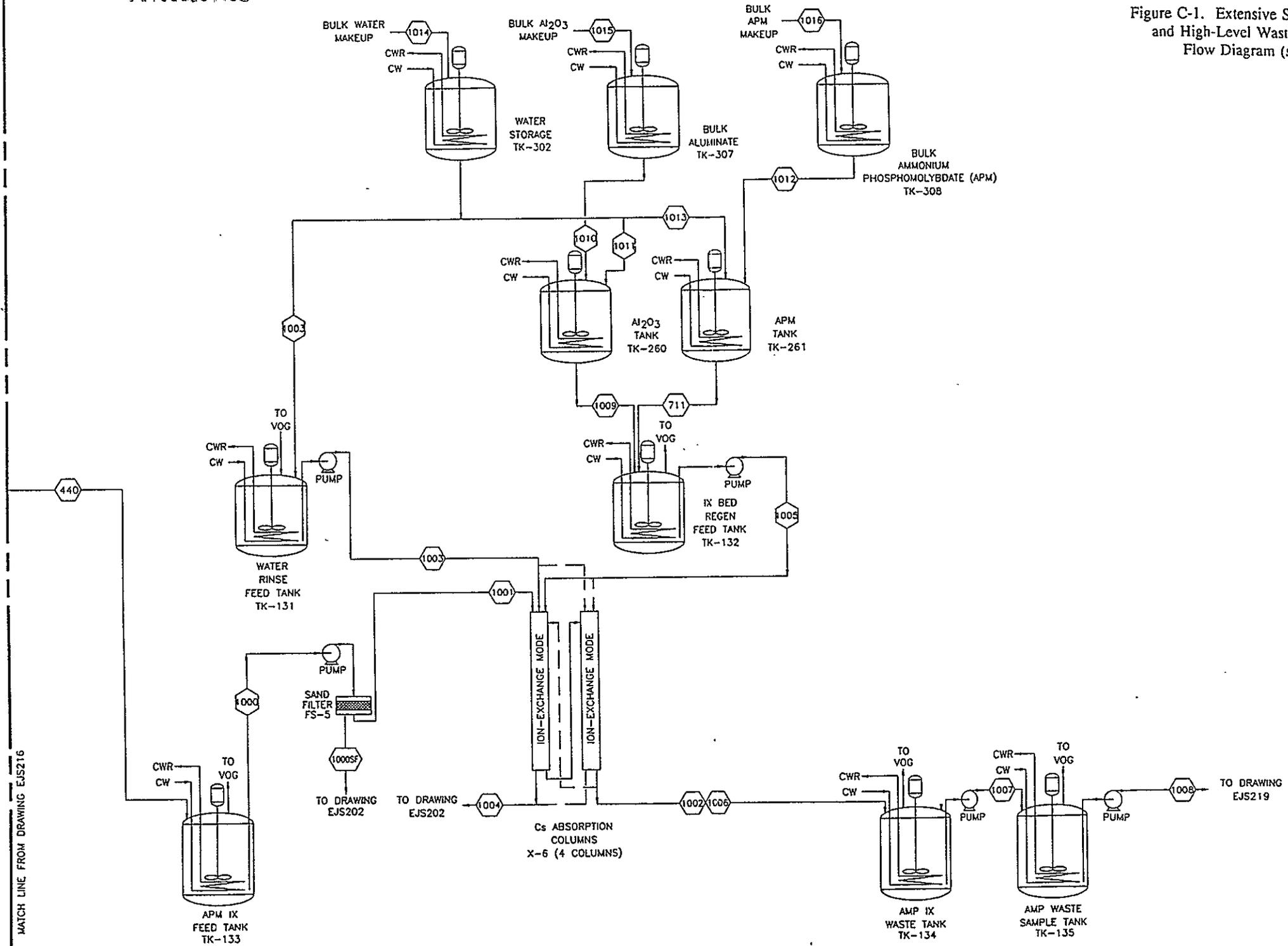
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 22 of 37).



MATCH LINE FROM DRAWING EJS215

MATCH LINE TO DRAWING EJS217

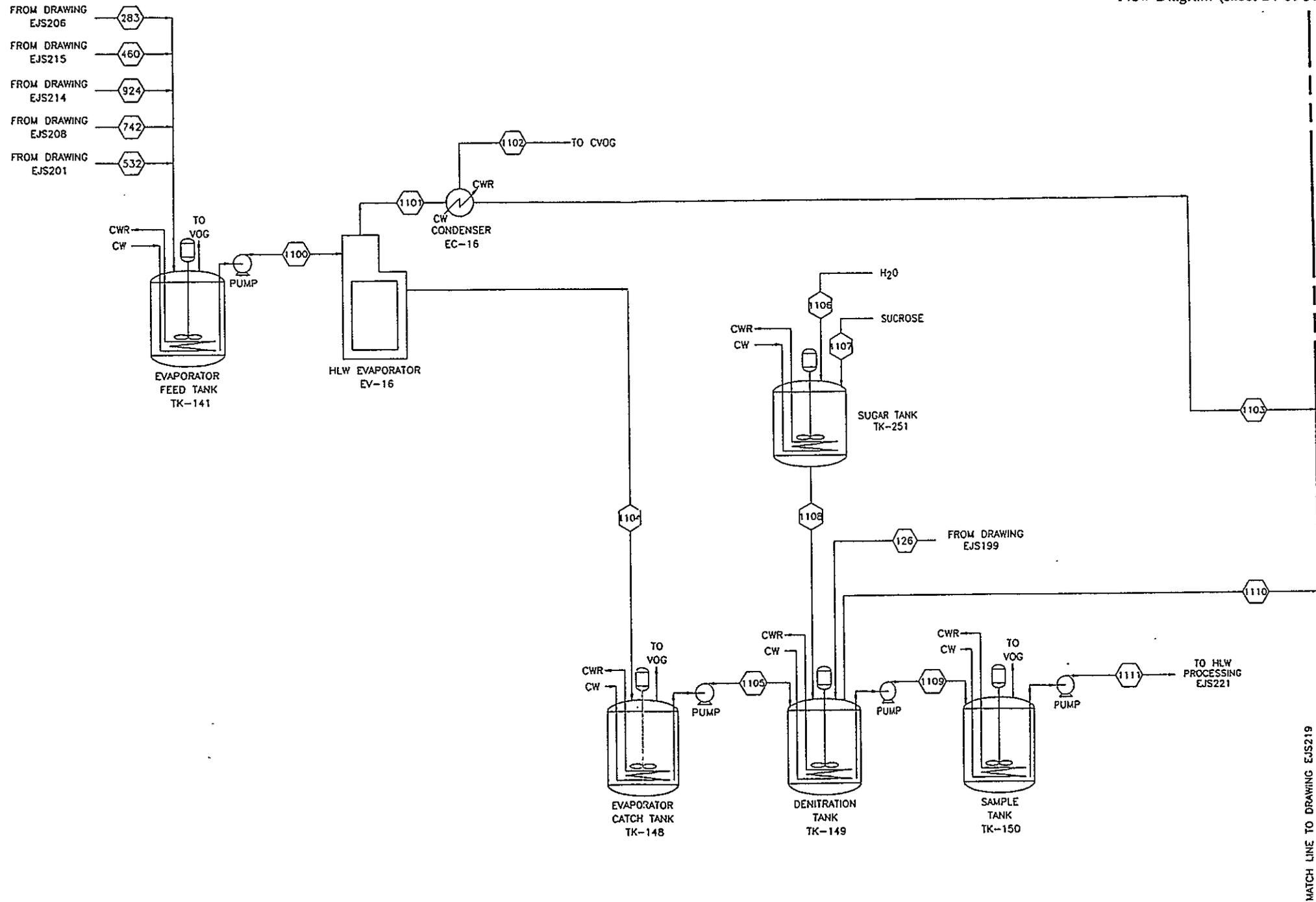
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 23 of 37).



MATCH LINE FROM DRAWING EJS216

CONTINUED ON DRAWING EJS218

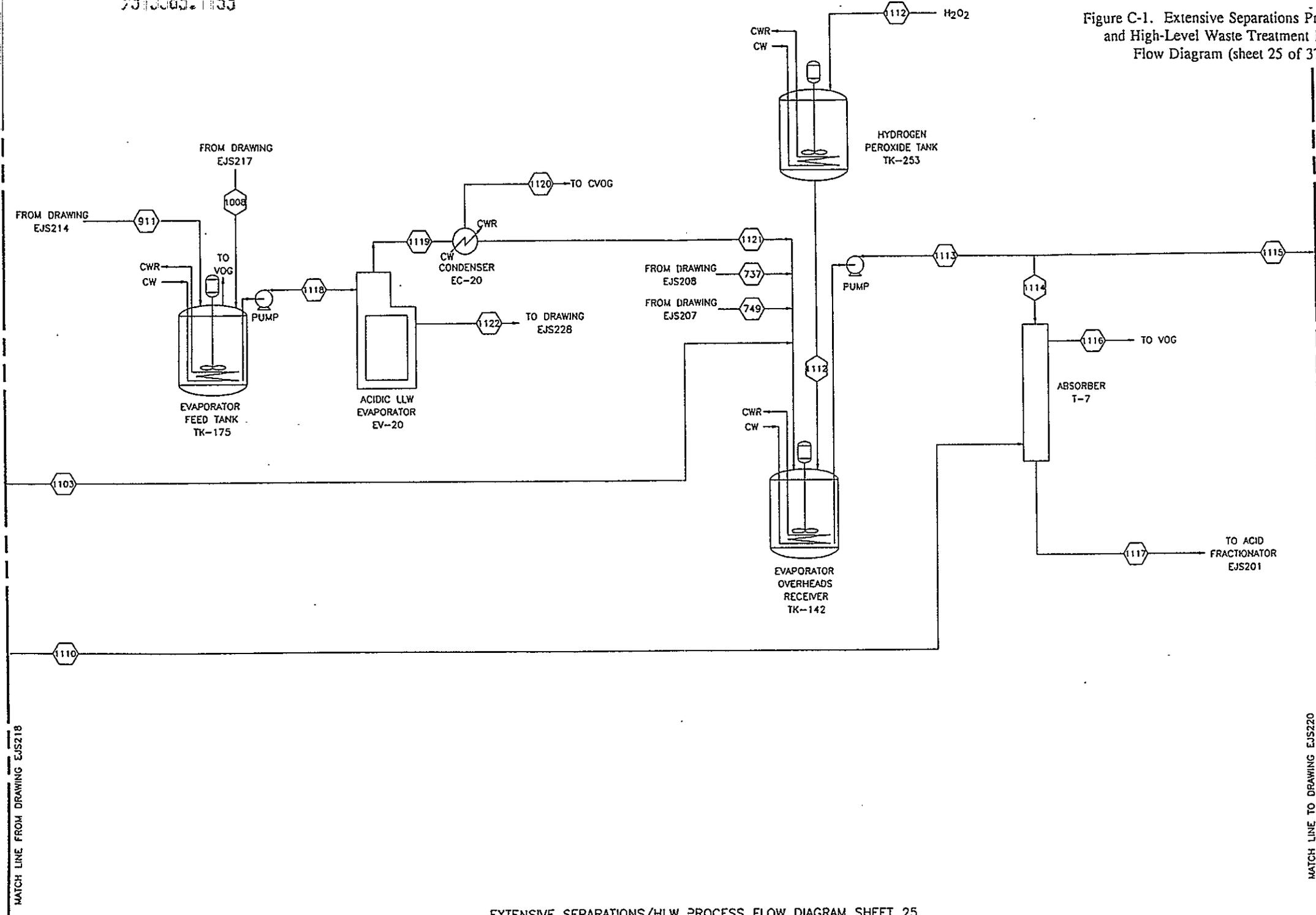
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 24 of 37).



MATCH LINE FROM DRAWING EJS217

MATCH LINE TO DRAWING EJS219

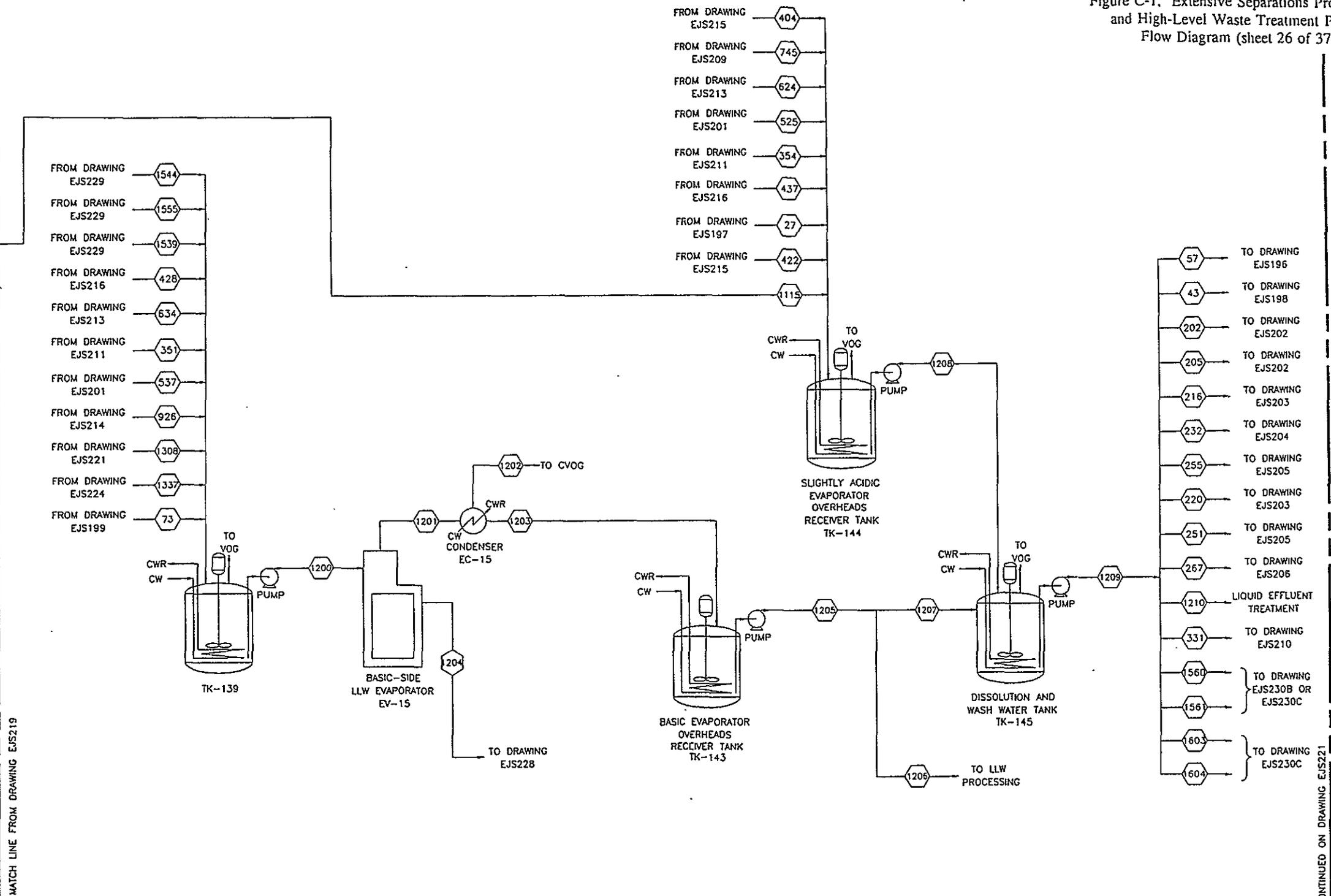
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 25 of 37).



MATCH LINE FROM DRAWING EJS218

MATCH LINE TO DRAWING EJS220

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 26 of 37).



MATCH LINE FROM DRAWING EJS219

CONTINUED ON DRAWING EJS221

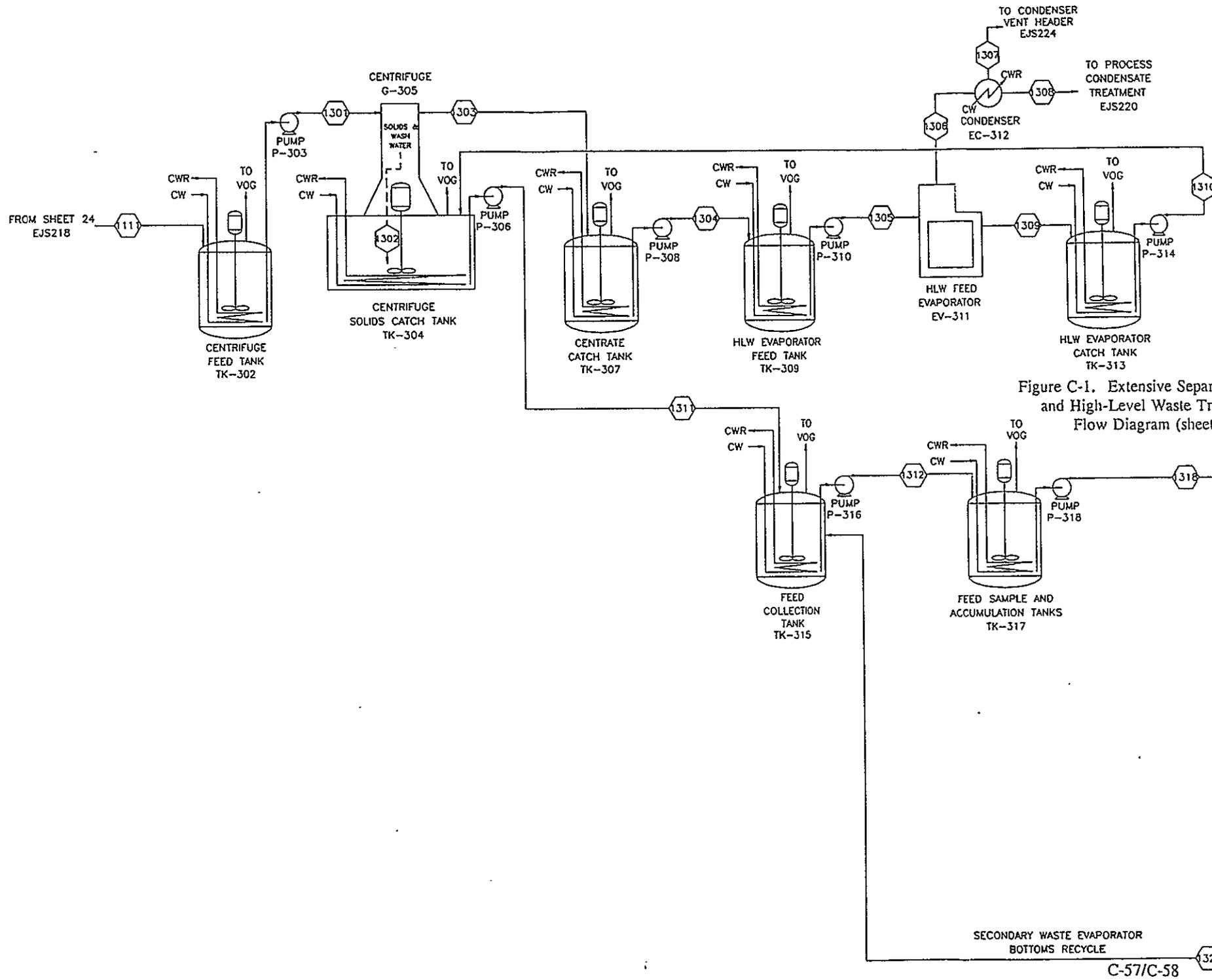
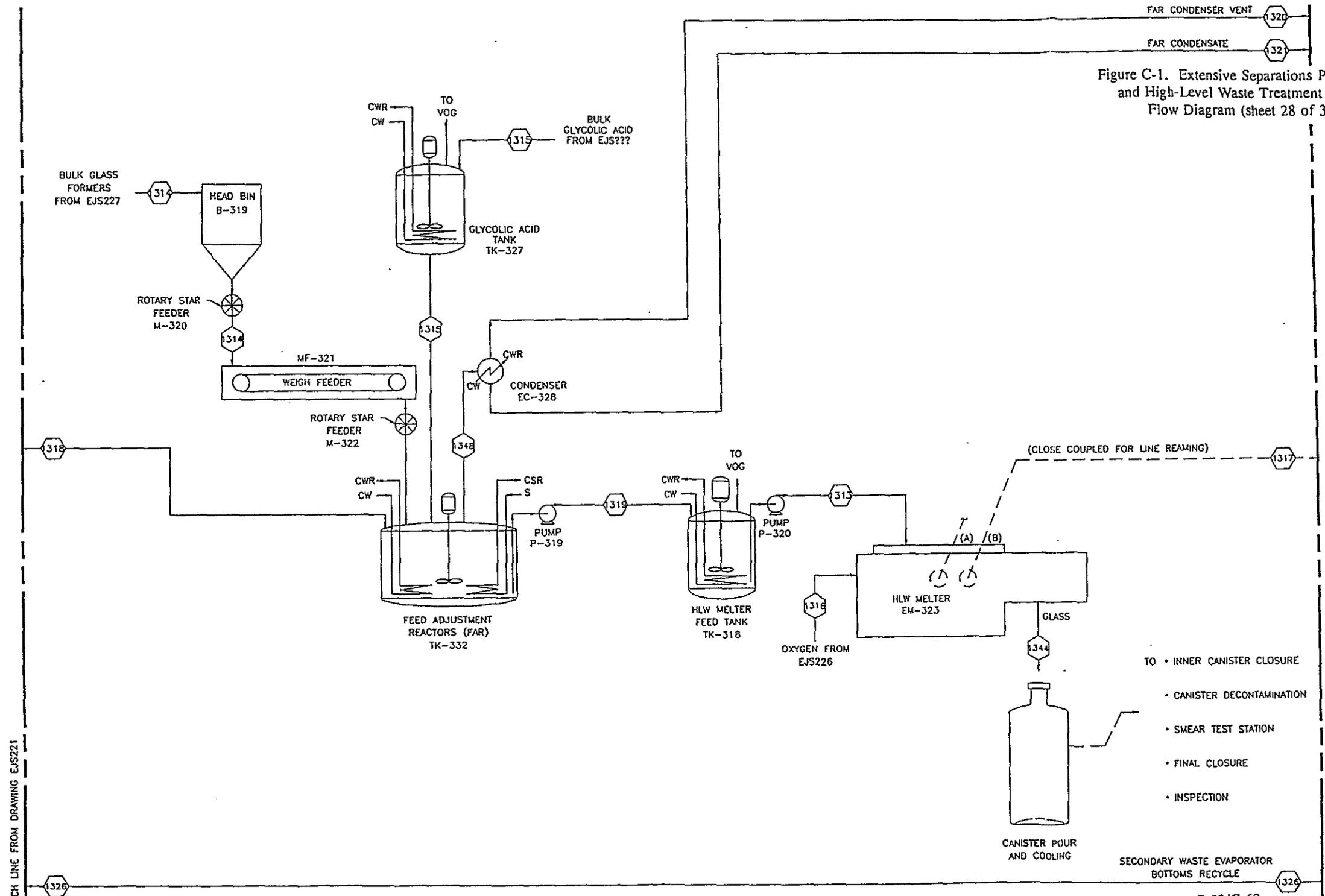


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 27 of 37).

MATCH LINE FROM DRAWING EJS220

MATCH LINE TO DRAWING EJS222

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 28 of 37).



- TO • INNER CANISTER CLOSURE
- CANISTER DECONTAMINATION
- SMEAR TEST STATION
- FINAL CLOSURE
- INSPECTION

MATCH LINE FROM DRAWING EJS221

MATCH LINE TO DRAWING EJS223

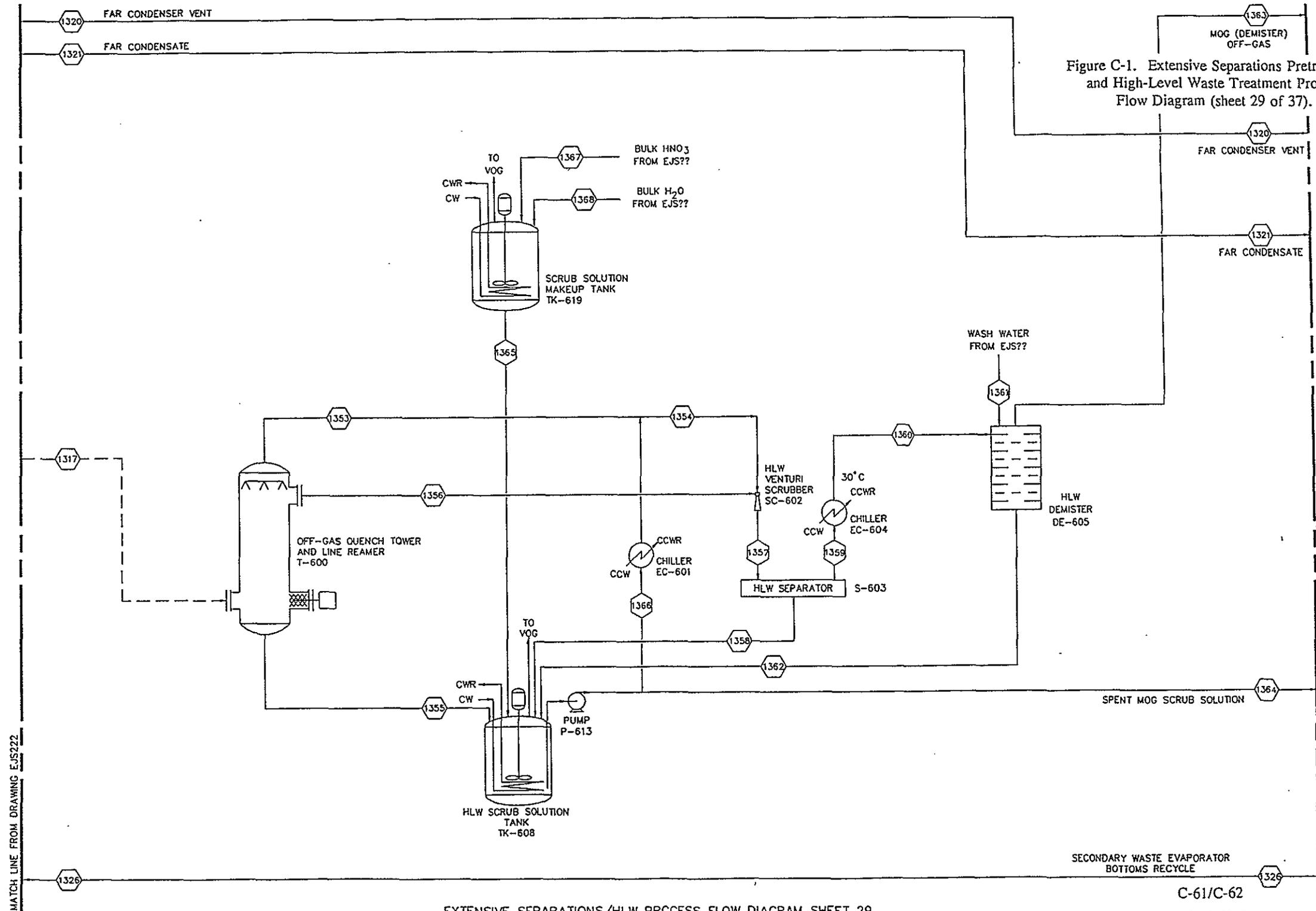


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 29 of 37).

MATCH LINE FROM DRAWING EJS222

MATCH LINE TO DRAWING EJS224

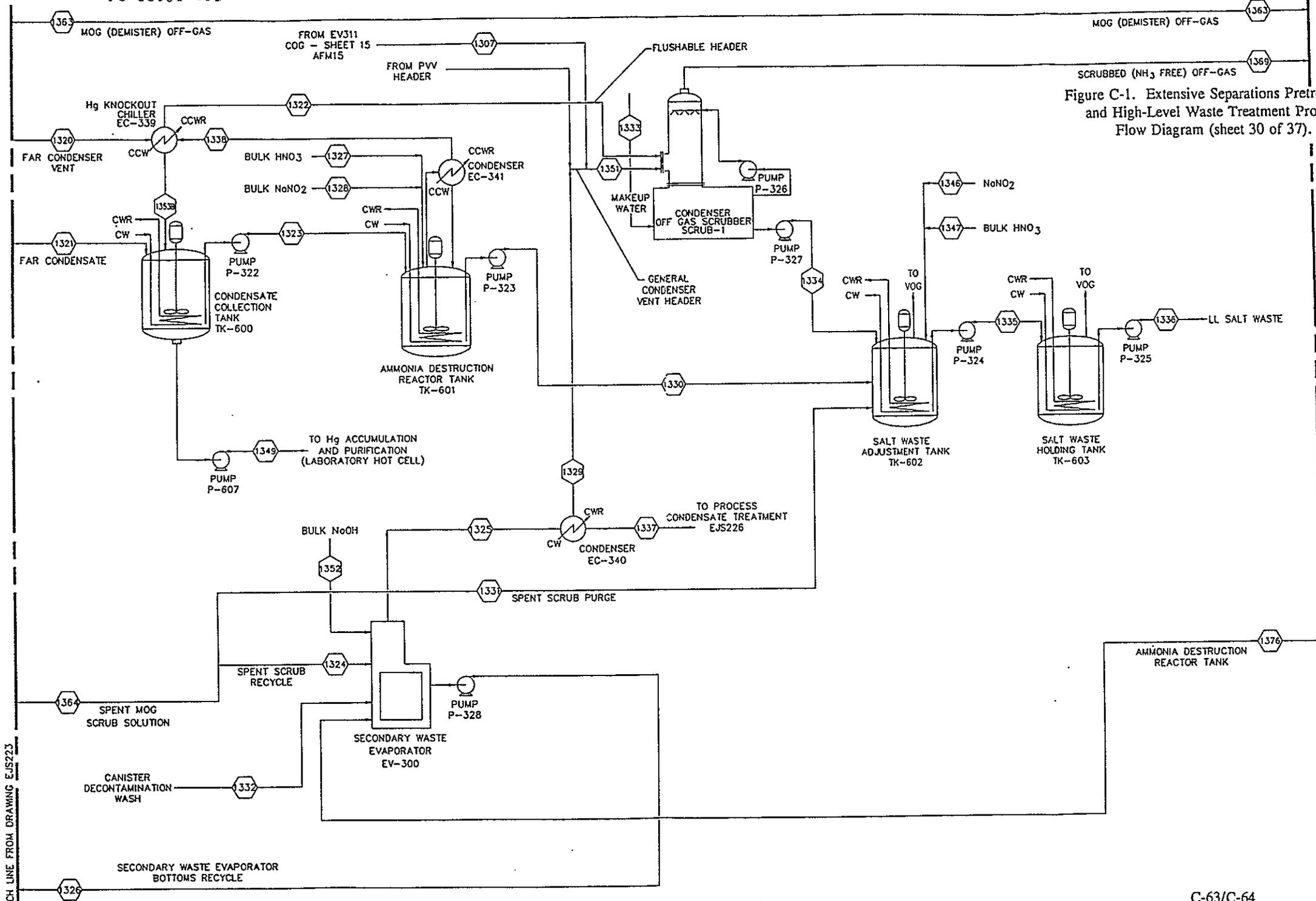
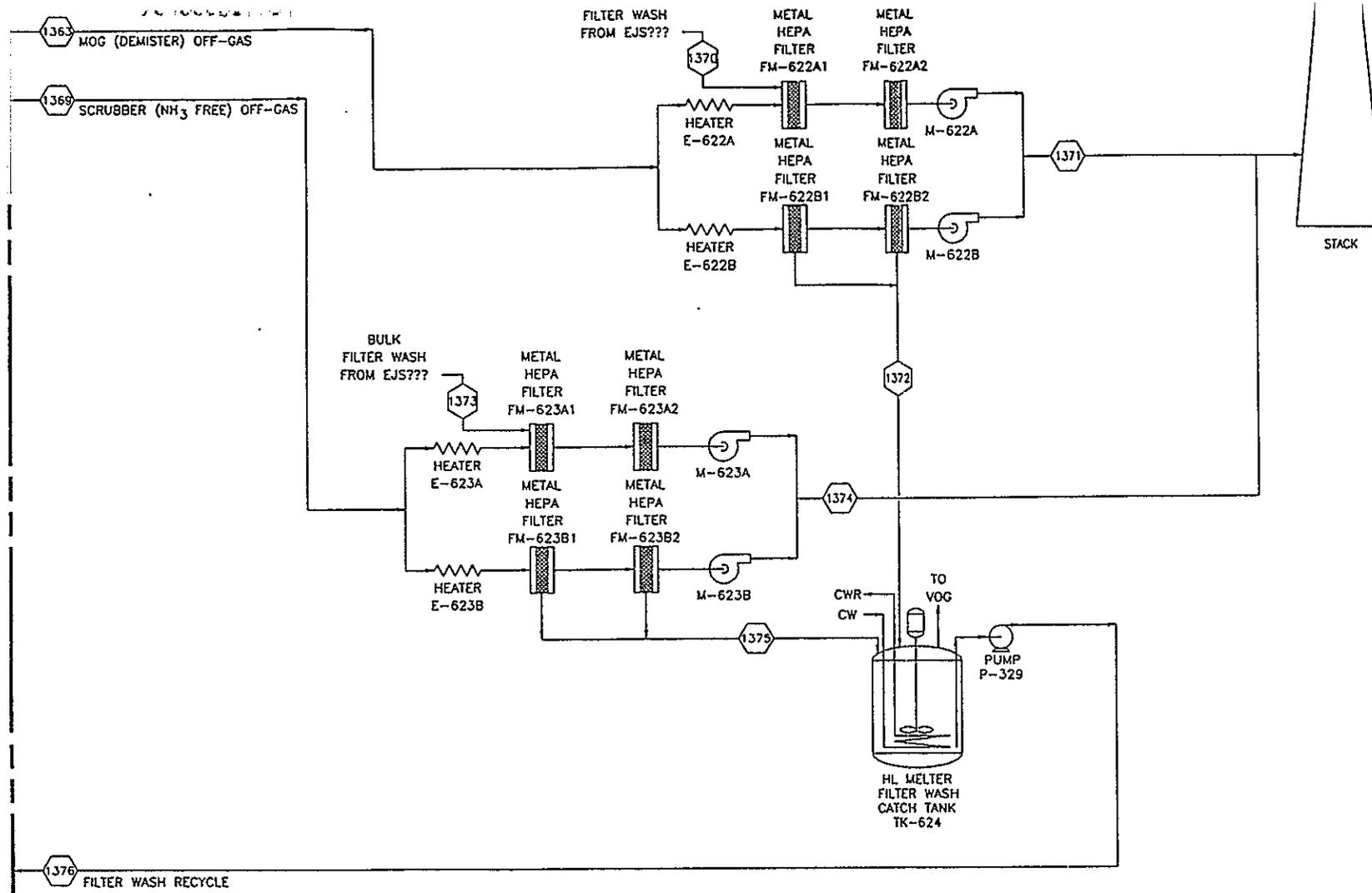


Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 30 of 37).

MATCH LINE FROM DRAWING EJS223

MATCH LINE TO DRAWING EJS225

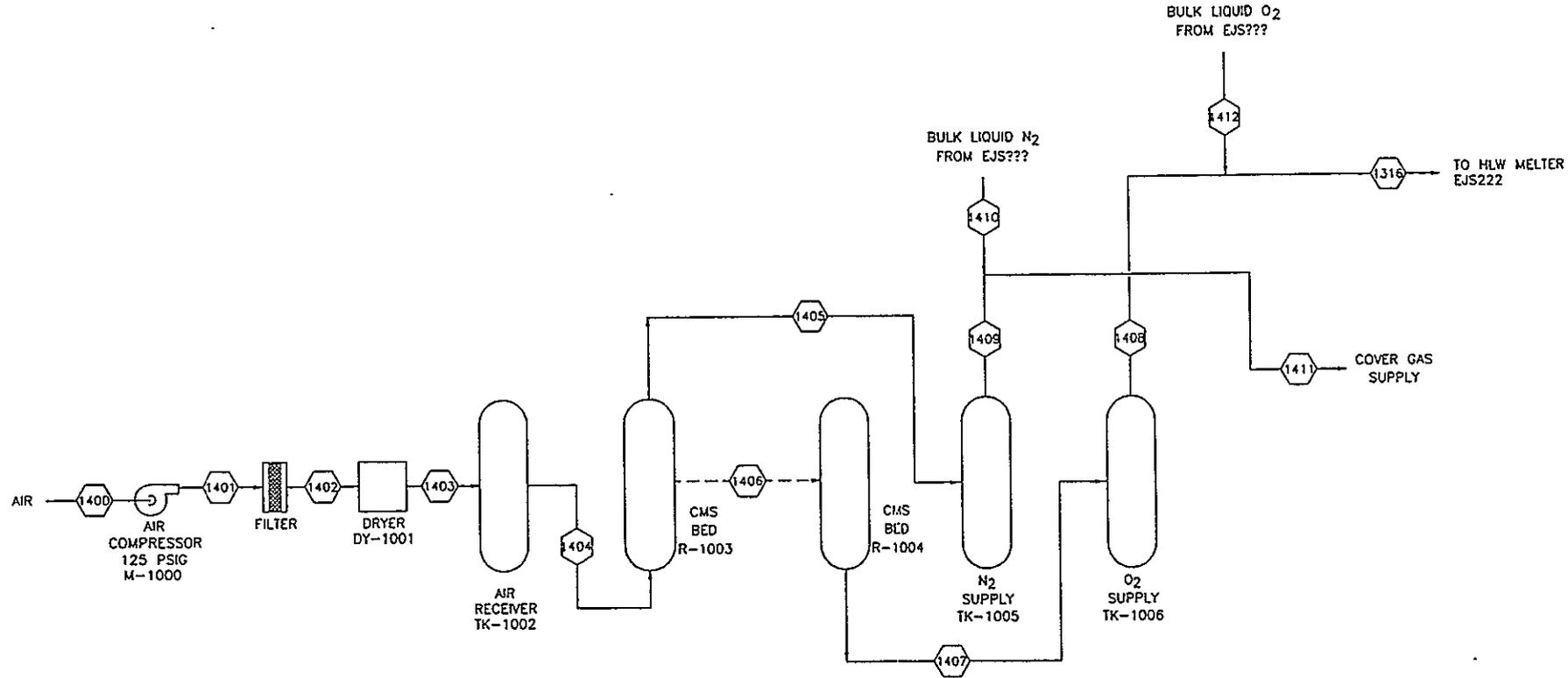
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 31 of 37).



MATCH LINE FROM DRAWING EJS224

CONTINUED ON DRAWING EJS226

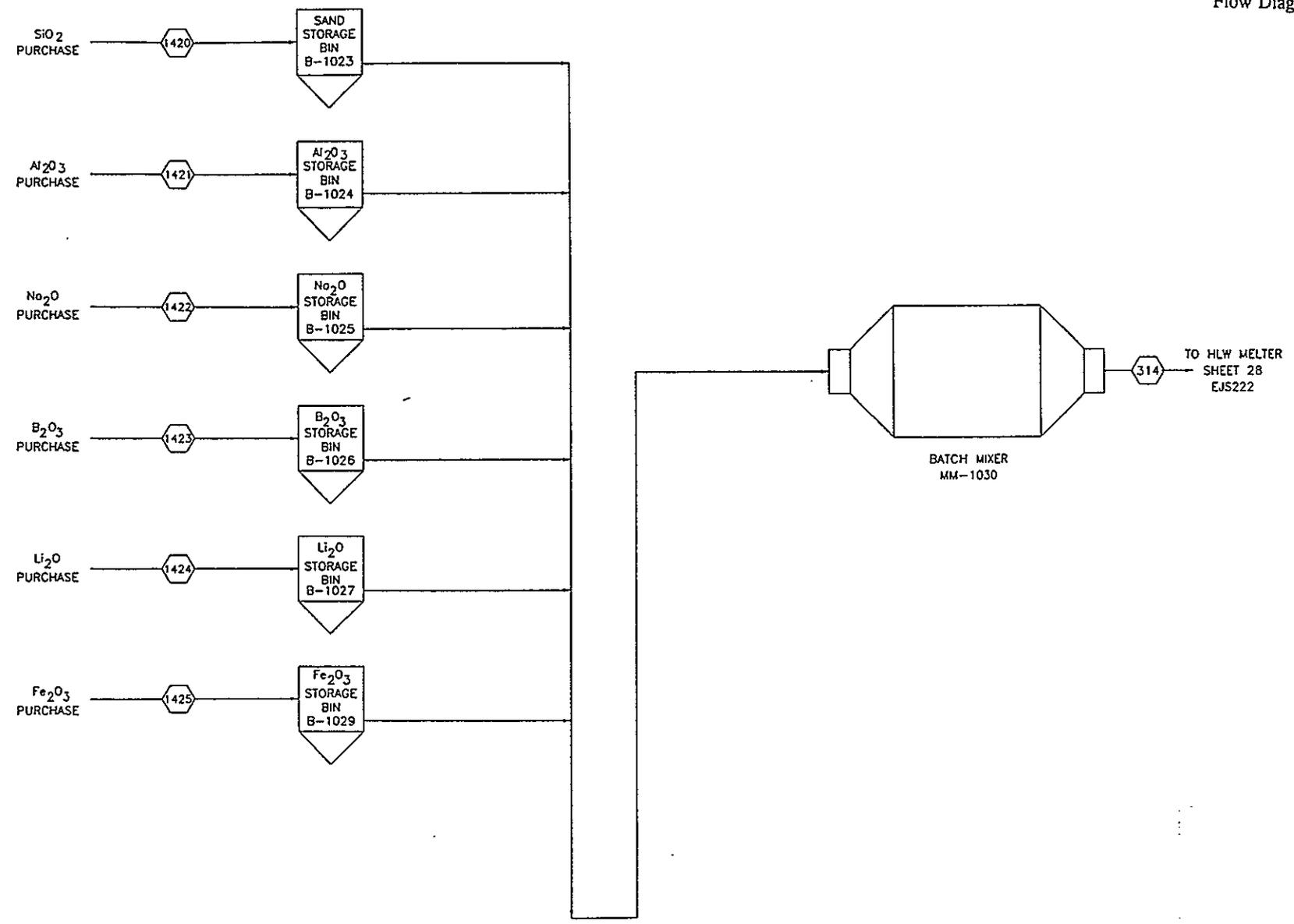
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 32 of 37).



CONTINUED FROM DRAWING EJS225

CONTINUED ON DRAWING EJS227

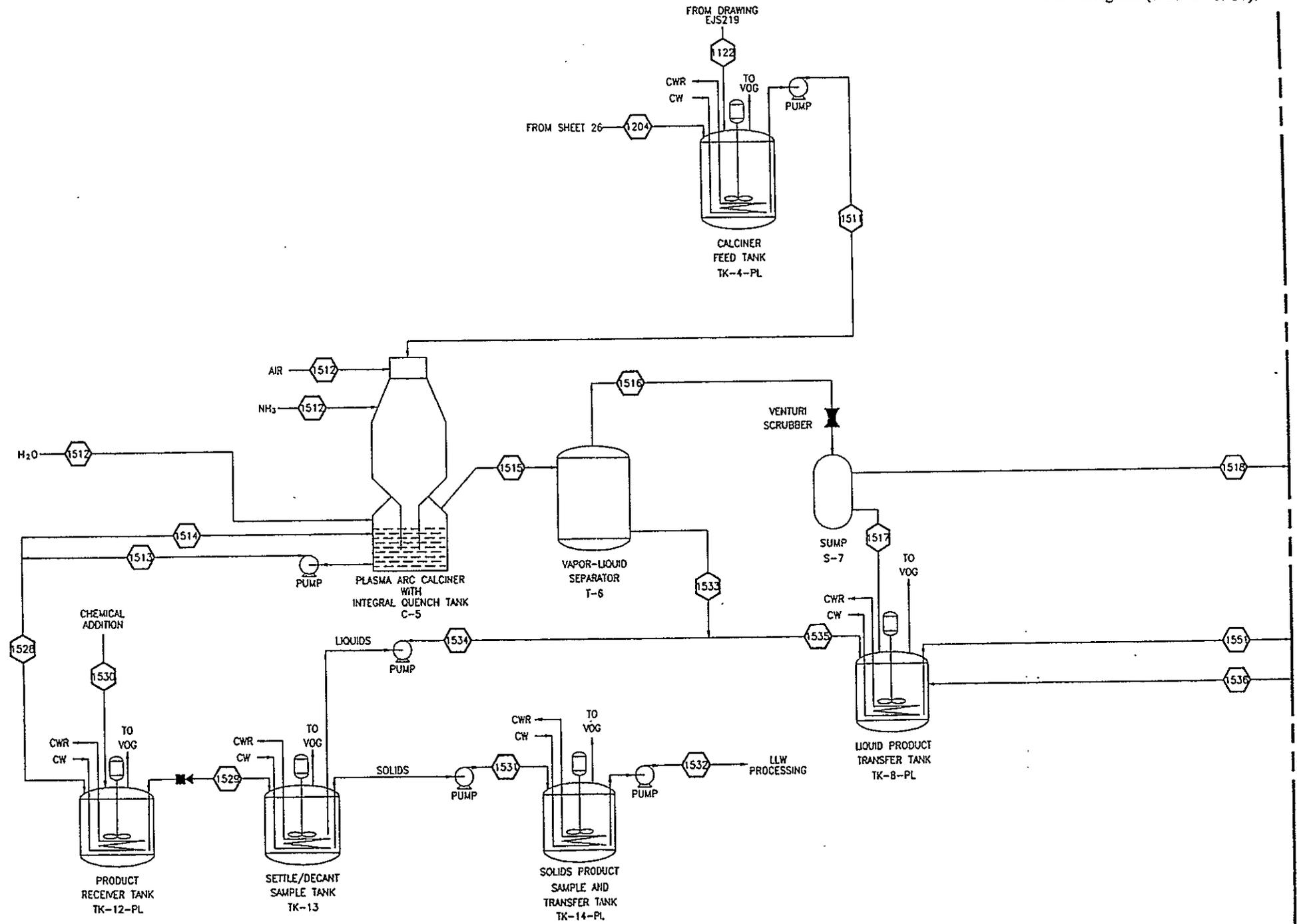
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 33 of 37).



CONTINUED FROM DRAWING EJS226

CONTINUED ON DRAWING EJS228

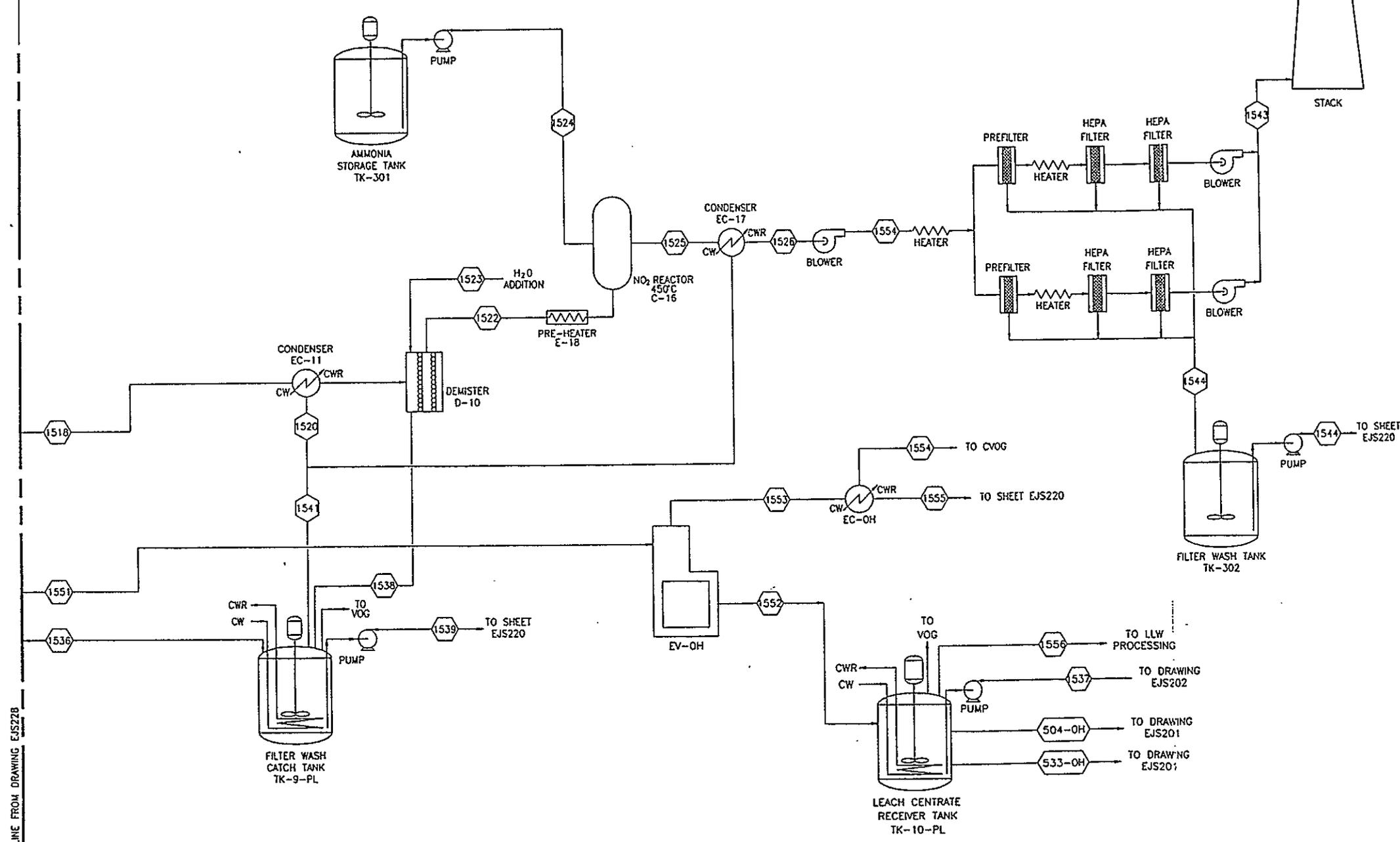
Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 34 of 37).



MATCH LINE TO DRAWING EJS227

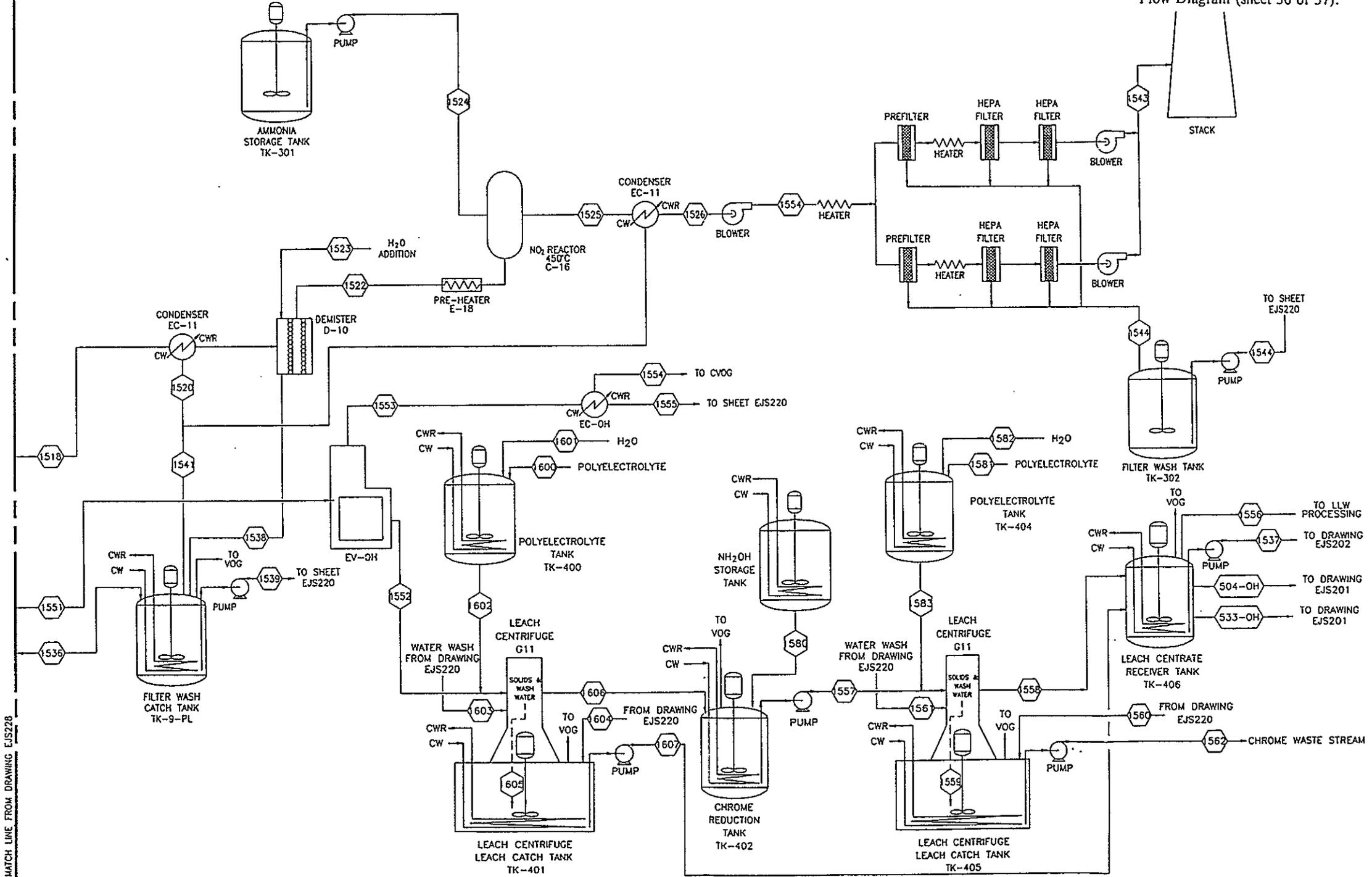
MATCH LINE TO DRAWING EJS229

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 35 of 37).



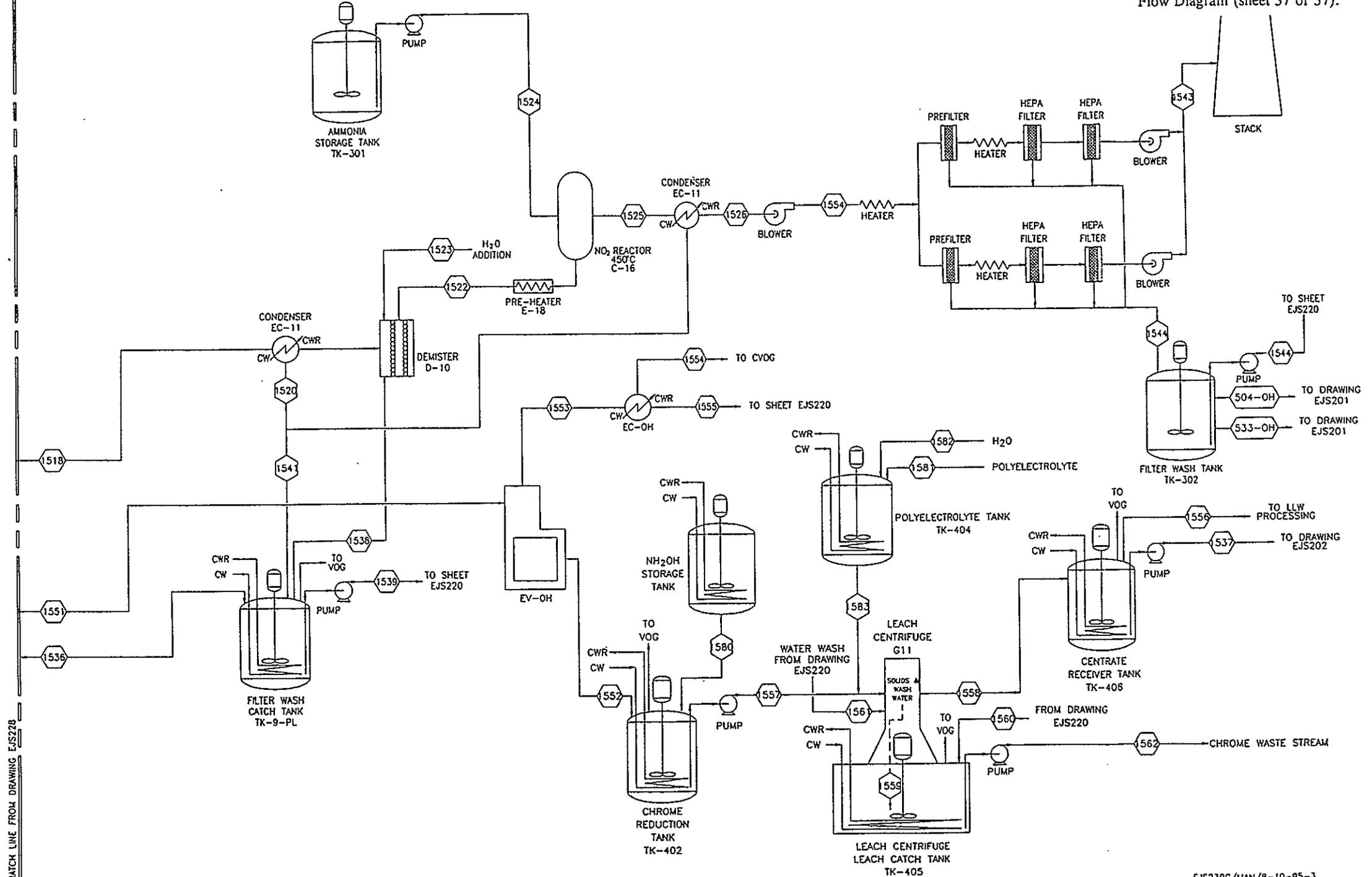
MATCH LINE FROM DRAWING EJS228

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 36 of 37).



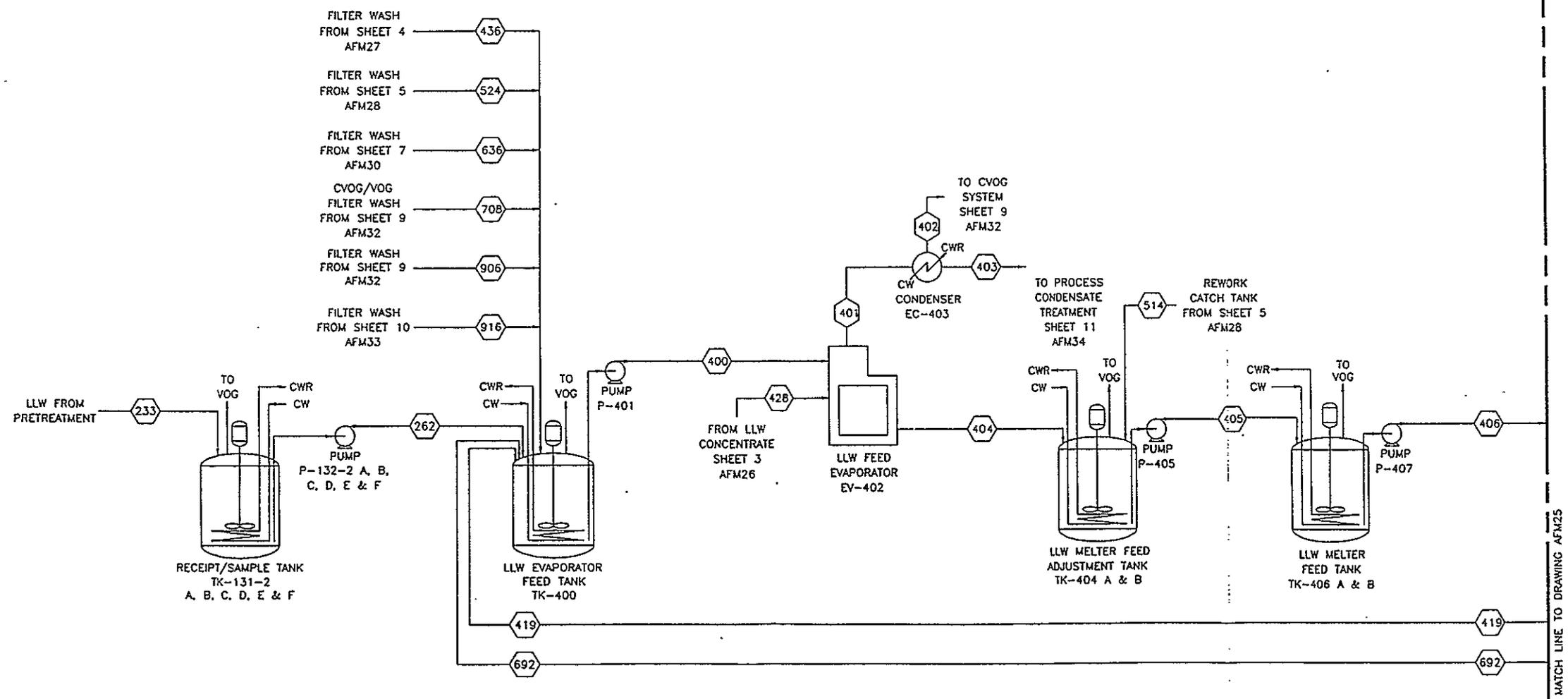
MATCH LINE FROM DRAWING EJS228

Figure C-1. Extensive Separations Pretreatment and High-Level Waste Treatment Process Flow Diagram (sheet 37 of 37).



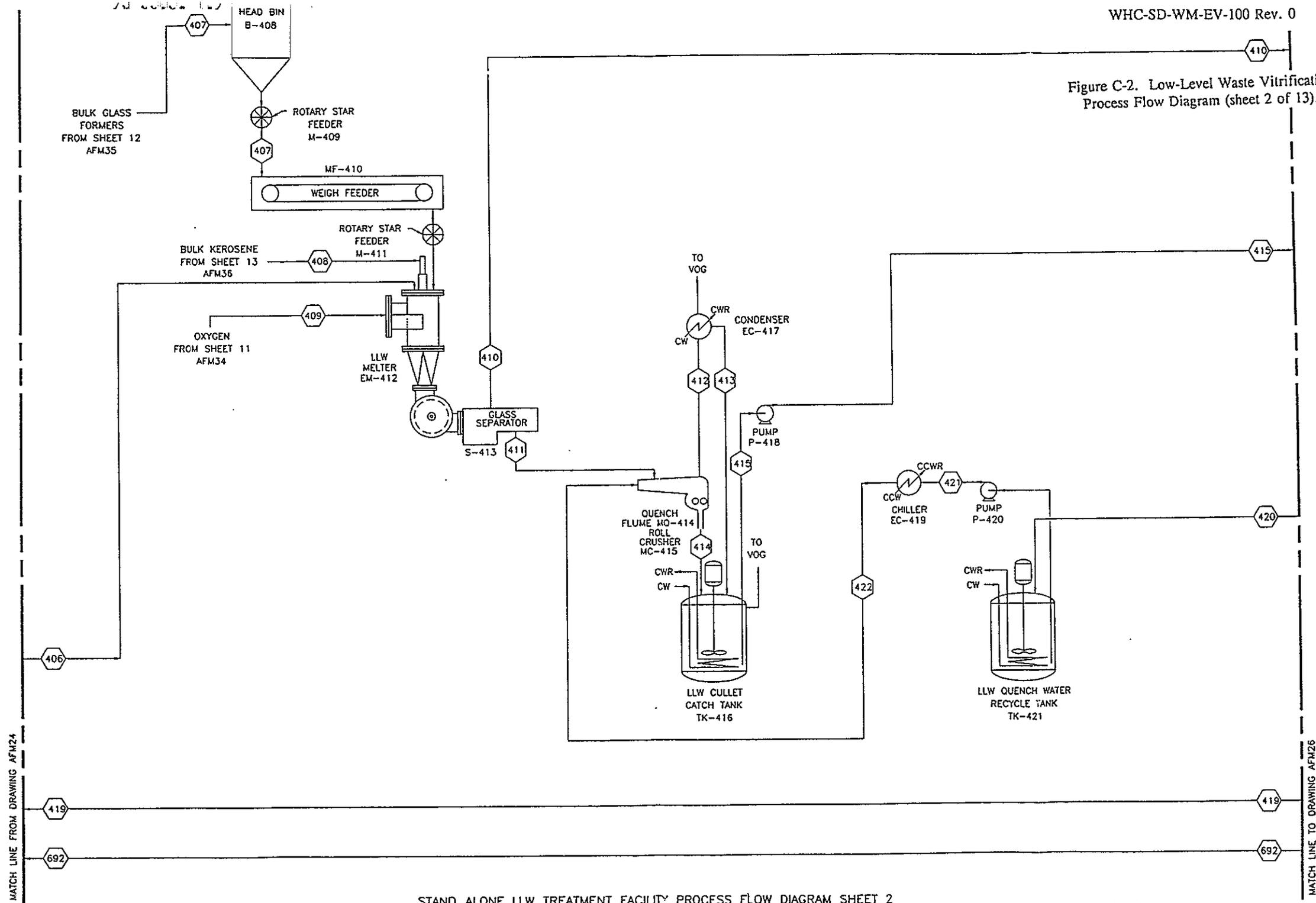
MATCH LINE FROM DRAWING EJS228

Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 1 of 13).



MATCH LINE TO DRAWING AFM23

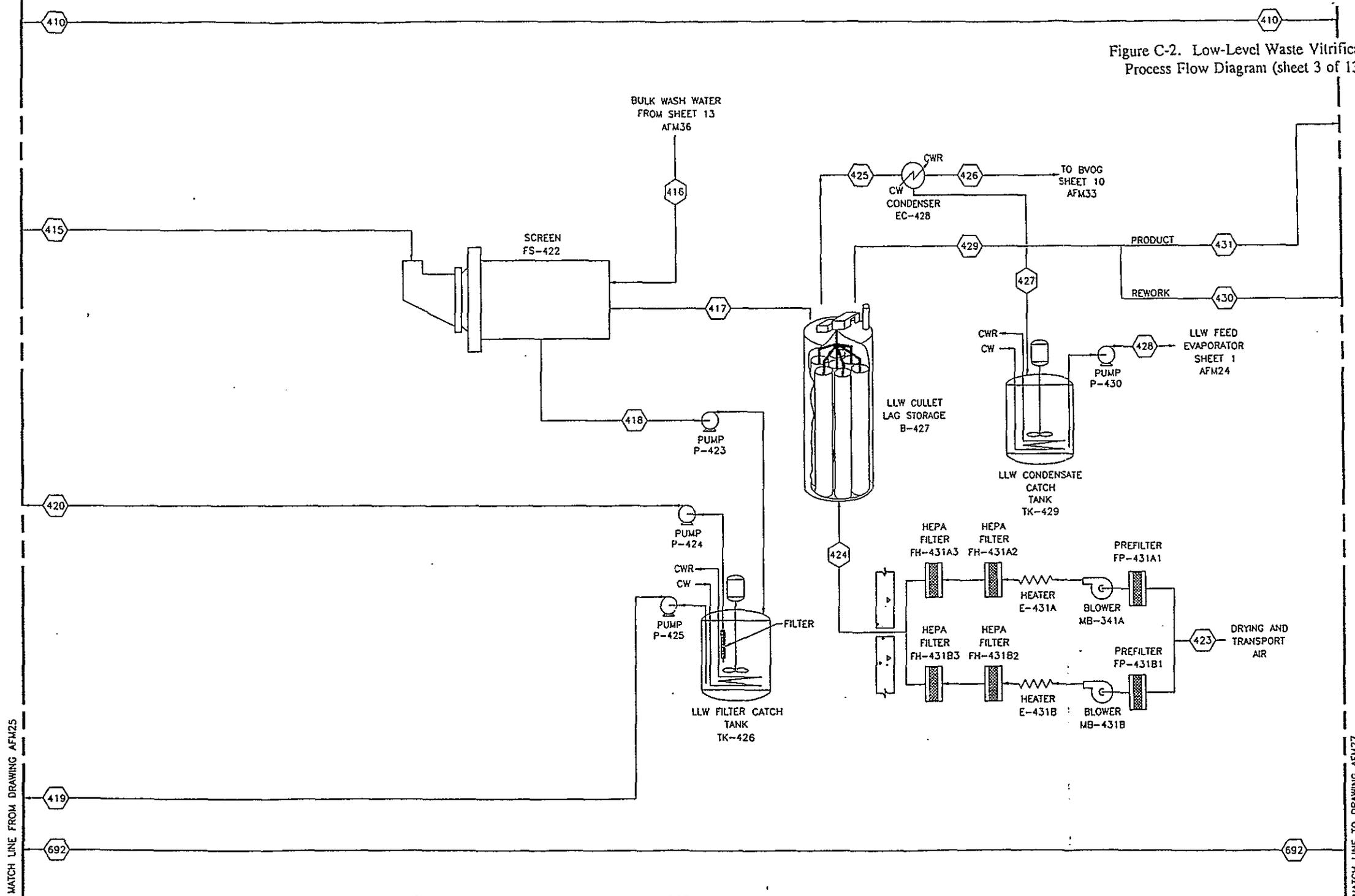
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 2 of 13).



MATCH LINE FROM DRAWING AFM24

MATCH LINE TO DRAWING AFM25

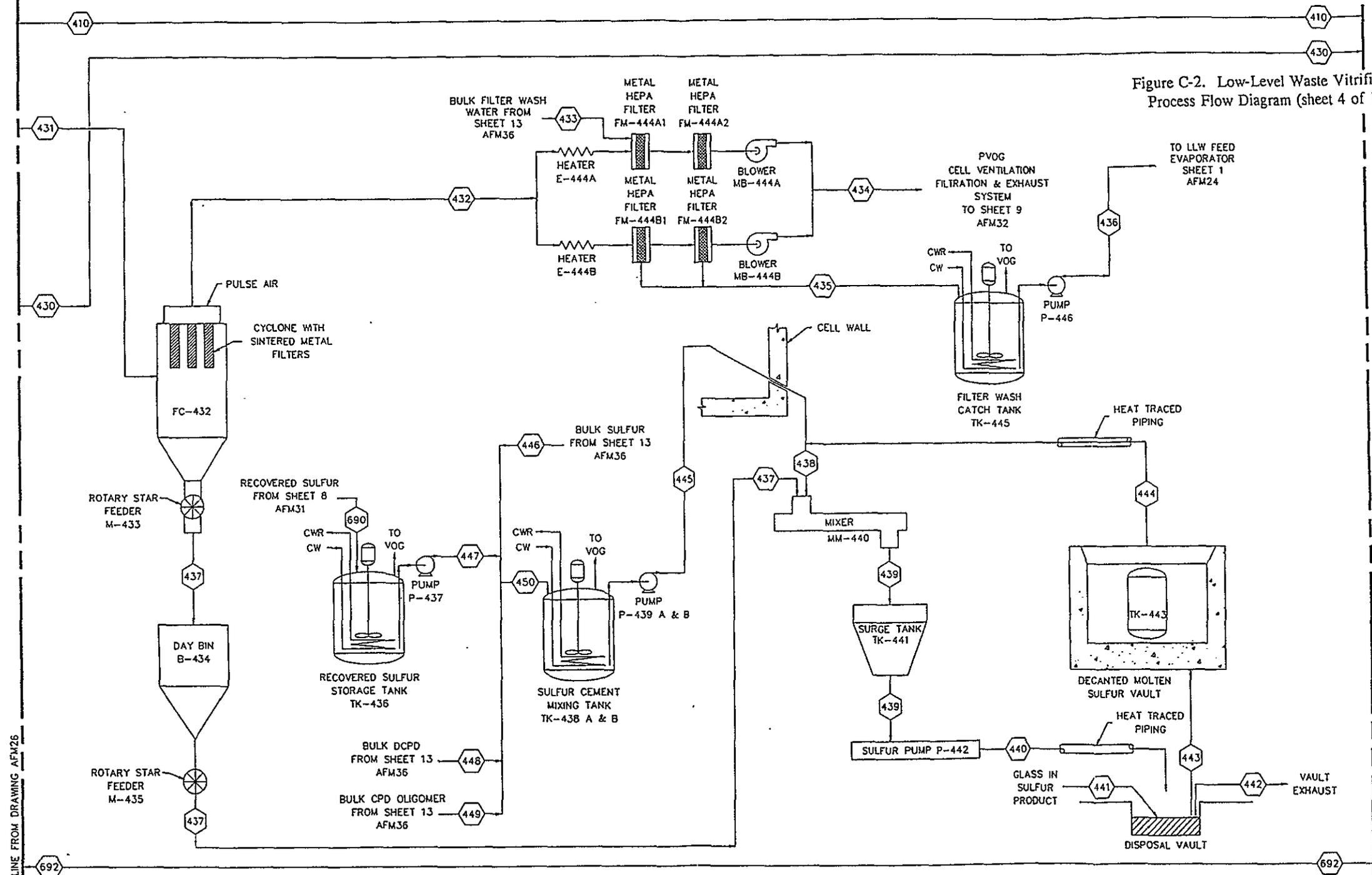
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 3 of 13).



MATCH LINE FROM DRAWING AFM25

MATCH LINE TO DRAWING AFM27

Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 4 of 13).

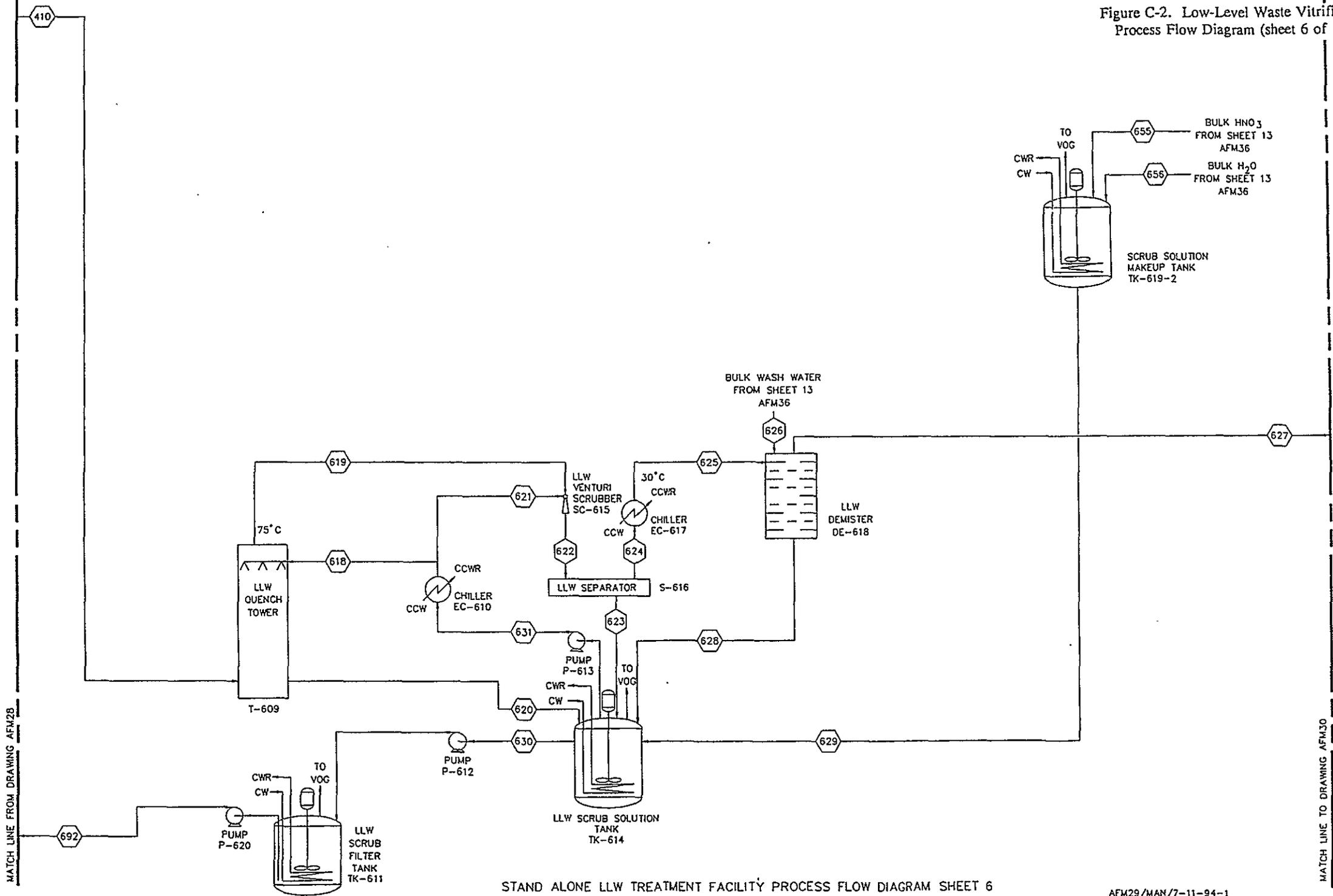


MATCH LINE FROM DRAWING AFM25

MATCH LINE TO DRAWING AFM28



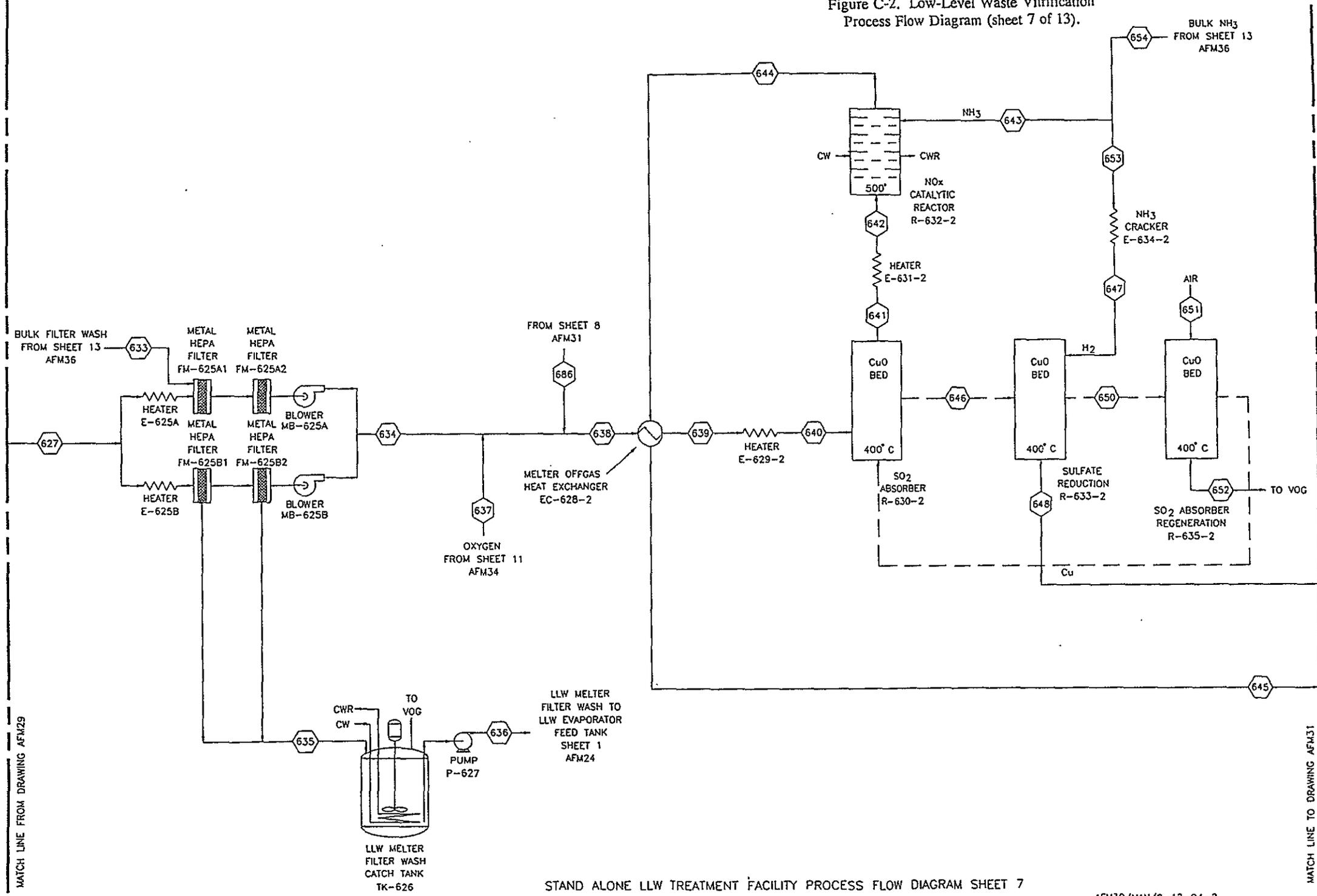
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 6 of 13).



MATCH LINE FROM DRAWING AFM29

MATCH LINE TO DRAWING AFM30

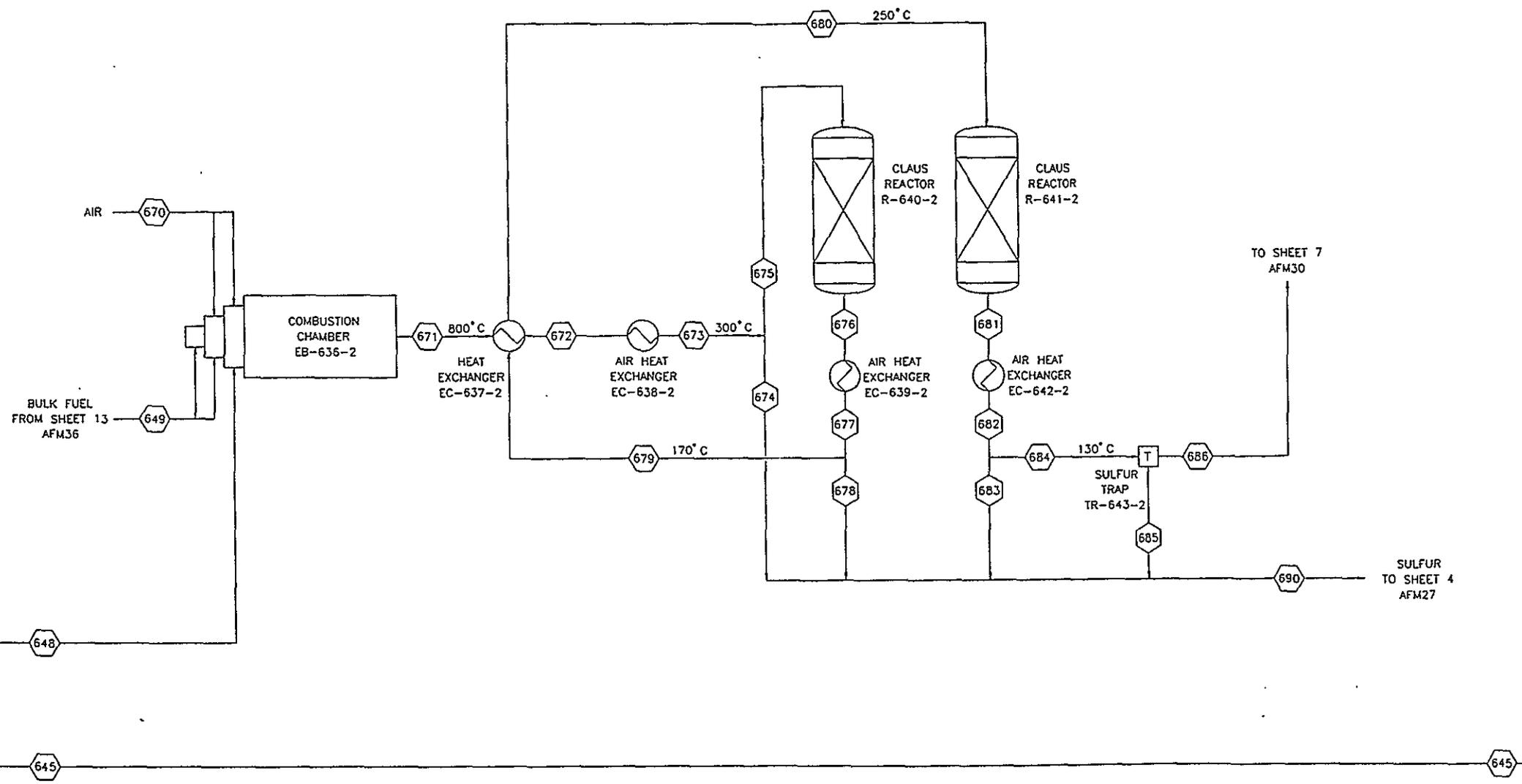
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 7 of 13).



MATCH LINE FROM DRAWING AFM29

MATCH LINE TO DRAWING AFM31

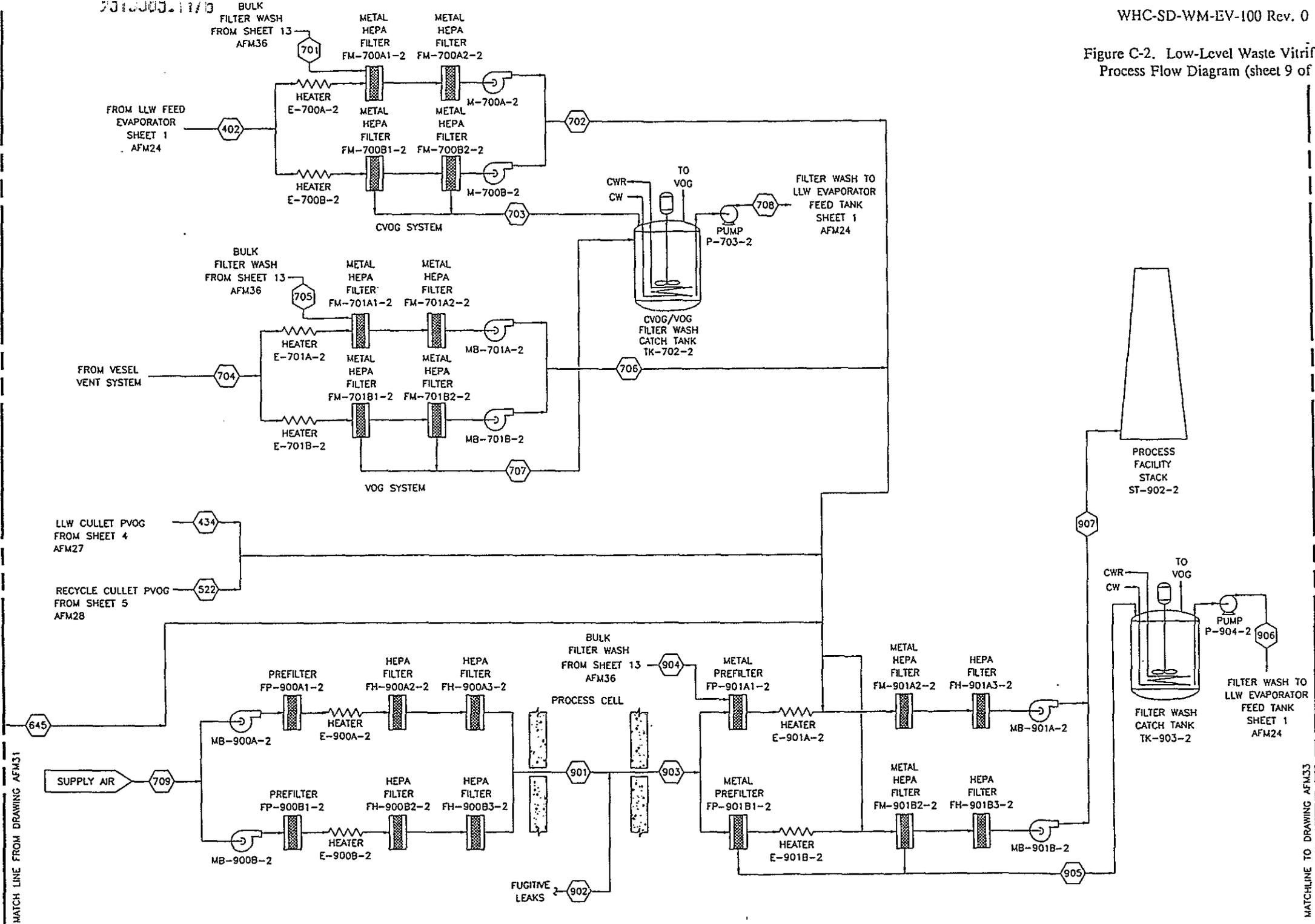
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 8 of 13).



MATCH LINE FROM DRAWING AFM30

MATCH LINE TO DRAWING AFM32

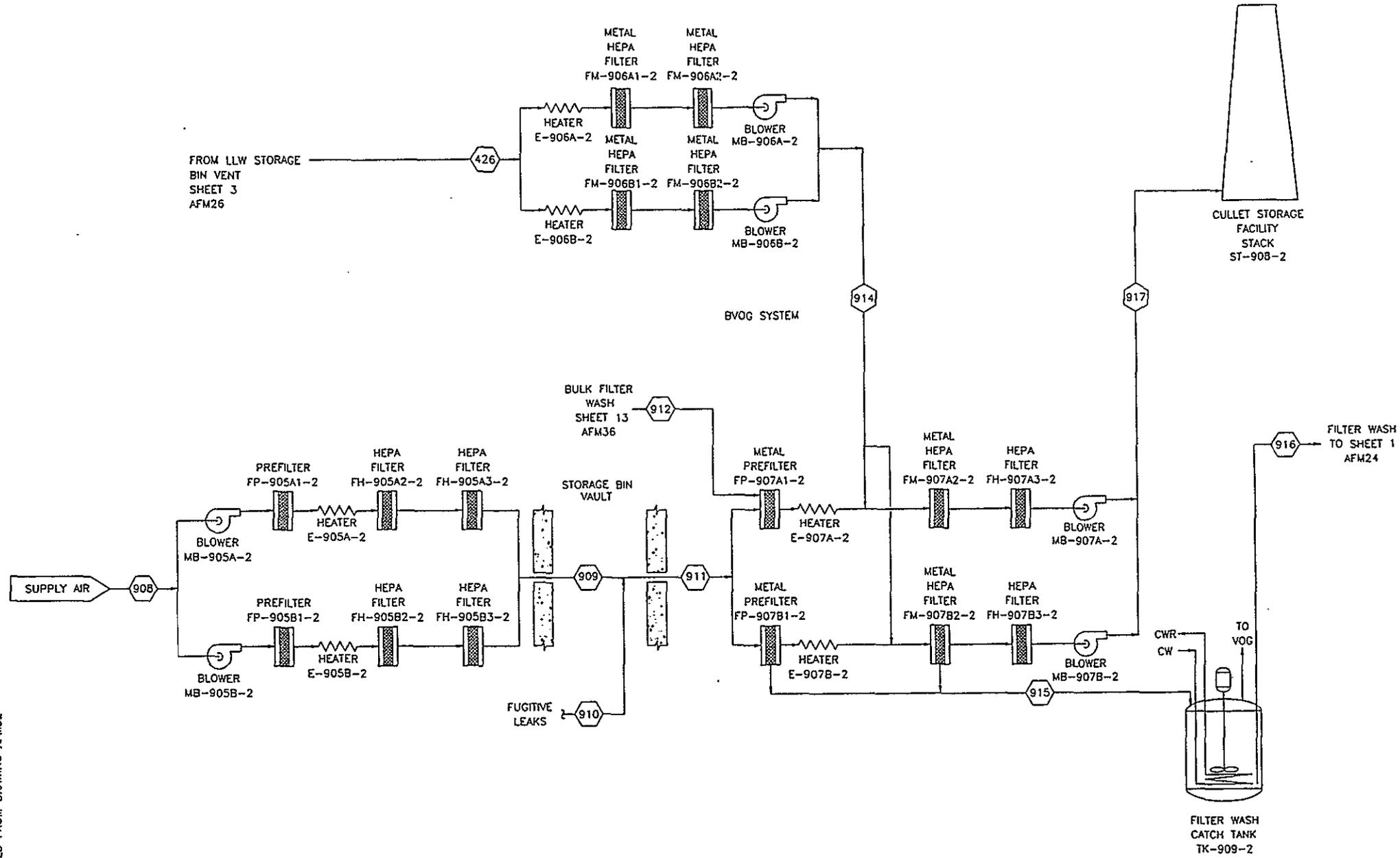
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 9 of 13).



MATCH LINE FROM DRAWING AFM31

MATCHLINE TO DRAWING AFM33

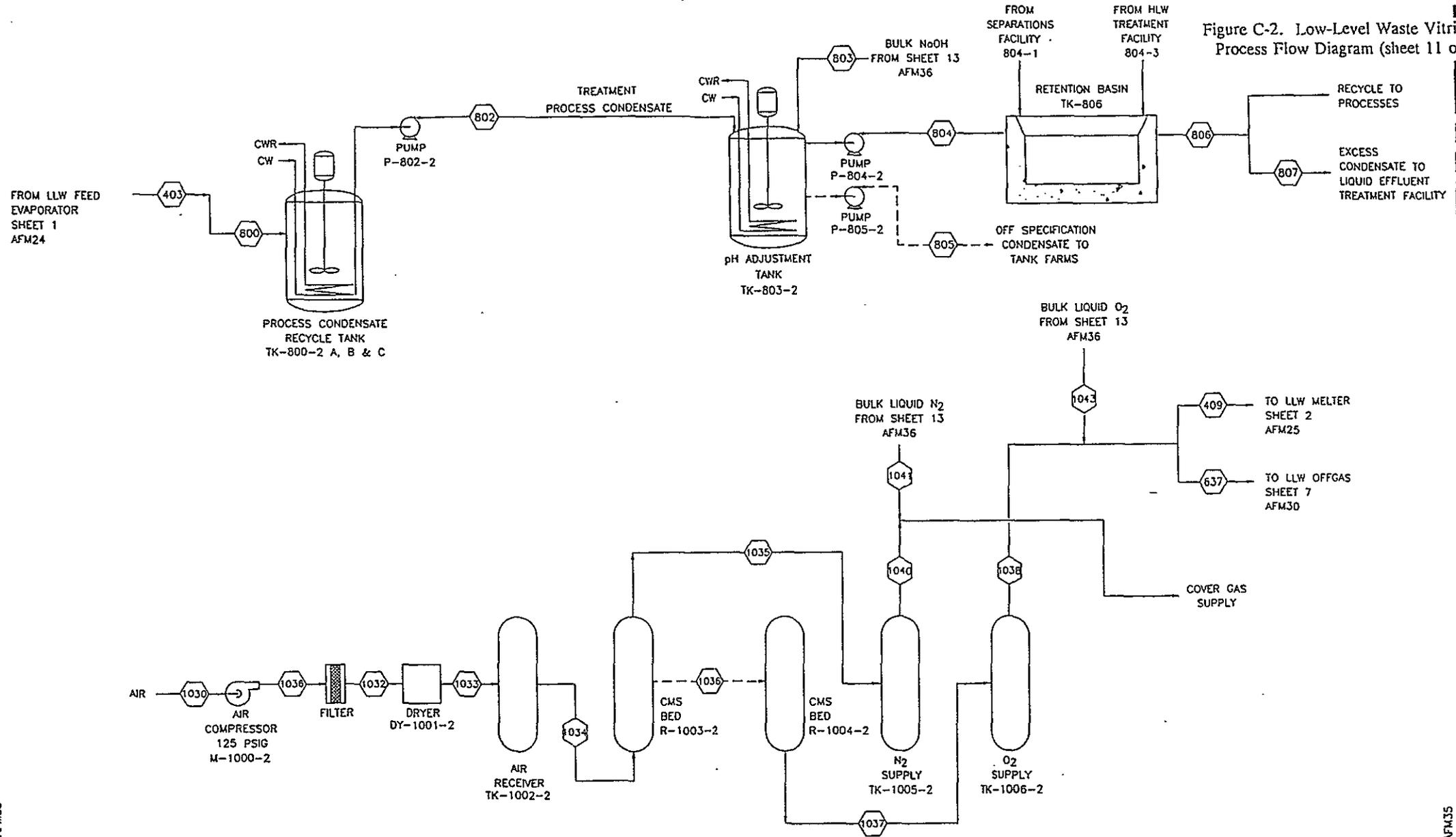
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 10 of 13).



CONTINUED FROM DRAWING AFM32

NEXT USED ON DRAWING AFM34

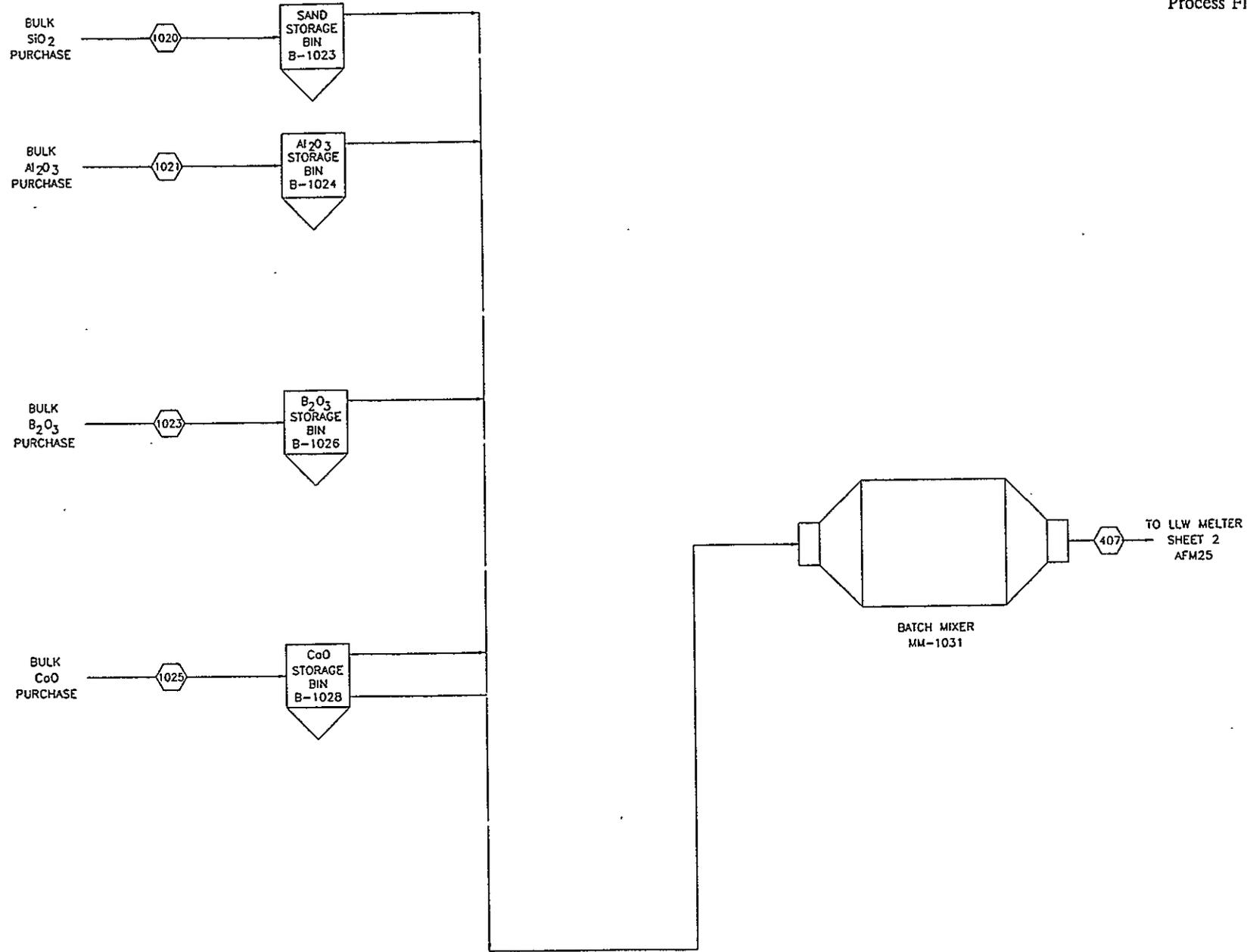
Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 11 of 13).



CONTINUED FROM DRAWING AFM33

NEXT USED ON DRAWING AFM35

Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 12 of 13).



CONTINUED FROM DRAWING AFM34

CONTINUED FROM DRAWING AFM36

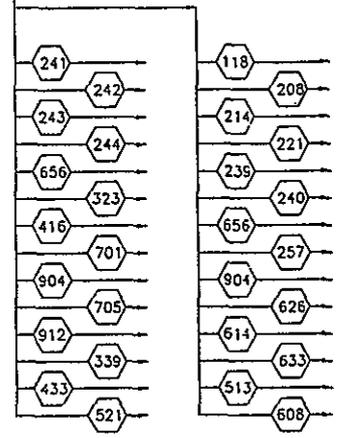
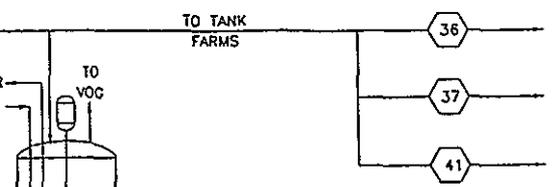
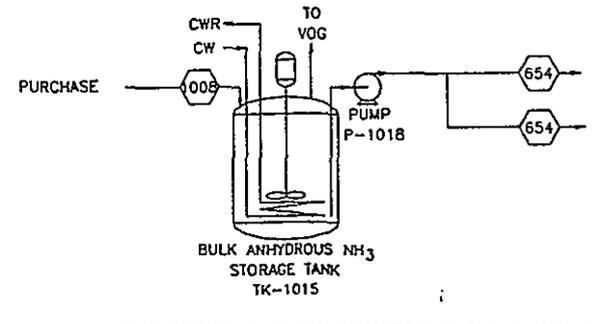
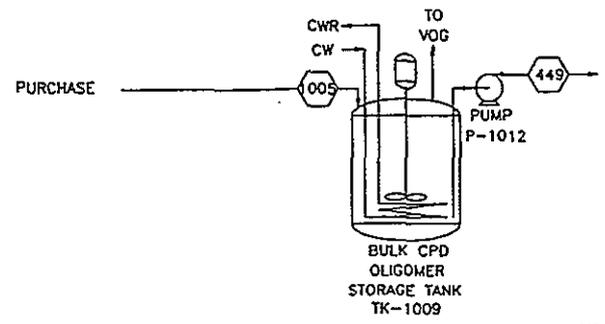
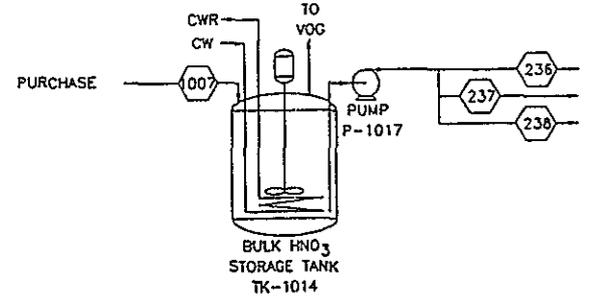
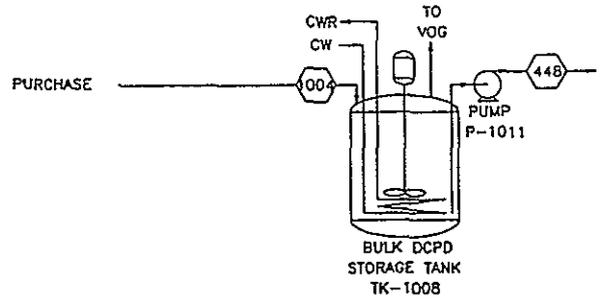
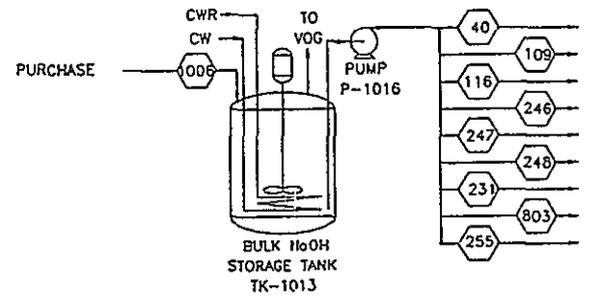
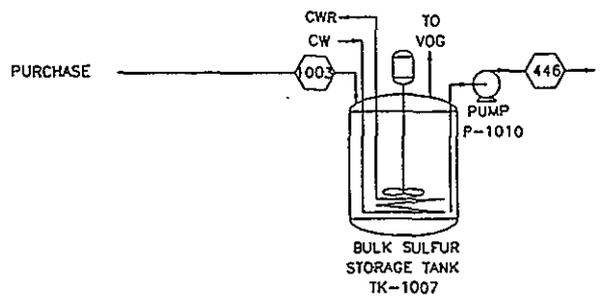
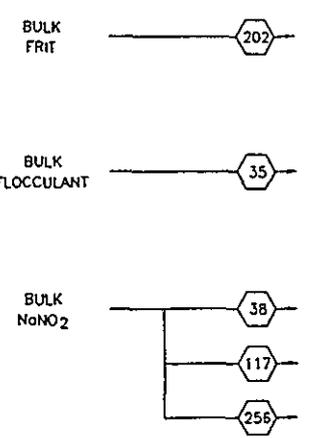
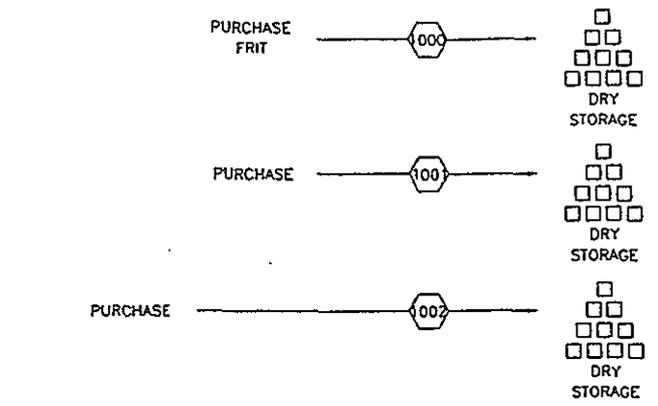
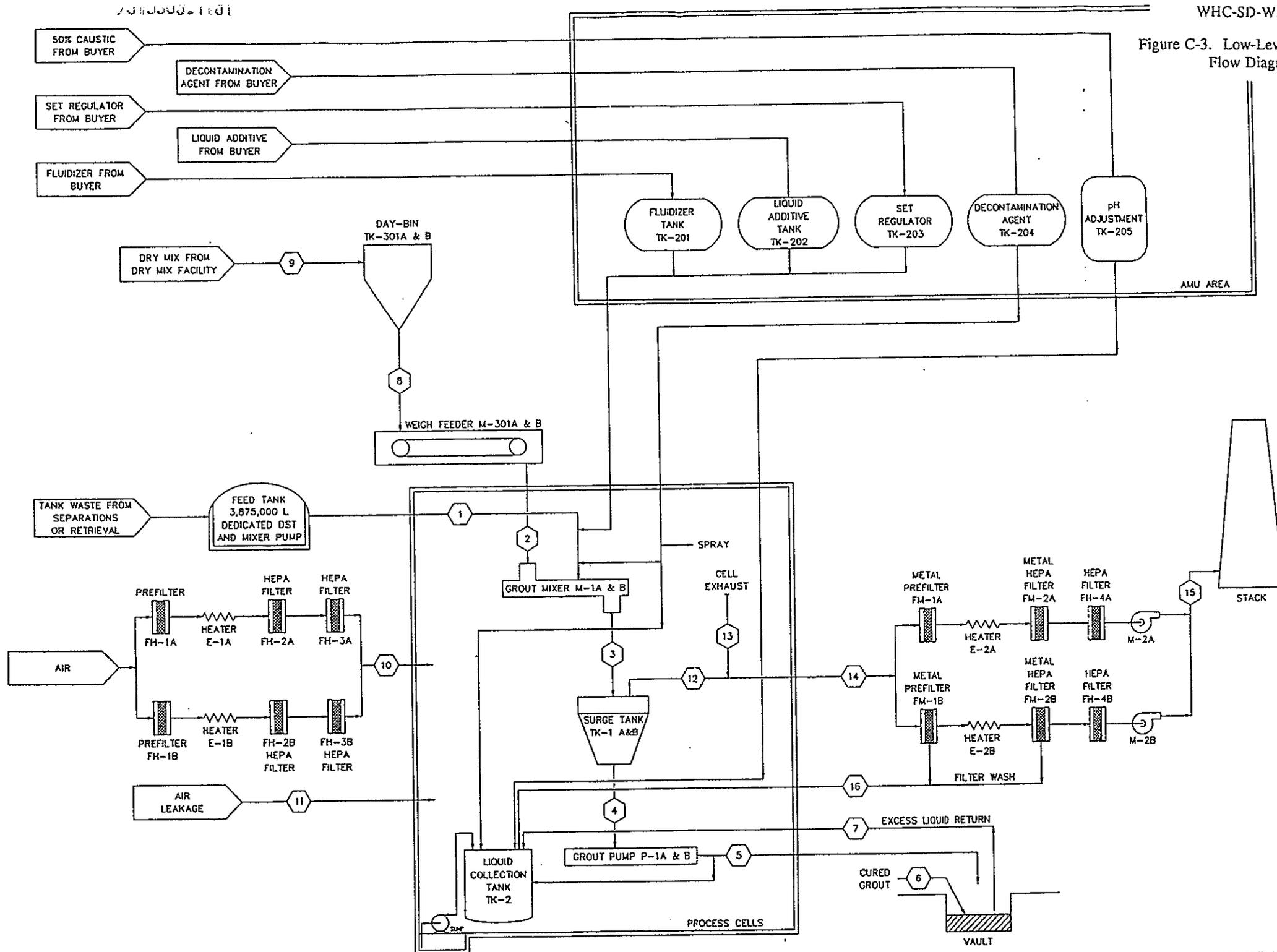


Figure C-2. Low-Level Waste Vitrification Process Flow Diagram (sheet 13 of 13).

CONTINUED FROM DRAWING AFM35

Figure C-3. Low-Level Waste Grout Process Flow Diagram (1 sheet)



SALT GROUT PROCESS FLOW DIAGRAM

**APPENDIX D**

**DESCRIPTION AND SPECIFICATIONS FOR UNIT PROCESSES**

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**LIST OF TERMS**

APM	Ammonium Phosphomolybdate
CMPO	octylphenyl-N, N-diisobutylcarbamoylmethylphosphine oxide
DST	double-shell tank
DTPA	Diethylene-triamine-penta-acetic acid
EDTA	Ethylenediaminetetraacetic acid
EIS	Environmental Impact Statement
ESP	Extensive Separations Pretreatment (alternative)
HEPA	high-efficiency particulate air (filter)
HVAC	heating, ventilating and air conditioning
HLW	high-level waste
IX	ion exchange
LLW	low-level waste
M	molarity
NO <sub>x</sub>	Oxides of Nitrogen
NPH	Normal Paraffin Hydrocarbon
psig	pounds per square inch gauge
SLS	solid/liquid separation
SST	single-shell tank
TBP	Tributyl Phosphate
TOC	Total Organic Carbon
TRU	Transuranic
TWRS	Tank Waste Remediation System

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## D1.0 INTRODUCTION

This appendix describes the Extensive Separations Pretreatment (ESP) alternative flowsheet model. To meet two principal objectives, this alternative contains additional acid-stream and alkaline-stream separation and chemical reaction steps beyond those in the Tri-Party Agreement baseline flowsheet. The objectives are to decrease the toxicity of the low-level radioactive waste form to U.S. Nuclear Regulatory Commission Class A limits and to decrease the volume of high-level radioactive waste to a limit of 1,000 canisters of glass. To reduce the production of glass, a high-level waste (HLW) vitrification facility would be included as part of the separations facility; therefore a separate HLW vitrification facility and associated tank farm lag storage would not be required. A proposed set of processes illustrates how the objectives could be met and have been combined into the process described below.

In the main body of this supporting document, Figure 3-1 shows a block diagram representing the Tank Waste Remediation System (TWRS) ESP alternative flowsheet. Figure 4-1 shows the differences between process alternatives included in the Environmental Impact Statement (EIS). Figure 4-2 shows the disposal of tank waste for the ESP alternative. Figure 4-3 shows the ESP flowsheet differences for the LLW form. The numerical bases of the inventories and separations factors for the unit processes are in Tables D-1 through D-8.

The following is a summary of the extensive processing unit operations. Many of the process descriptions in this appendix are summaries of information in WHC-EP-0616 (Boomer et al. 1993) and PNL-8388 (Swanson 1993).

### SEPARATE AND DISSOLVE SOLIDS

- Separate and wash sludge
- Leach sludge with sodium hydroxide
- Dissolve sludge with nitric acid
- Dissolve sludge with nitric acid containing fluoride.

### PURIFY ACID SOLUBLE RADIONUCLIDES

- Extract uranium, plutonium, neptunium, and thorium with tributyl phosphate
  - Purify uranium with tributyl phosphate
- Extract americium and lanthanides with octylphenyl-N, N-diisobutylcarbamoylmethylphosphine oxide (CMPO)
  - Purify americium and higher lanthanides by chromatographic ion exchange (IX)

- Extract technetium, strontium and barium with crown ether
- Remove and recycle cesium by ammonium phosphomolybdate.

#### **REMOVE RADIONUCLIDES FROM ALKALINE LIQUIDS**

- Destroy complexants hydrothermally
- Remove cesium by resorcinol formaldehyde IX
- Remove and recycle strontium by crystalline silicotitanate IX
- Remove technetium by anion exchange.

#### **RECOVER AND REUSE BULK CHEMICALS**

- Evaporate and reuse water
- Distill fractionate and reuse nitric acid
- Destroy nitrate by calcination or electrolysis
- Recover and recycle sodium hydroxide.

#### **REMOVE HEAVY METALS**

- Chromium reduction and removal from recycled sodium hydroxide for LLW grout option.

The following processing alternatives were not considered in the Extensive Separations alternative.

- In-tank sludge wash
- Duolite<sup>TM</sup>\* CS-100 cesium ion exchange
- Irreversible high selectivity cesium absorbents
- Organic destruction by supercritical water oxidation (by heating and digesting at atmospheric pressure, by adding oxygen and potassium permanganate, or by higher temperature and pressure hydrothermal oxidation)
- Combined single-solvent extraction of americium, plutonium and strontium
- Purification of strontium from barium by IX chromatography

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\*Duolite is a trademark of the Rohm & Haas Company.

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- Disposal of uranium in grout
- Ferric hydroxide precipitation of plutonium
- Removal of iodine from low-level waste or vitrification offgas (Boldt 1995)
- Sodium nitrate purification by clean salt precipitation (see Section D23.0)
- Conversion of sodium hydroxide to sodium carbonate.

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## D2.0 SOLID-LIQUID SEPARATION TREATMENT/ASSUMPTIONS

The CLEAN flowsheet employs centrifuges for solid/liquid separation (SLS) in seven parts of the flowsheet (Figure C-1, Appendix C). These systems are located as follows: after waste tank retrieval (sheet 2, G-1), after organic destruction processing (sheet 4, G-4), after sodium hydroxide dissolution (sheet 8, G-2), after both acid dissolution steps (sheets 10 and 12, G-3 and G-5, respectively), after the chromium reduction step (sheet 35B, G-10) before sodium hydroxide production and recycling, and prior to the HLW melter (sheet 27, G-305). Process specifications which apply to some or all of these systems are listed below.

1. Liquid from the centrifuge would have a solid/liquid weight ratio of 0.001.
2. Supernate entrainment in the solids from the centrifuge is 12 percent of the solid weight in the centrifuge feed.
3. Polyelectrolyte solution is blended in line with the slurry feed to the centrifuge to coagulate colloidal solids and thereby improve solids recovery. This solution would have a weight equivalent to 19 percent of the solids in the centrifuge feed and would have a composition of 1 wt% polyelectrolyte and 99 wt% water.
4. Centrifuge wash water amount would be four times the total solids weight in the centrifuge feed. Of this 400 percent solid weight equivalent wash water, a one-to-one weight ratio of this wash water to solids would accompany these solids and entrained supernate to the centrifuge catch tank; the remainder of the wash water would accompany the clarified supernate.
5. Sufficient dilution water would be added to the centrifuge catch tank to achieve a pumpable slurry containing 10wt% solid content.
6. 0.25 wt% of the clarified liquid from the centrifuge would be recycled to the centrifuge feed tank. The remainder of the liquid would be sent to the inertial filters.
7. The inertial filters would produce a clarified liquid containing 12.5 percent of the total liquids and 1 percent of the total solids from the incoming flow. The remainder of the liquids and solids would be returned to the inertial filter feed tank.

8. 25 percent of the clarified liquid from the inertial filters would be recycled to the inertial filters.

Specifications 1 and 2 apply to all systems. Specification 3 applies to all except the G-305. Specifications 6, 7, and 8 apply to all systems except G-305, G-2, and G-10.

The ESP flowsheet also includes sand (or glass frit) filters to capture solids before all of the ion exchange column processes. These filters are assumed engineering judgment to capture 99.9 percent of the solids. These solids are recycled back to the dissolution process head end. This assumption impacts the amount of solids reading the LLW, affecting the LLW radionuclide concentrations. It also impacts the HLW volume by forcing more solids towards HLW. The addition of the filter glass frit was assumed to be compensated by reducing the glass former addition in the subsequent glass making processes.

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### D3.0 INITIAL SOLID-LIQUID SEPARATION

Sludge washing steps are shown in Figure C-1, sheets 1 and 2. Waste feed from retrieval and transfer would enter receipt and sample tanks (TK-1). The waste would be mixed (TK-2) with purge from an inertial filter, mixed with polyelectrolyte, and sent to a centrifuge (G-1). The polyelectrolyte would coagulate colloidal solids to increase the solids recovery during centrifuging and filtration. The solids would be water washed and collected in the centrifuge catch tank (TK-3). Additional water would be added to produce a pumpable slurry which would be sent to the sodium hydroxide leach tank (sheet 3, TK-31). The centrate would be sent to an inertial filter (FI-1) to remove residual solids and pass through the inertial filter catch tank (TK-9). The clarified liquid from the catch tank then would be sent to the organic destruction step (sheet 3).

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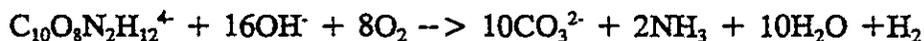
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#### D4.0 ORGANIC DESTRUCTION (ZIMPRO PROCESS)

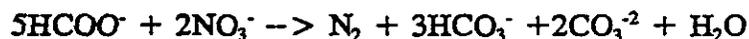
After initial solid/liquid separation of retrieved tank waste, the resulting liquid would undergo wet air oxidation to destroy organics and ferrocyanides as shown in Figure C-1, sheets 3 and 4. This technology has been utilized for 30 years by Zimpro-Passavant Environmental Systems, Inc. in designing commercial processes for organic destruction.

Diluted tank waste liquid from the preceding solid/liquid separation step (Stream 14, sheet 2) and the liquid centrate from sodium hydroxide leaching (Stream 208, sheet 8) would be fed to the organic destruction feed tank (TK-80) and would enter the wet air oxidation process at 25 °C and less than 50 pounds per square inch gauge (psig). A high-pressure pump (P-1) would pressurize feed from this tank to 2,000 psig. The pressurized feed would be heated to approximately 300 °C by a heat exchanger (E-11) that would recover heat from processed tank waste exiting the reactor. The exchanger would recover approximately 90 percent of the heat necessary to bring the feed stream from ambient to reactor temperature (325 °C). After passing through the heat exchanger, the feed would enter a jacketed tubular reactor (TK-81) along with air compressed from ambient to reactor pressure. Heat transfer fluid in the jacket of the reactor would supply the additional heat required to raise the feed stream to the target reactor temperature. The jacketed reactor concept would maintain constant reactor temperature.

At reactor conditions and 29 percent excess air (based on 100 percent organic destruction), oxygen and hydroxide would react with the organic constituents to form carbonate, oxalate, nitrogen, ammonia, and some hydrogen. Organic would decompose according to the following overall reaction stoichiometry based on experimental data for ethylenediaminetetraacetic acid (EDTA) (Schmidt et al. 1994):



Ferrocyanide destruction would proceed according to the following reactions (Schmidt et al. 1993):



A heat exchanger (E-10) would cool the reactor effluent from 325 °C to 50 °C. Two liquid/gas separators in series (T-3, T-4) would reduce the pressure of the cooled processed feed from 2,000 psig to atmospheric pressure. Gases from these two separators would be combined and cooled (EC-102) before being sent to offgas treatment.

A reactor residence time of one hour at 325 °C and 2,000 psig would achieve 99 percent organic and ferrocyanide destruction. Metals complexed with organic in the reactor feed would precipitate as hydroxides upon cooling. Hydroxides of strontium, nickel, calcium, and iron would be likely along with coprecipitation of transuranic (TRU) components and lanthanides. These solids would be removed by centrifuging (G-4) and inertial filter (FI-4) systems and routed to sodium hydroxide leaching (sheet 8, TK-31). The clarified liquid from organic destruction would be combined with the wash raffinate from the actinide extraction with tributyl phosphate (TBP) in the ion exchange evaporator feed tank (TK-89).

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## D5.0 SOLIDS DISSOLUTION

Three dissolution steps, along with undissolved sludge recycle, would minimize the amount of sludge routed to HLW. One dissolution cycle would involve sequential sludge processing through the following steps: (1) sodium hydroxide leach, (2) nitric acid/oxalic acid dissolution, and (3) nitric acid/hydrofluoric acid dissolution. Solid/liquid separation (SLS) would occur after each of these steps. Each of these processes would require several hours of digestion at approximately 90 °C under the following basic/acidic conditions:

Sodium hydroxide Leach:  $[\text{OH}^-] = 4.0\text{M}$

Nitric/Oxalic Acid Dissolution:  $[\text{H}^+] = 4.5\text{M}$   
 $[\text{C}_2\text{O}_4^{2-}] = 0.3\text{M}$  (approx. 1 mole/mole cations dissolved)

Nitric/HF acid Dissolution:  $[\text{H}^+] = 4.5\text{M}$   
 $[\text{F}^-] = 1.0\text{M}$  (approx. 1 mole/mole equivalent of cations dissolved).

The extent of dissolution assumed by component is shown in Table D-1 for each dissolution process.

The solids dissolution steps are shown in Figure C-1, sheets 8 through 12. Feeds to the sodium hydroxide leach tank (TK-31) would include washed solids collected from the solid wash (sheet 2, TK-6) and organic destruction (sheet 4, TK-88) steps, solids recycled from the second acid dissolution step (sheet 12, TK-116), ammonium phosphomolybdate from the acid cesium ion exchange step (sheet 23, X-6), and sodium hydroxide recycled from LLW treatment (sheet 35, TK-10-PL). After digestion by sodium hydroxide, the contents would be sent to a centrifuge (G-2). The centrate (Stream 208) would be routed to the organic destruction feed tank (sheet 3, TK-80).

Solids remaining after sodium hydroxide leaching would be collected (TK-34) and sent to a nitric acid-oxalic acid dissolution step (sheet 9). The solids would be combined with nitric and oxalic acids in a dissolver (TK-29). Strontium-loaded crystalline silicotitanate resin from acid-side treatment (sheet 6, TK-132) would also be fed to this dissolver. Offgas from the dissolver would be treated in a condenser-scrubber offgas treatment system (T-1, T-2, TK-30). This system would convert 90 percent of evolved nitrogen dioxide to nitric acid for recycling to the dissolver. The digested solid-liquid mixture would be sent to a centrifuge (G-3) and filter (FI-5) sequence. The filtrate would be sent to the TBP extraction step (sheet 13) after blending with filtrate from the nitric-hydrofluoric dissolution step.

Solids remaining after the nitric acid-oxalic acid dissolution would be collected (TK-104) and sent to a nitric acid-hydrofluoric acid dissolution step (sheet 11). The solids would be combined with nitric and hydrofluoric acid in a dissolver (TK-110) with an associated offgas

treatment system (T-8, T-9, TK-111) as described for previous acid dissolution step. The digested solid-liquid mixture would be sent to a centrifuge (G-5) and filter (FI-6) sequence. The filtrate would be sent to the TBP extraction step (sheet 13) after blending with filtrate from the nitric-oxalic dissolution step. Solids remaining after the sodium hydroxide, nitric-oxalic, and nitric-hydrofluoric dissolution steps would be collected (TK-114) and sampled (TK-115). In this flowsheet, 5 percent of the remaining solids would be purged to HLW treatment (sheet 24), and the remainder would be recycled to the sodium hydroxide dissolution step (sheet 8).

Table D-1. Component Dissolution by Base/Acid Process. (3 sheets)

Solid phase species	Percent Dissolution		
	NaOH	Nitric/oxalic	Nitric/HF
AG+		90%	90%
AL+3	90%	50%	50%
AL2O3	50%		
AM+3		90%	50%
APM-	90%		
AS+5		90%	90%
B+3		90%	90%
BA+2		90%	90%
BE+2		90%	90%
BI+3		90%	90%
C14		90%	90%
CA+2		90%	90%
CANCRINITE	50%		10%
CD+2		90%	90%
CE+3		90%	90%
CL-		90%	90%
CM+3		90%	90%
CO+3		90%	90%
CO3-2		90%	90%
CR+3		90%	50%
CS+		90%	90%
CU+2		90%	90%
F-		90%	90%

Table D-1. Component Dissolution by Base/Acid Process. (3 sheets)

Solid phase species	Percent Dissolution		
	NaOH	Nitric/oxalic	Nitric/HF
FE+3		90%	50%
FECN6-3		90%	90%
HG+2		90%	90%
I-		90%	90%
K+		90%	90%
LA+3		90%	90%
LI+		90%	90%
MG+2		90%	90%
MN+2		90%	90%
MNO2		90%	90%
MO+6		90%	90%
NA+		90%	90%
NB+5		90%	90%
NI+3		90%	90%
NI2FECN6	90%		
NO2-		90%	90%
NO3-		90%	90%
NP+4		90%	50%
OH-		90%	90%
P2O5:24W		90%	90%
PB+4		90%	90%
PO4-3	90%	90%	90%
PU+4		90%	50%
RB+		90%	90%
RE+7		90%	90%
RH+3		90%	90%
RU+3		90%	90%
SB+5		90%	90%
SE+6		90%	90%
SI+4		90%	90%

Table D-1. Component Dissolution by Base/Acid Process. (3 sheets)

Solid phase species	Percent Dissolution		
	NaOH	Nitric/oxalic	Nitric/HF
SM+3		90%	90%
SN+4		90%	90%
SO4-2	90%	90%	90%
SR+2		90%	90%
TCO4-		90%	90%
TE+6		90%	90%
TH+4		90%	50%
TI+4		90%	90%
TL+3		90%	90%
TOC		90%	90%
UO2+2		90%	90%
V+5		90%	90%
W+6		90%	90%
ZN+2		90%	90%
ZR+4		90%	90%
ZRO2:2H2		90%	50%

The base case assumption (see Table A-1) of 90 percent dissolution for major matrix components of the sludge per pass for the series of three dissolution processes closely parallels the clean option example flowsheet assumptions (Swanson 1993). Engineering judgment, based on experience in B-Plant, is that in multiple passes with a single dissolvent, the fraction of sludge dissolved decreases with each pass. To show the effect of this poor dissolution assumption, the number of passes was limited to three cycles and the percent dissolution was arbitrarily reduced from 90 percent to 10 percent to 5 percent on the first, second, and third passes, respectively. The result of the poor dissolution assumption given in Table A-6 is a large increase in the HLW glass volume. Clearly, verification of the dissolution assumption in the base case will be critical to the success of the Extensive Separations concept and should receive high priority in the experimental program.

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## D6.0 TRIBUTYL PHOSPHATE SOLVENT EXTRACTION OF TRANSURANIC COMPONENTS

The TBP extraction separation step is shown in Figure C-1, sheets 13 through 15. Feed for TBP extraction would consist of clarified dissolver supernatant from the acid dissolution steps (sheets 10 and 12) and acid recycled (sheet 17, TK-130) from the backcycle evaporator in the uranium concentration step. This combined feed would be contacted with TBP in a solvent extractor (CB-5). The solvent composition would be the 30 percent TBP in hydrocarbon diluent as used in the well-known plutonium-uranium extraction (PUREX) process.

In the first extraction contactor (CB-5) of this TBP cycle, uranium, plutonium, neptunium, and thorium would be extracted into the TBP organic phase, which would then be scrubbed with nitric acid to increase the separation from the bulk metal ions, some of which have a slight extractability. The raffinate from this contactor, 3.5M nitric acid containing lanthanides and the remaining components of the dissolved sludge solution, would be steam stripped (EV-1) before being sent to the CMPO extraction step (sheet 18).

The solvent phase in this first extraction contactor would be scrubbed using 0.6M nitric acid and then would be stripped in a reduction step (sheet 14, CB-6) using 0.1M hydroxylamine and 0.2M nitric acid. This contacting would concentrate plutonium, neptunium, and thorium in the acidic water phase (Stream 716) which would be concentrated by evaporation (EV-12) and sent to HLW treatment (sheet 24). The solvent would be stripped a second time (CB-6) using water to remove uranium and using 0.01M 1-hydroxyethyl-1,1-diphosphonic acid (HEDPA) (TK-247) to remove trace TRU not stripped in the previous steps. This would remove uranium, which would be concentrated by evaporation (EV-13). After adding nitric acid to a concentration of 1.4M (TK-129), the concentrate would be sent to uranium purification (sheets 16 and 17). The solvent would be washed using 0.25M sodium carbonate (TM-4) and would be recycled (TK-128). The wash solution would be collected (TK-127) and sent to basic side processing (sheet 4).

Table D-2 presents the percent split of feed components among the various product streams from this process. The dissolved sludge stream would be decontaminated from uranium and neptunium by a factor of  $10^4$  and from thorium and plutonium by a factor of  $10^2$ . Cesium, strontium, americium, and trivalent lanthanides would be inextractable by TBP, and very little technetium would be extracted at the high acidity used here ( $\sim 4\text{M HNO}_3$ ). Thus, these elements would remain in the dissolved sludge stream (Stream 721) leaving this step, along with the nonradioactive elements present in the stream.

For this solvent extraction process as well as other extraction operations described later in this appendix, evaporation steps downstream of the process would recover solvent from the aqueous streams which would contact the organic solvent. Where evaporator condensate contains solvent, this condensate would be sent to a decanter, and the organic phase would be

recycled to the closest upstream solvent extraction system to minimize discharging solvent to water treatment. This is shown at various locations in Figure C-1, but the individual cases are not described in this appendix.

Table D-2. TBP Extraction of Transuranics Section: Component Distribution.

Component distribution among outlet streams					
Species	Stream ID				
	Stream 721	Stream 742	Stream 746	Stream 732	Stream 729
	To Am/La Separation	To HLW	To uranium purification	To basic side feed	Solvent recycle
Am+3	99.99%	.01%			
Bi+3	99.90%	.10%			
Ce+3	99.99%	.01%			
Cm+3	99.99%	.01%			
Fe+3	99.99%	.01%			
La+3	99.99%	.01%			
Np+4	.01%	99.98%	.01%		
Pu+4	1.00%	98.99%	.01%		
Sm+4	99.99%	.01%			
TcO4-	99.90%	.05%	.045%	.005%	
Th+4	1.00%	98.99%	.01%		
UO2+2	.01%	.10%	99.88%	.01%	
Zr+4	99.90%	.10%			
TBP Solvent				.13%	99.87%

## D7.0 URANIUM PURIFICATION

Uranium purification is shown in Figure C-1, sheets 16 and 17. Concentrate from the TBP separation (TK-129, sheet 15) would be contacted (CB-3) with TBP. This contactor would run at low acidity (0.01M nitric acid) and a high degree of saturation of the TBP by uranium. Hydroxylamine reductant would be employed in the scrub stream to enhance removal of fission product and TRU elements. These conditions would lead to a significant uranium "loss" to the contactor raffinate catch tank (TK-51), but the "loss" would be recovered by evaporation (EV-14), collection (TK-130), and recycling of the raffinate to the first TBP extraction separation (TK-120, sheet 13). Finally, the purified uranium would again be stripped into dilute acid (Stream 326). This stream would be concentrated by evaporation (EV-6), sampled, and stored for shipment for use offsite (TK-56). The solvent from this cycle would be washed with sodium carbonate to prepare it for reuse.

The distribution of key components among the outlet streams from this operation is shown in Table D-3.

Table D-3. Uranium Purification Section: Component Distribution.

Component distribution among outlet streams				
Species	Stream ID			
	Stream 357	Stream 336	Stream 351	Stream 344
	To extract transuranic	Uranium to stockpile	To LLW	Solvent Recycle
Np+4	100%			
Pu+4	100%			
TcO <sub>4</sub> <sup>-</sup>	100%			
Th+4	100%			
[UO <sub>2</sub> +2]	5.7%	94.29%	.01%	
Solvent			.05%	99.95%

**D8.0 CMPO SOLVENT EXTRACTION FOR AMERICIUM AND LANTHANIDE REMOVAL**

The CMPO extraction step is shown in Figure C-1, sheets 18 and 19. Raffinate from the TBP extraction step would be adjusted for complexation (TK-39), contacted with CMPO (CB-1) to remove 99.99 percent of the americium and the lanthanides and 99.5 percent of the bismuth, and scrubbed using 0.6M nitric acid. The raffinate, which would contain 2.9M hydrogen ion, would retain essentially all the cesium, strontium, and technetium in the dissolved sludge feed to this step. The raffinate would be sent to the crown ether extraction step (sheets 21 and 22).

The solvent from the extractor would be stripped (contactor CB-2) using 0.05M nitric acid to remove the americium and lanthanum. The strip solution would be concentrated by evaporation (EV-4) and would be sent to an americium/lanthanum band ion exchange separation (sheet 20). Bismuth is retained in the extract under these stripping conditions and would be removed from the solvent in a subsequent sodium bicarbonate wash step (TM-1). The wash solution containing bismuth would be collected (TK-46) and sent to LLW treatment (sheet 26). The washed solvent would be recycled (TK-47). Table D-4 shows the distribution of key components among the outlet streams from this process.

Table D-4. CMPO Extraction Am/La Section: Component Distribution.

Component distribution among outlet streams				
Species	Stream ID			
	Stream 606	Stream 628	Stream 634	Stream 633
	To Sr/Tc extraction	To Am/La Separation	To LLW Processing	Solvent recycle
Am+3	.01%	99.98%	.01%	
Bi+3	.50%	0.50%	99.00%	
Ce+3	.01%	99.99%		
Cm+3	.01%	99.99%		
Cs+	100.00%			
Fe+3	99.99%	.01%		
H+	91.40%	8.60%		

Table D-4. CMPO Extraction Am/La Section: Component Distribution.

Component distribution among outlet streams				
Species	Stream ID			
	Stream 606	Stream 628	Stream 634	Stream 633
	To Sr/Tc extraction	To Am/La Separation	To LLW Processing	Solvent recycle
La+3	.01%	99.99%		
Sm+3	.01%	99.99%		
Sr+2	100.00%			
TcO4-	99.90%	.10%		
	99.90%	.10%		
Solvent			.26%	99.74%

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### D9.0 AMERICIUM/LANTHANIDE SEPARATION BY BAND DISPLACEMENT CATION EXCHANGE

The americium/lanthanum separation is shown in Figure C-1, sheet 20. Concentrate from CMPO stripping (TK-43, sheet 19) would be contacted with a cation exchange resin (X-3) in preparation for separation by band displacement cation exchange. Wheelwright (1974) recommends Dowex 50w, X-8 (100 to 200 mesh) for both the feed adsorption column and the subsequent elution columns where band development of DTPA complexes with the +3 metals occurs. Following a water wash of the resin, the trivalent ions would be eluted with a pH-adjusted 0.05M diethylenetriaminepentaacetic acid (DTPA) solution (TK-257) onto a zinc-loaded cation exchange resin column. Continued elution through a series of columns (X-3) would establish discrete bands of metal ions in a sequence that would depend on the magnitudes of the constants governing the formation of the metal ion-DTPA complexes. This order would be zinc; curium, americium, terbium, and dysprosium (together); gadolinium; europium; samarium; yttrium; and light lanthanides.

The first effluent cut would contain most of the zinc; the second would contain the remainder of the zinc and some of the yttrium along with all of the americium, curium, and lanthanides heavier than promethium; and the third would contain the remainder of the yttrium plus the lanthanides lighter than samarium.

Final removal of light lanthanides from the resin would be accelerated by using a more highly concentrated DTPA solution (TK-256) to strip them. The resin beds then would be regenerated, following a water wash, to prepare them for the next cycle. The first (sorption) bed would be regenerated with  $\text{HNO}_3$ ; subsequent (band displacement) beds would be regenerated with  $\text{Zn}(\text{NO}_3)_2$ . After another water wash, the resin beds would be ready for the next cycle of operation.

The first and third effluent cuts, which would contain most of the zinc, some yttrium, and lanthanides lighter than samarium, would be collected (TK-160) and sent to acidic LLW treatment (sheet 26). The second effluent cut, which would contain some zinc and yttrium and all of the americium, curium, and lanthanides heavier than promethium, would be collected (TK-158) and sent to HLW treatment (sheet 27). Component distributions among the outlet streams from this operation are provided in Table D-5.

Table D-5. Am/La Band Displacement Section: Component Distribution.

Component distribution among outlet streams			
Species	Stream ID		
	Stream 60	Stream 61	Stream 62
	To acidic LLW	To basic LLW	To HLW
Am+3	.01%	.10%	99.89%
Ce+3	.01%	99.89%	.10%
Cm+3	.01%	.10%	99.89%
La+3	.01%	99.89%	.10%
Sm+3	.01%	.10%	99.89%
DTPA-3 (from .05M DTPA strip solution)		99.00%	1.00%
Na+ (from 0.5M DTPA strip solution)		99.00%	1.00%
Zn+2 (from Zn regeneration solution)		99.00%	1.00%

### D10.0 CROWN ETHER SOLVENT EXTRACTION FOR STRONTIUM AND TECHNETIUM REMOVAL

The crown ether extraction separation is shown in Figure C-1, sheets 21 and 22. Raffinate from the CMPO separation (sheet 18, TK-41) would be concentrated by evaporation (EV-7); contacted (CB-4) with a crown ether solvent (0.2M in diluent) to remove strontium, barium and technetium; and scrubbed using 0.6M nitric acid. The raffinate, which is 3.5M in hydrogen ion, would be steam stripped (EV-9) before being sent to a cesium ion exchange separation (sheet 24). The solvent would be stripped using 0.01M nitric acid to remove the strontium, barium, and technetium. The strip solution would be concentrated by evaporation (EV-8) and would be sent to a HLW treatment (sheet 24). The solvent would be washed using sodium bicarbonate (TM-3) and recycled (TK-66). The wash solution would be collected (TK-65) and would be sent to LLW treatment (sheet 26).

As shown in Table D-6, the aqueous dilute acid HLW stream from this extraction step would contain 99.99 percent of the strontium and barium and 99 percent of the technetium in the feed to this operation.

Table D-6. Crown Ether Extraction Sr/Tc Section: Component Distribution.

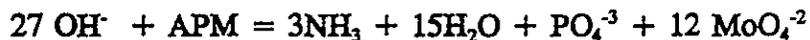
Component distribution among outlet streams				
Species	Stream ID			
	Stream 460	Stream 440	Stream 428	Stream 429
	To HLW Treatment	To Cs APM extraction	To Basic LLW Treatment	Solvent recycle
Ba+2	99.99%	.01%		
Sr+2	99.99%	.01%		
TcO4-	99.00%	1.00%		
Solvent			.05%	99.95%

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### D11.0 AMMONIUM PHOSPHOMOLYBDATE ION EXCHANGE FOR CESIUM REMOVAL

The acidic cesium ion exchange separation is shown in Figure C-1, sheet 23. Raffinate from the crown ether extraction separation step (sheet 22) would be sent to a feed tank (TK-133), filtered (FS-5), and contacted with ammonium phosphomolybdate (X-6) to sorb 99.99 percent of the cesium. Raffinate is sent to LLW treatment (sheet 25, TK-175). The resin would consist of Ammonium Phosphomolybdate (APM) on an alumina substrate (10 wt% APM) to meet the mass and heat transfer requirements for this application. Because cesium cannot be readily eluted from APM, the loaded sorbent (50 grams of cesium per kilogram of APM) would be routed to the sodium hydroxide leach step of the sludge dissolution process (TK-31, sheet 8). Sodium hydroxide conditions would dissolve 90 percent of the alumina, APM and its associated cesium according to the following dissolution reactions, with ultimate cesium removal by basic-side ion exchange (sheet 5, X-1).



**D12.0 RESORCINOL-FORMALDEHYDE ION EXCHANGE FOR CESIUM REMOVAL**

Cesium ion exchange is shown in Figure C-1, sheet 5. A 4-column carousel arrangement would be employed, in which 3 columns in series would be in operation while the fourth column would be undergoing regeneration. Table D-7 provides volumetric information on the regeneration/wash streams required for this ion exchange process.

Evaporation (EV-0) would concentrate basic-side feed (sheet 4, TK-89) and cesium ion exchange wash streams (water and sodium hydroxide) to meet a cesium ion exchange feed specification of 7M NaOH. The ion exchange feed (TK-11) would be filtered (FS-1) and contacted with a resorcinol-formaldehyde ion exchange resin (X-1) to achieve 99.61 percent cesium removal. The column raffinate would be sent to strontium ion exchange (sheet 6, TK-152).

For regeneration, the resin would be washed using 2M NaOH and water, then eluted with 1M formic acid. The eluted cesium would be sent to HLW treatment (sheet 24). Next the resin would be washed with water and regenerated with 2M NaOH. These water and sodium hydroxide wash streams would be recycled to the evaporator (EV-0) upstream of this ion exchange operation.

Table D-7. Basic Cesium IX: Relative Flow Volumes of Streams.

Stream name	Stream number	Column volumes	Composition
Column Feed	102	180	---
Regeneration Streams			
Feed flush	113	3	2M NaOH
Sodium hydroxide flush	110	6	Water
Cesium Eluant	108	20	1M Formic acid
Eluant Flush	117	3	Water
Regeneration	115	3	2M NaOH

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### D13.0 CRYSTALLINE SILICOTITANATE ION EXCHANGE FOR STRONTIUM REMOVAL

Strontium ion exchange is shown in Figure C-1, sheet 6. Raffinate from the cesium ion exchange step would be sent to a feed tank (TK-152), filtered (FS-6), and contacted with crystalline silicotitanate (CST) cation exchange resin (X-6) to achieve 99.6 percent strontium removal. The raffinate would be sent to technetium ion exchange (sheet 7, TK-93).

The CST bed would not be regenerated since strontium cannot be eluted from the resin. The strontium loaded resin (4 wt% strontium loading for engineered CST) would be recycled to the acid dissolution section (sheet 9, TK-29). Dissolution of the resin and associated strontium (90 percent dissolution per pass) would occur in the nitric acid-hydrofluoric acid dissolver (sheet 11, TK-110). The nominal composition of the engineered CST would be 32 wt%  $TiO_2$ , 16 wt%  $Na_2O$ , 44 wt%  $SiO_2$  and 8 wt% water.

The CST will coexchange plutonium and cesium along with strontium at approximately the same extent as the strontium when the lower cesium and plutonium concentrations are considered. However, only the strontium exchange is represented in the flowsheet.

The strontium removal assumptions impact whether the LLW radiotoxicity objective is met, and also impact the HLW volume because CST material adds oxides of sodium, titanium, and silicon during vitrification. The additional removal of cesium and plutonium could also aid in meeting the LLW radiotoxicity objective, and while not represented in the flowsheet, will be addressed in the presentation of the LLW product composition results (see Table 9-1) relative to the radiotoxicity limits.

**D14.0 ANION EXCHANGE REMOVAL OF TECHNETIUM**

Technetium ion exchange is shown in Figure C-1, sheet 7. Raffinate from the basic-side strontium ion exchange step would be sent to a feed tank (TK-93), filtered (FS-2), and contacted with an anion exchange resin (X-2) to achieve 99.5 percent technetium removal as pertechnetate ion ( $TcO_4^-$ ). Table D-8 provides volumetric information on the regeneration/wash streams required for this ion exchange process.

The *Tank Waste Technical Options Report* (Boomer et al. 1993) contains a summary of feed and regeneration rates for technetium ion exchange. Using a strong base anion exchange resin, 100 stream or column volumes of feed were assumed processed to reach breakthrough. This data is based on experimental work done by Schultz (1980). That work used  $NO_3^-$  form of three different resins: Dowex 1x4, Dowex MSA-1, and IRA 401. Results are provided in the body of that report for only the last two of these three resins.

The loaded resin bed would be washed with sodium hydroxide to remove residual feed solution, then with water in preparation for acid elution. A dilute nitric acid (0.25M) would be used first to ensure conversion of the resin to the nitrate form before technetium elution begins. The technetium would be eluted with 6M nitric acid. The bed would be prepared for the next loading cycle by displacing the acid first with water, then with sodium hydroxide.

The eluted technetium would be concentrated by evaporation (EV-10) and sent to HLW treatment (sheet 24). Vapor from the evaporation would be sent to a fractionator (T-5) where nitric acid would be recovered and recycled (TK-101) for use in the technetium ion exchange step. Raffinate would be sent to basic LLW treatment (sheet 26, TK-139).

Table D-8. Basic Technetium IX: Relative Flow Volume of Streams.

Stream Name	Stream Number	Column Volumes	Composition
Column Feed	500	100	---
Regeneration streams			
Feed flush	504	8	2M NaOH
Sodium hydroxide flush	507	12	Water
OH <sup>-</sup> displacement	513	6	0.25M HNO <sub>3</sub>
Technetium Eluant	515	20	6M HNO <sub>3</sub>
Eluant Flush	518	6	Water
Regeneration	533	8	2M NaOH

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## D15.0 STRONTIUM AND CESIUM CAPSULE DISSOLUTION AND METATHESIS (HIGH-LEVEL WASTE OPTION 2)

The cesium/strontium capsules would be transferred in a Beneficial Uses Shipping System cask from the waste encapsulation storage facility to a capsule cut-up cell in the extensive separations complex. After delidding using a cask lid/delid machine (CN15), the cask would be transferred to the hot cell.

Once in the hot cell, inner capsules would be transferred from the turnstile to the end crop machine (FX7) using a manipulator or tongs. After end cropping, the cesium capsules would be transferred to the cesium rinse tank (TK21) via a basket (BB9). Dissolved cesium would be transferred from the cesium collection tank (TK22), to ion exchanger (IX25) to remove the chloride ions prior to blending with the main HLW stream.

To prevent diffuse breakthrough of chloride, the anion exchanger must have a selectivity for chloride greater than 1.0. A bed composed of 50/100 mesh Dowex<sup>TM</sup>\* 2x8 in the hydroxyl form would achieve this purpose (Helfferich 1962). The capacity of the exchanger on a mole equivalent basis would only need to be 1.5 times the equivalents of chloride in the CsCl feed, since there are no other anions in the feed stream. One bed volume of wash water would be adequate to rinse the exchanger free of cesium ions. The resin would not be regenerated, but the chloride loaded bed would be discarded periodically and mixed into the LLW glass melter feed stream. The chloride concentration due to resin would be very low; therefore, it would not affect LLW melter operations or the quality of the glass product.

Strontium fluoride capsule contents would be removed by a hydraulic press on the end crop machine table and transferred to a slurry tank (TK20) via roll crushers. The strontium fluoride would then be transferred to HLW for blending and vitrification.

Outer and inner capsule hulls would be decontaminated in tank TK24 and shredded for size reduction (FX13). The shredded material would be drummed and removed from the facility for burial.

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\*Dowex is a trademark of the Dow Chemical Company.

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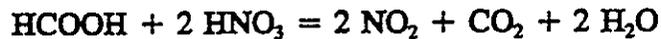
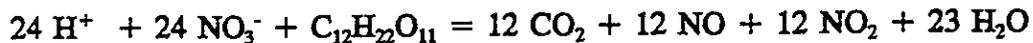


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## D16.0 HIGH-LEVEL WASTE STREAM VITRIFICATION TREATMENT

### D16.1 HIGH-LEVEL WASTE CONCENTRATION/DENITRATION

High-level waste stream treatment preceding vitrification is shown in Figure C-1, sheet 24. The HLW streams, from the separation steps described in the previous sections (undissolved sludge purge from acid dissolution, strontium and technetium from crown ether extraction, americium and lanthanides from band ion exchange, transuranics from TBP extraction, and technetium from basic side ion exchange) would be combined, (TK-141) and a concentrated nitric acid liquid (specific gravity=1.37) would be produced by evaporation (EV-16). The HLW evaporator would operate under a vacuum in order to remove the nitric acid. Dilute acid vapor from this operation would be sent to acid recovery (sheet 25, TK-142). The concentrated nitric acid solution and formic acid eluant containing cesium from resorcinol-formaldehyde ion exchange would be mixed (TK-149) and would undergo denitration by reaction with sucrose and formic acid destruction.



Sufficient 2M sucrose would be supplied to this denitration step to achieve 0.5M HNO<sub>3</sub> in the liquid after complete sucrose conversion. This liquid, containing undissolved solids from the acid dissolution operation, would be fed to the HLW centrifuge feed tank (sheet 27, TK-302). The NO<sub>x</sub> produced by denitration would be sent to the acid recovery portion of the flowsheet (sheet 25) for conversion back to HNO<sub>3</sub>.

### D16.2 HIGH-LEVEL WASTE VITRIFICATION

The denitrated HLW solid slurry would be routed to the HLW centrifuge. The centrifuged HLW solids (assumed 20 wt%) would be sent to the melter feed section; the liquid would go to a HLW evaporator. The HLW evaporator would concentrate the stream to decrease the amount of water entering the HLW melter. The evaporator bottoms would be routed to the melter feed section.

In the melter feed section, the solids from the HLW centrifuge and the melter offgas processing (MOG) treatment recycle would be combined together. Glass formers and an organic acid (glycolic) would be added to this stream before it enters the melter section. The organic acid would be added for reduction/oxidation control in the melter. The melter would be a joule-heated, liquid-slurry-fed glass melter. The molten glass would exit the melter and proceed to the canister filling and handling section. The MOG would be routed to the MOG treatment section.

A high-temperature or a stirred melter would convert the incoming feed slurry into molten borosilicate glass containing > 50 wt% waste oxides (based on glass composition optimization calculations in the flowsheet simulation). Volatilized melter feed components would form a separate stream of offgas that passes overhead. The hot glass would be semi-continuously poured into cylindrical, stainless steel canisters.

The melter used in the HLW vitrification process would be a joule-heated design that would operate at 1200 °C. The composition and flowrate of glass formers (containing primarily SiO<sub>2</sub>, B<sub>2</sub>O<sub>3</sub>, Fe<sub>2</sub>O<sub>3</sub>, Li<sub>2</sub>O, Na<sub>2</sub>O, Al<sub>2</sub>O<sub>3</sub>, and CaO) would be adjusted to minimize total HLW glass weight subject to the following glass constraints:

SiO <sub>2</sub>	content	≥	42	wt%	≤	57	wt%
B <sub>2</sub> O <sub>3</sub>	content	≥	5	wt%	≤	20	wt%
Na <sub>2</sub> O	content	≥	5	wt%	≤	20	wt%
Li <sub>2</sub> O	content	≥	1	wt%	≤	7	wt%
Fe <sub>2</sub> O <sub>3</sub>	content	≥	2	wt%	≤	15	wt%
CaO	content	≤	10	wt%			
MgO	content	≤	8	wt%			
Al <sub>2</sub> O <sub>3</sub>	content	≤	15	wt%			
ZrO <sub>2</sub>	content	≤	13	wt%			
Cr <sub>2</sub> O <sub>3</sub>	content	≤	0.5	wt%			
P <sub>2</sub> O <sub>5</sub>	content	≤	3	wt%			
SO <sub>3</sub>	content	≤	0.5	wt%			

The separation of glass/offgas was assumed to send 100 percent gas and 1 percent solids to HLW offgas treatment with 99 percent solids to glass. Glass rework of off-specification material would be processed through a separate cyclone/roll crusher/catch tank to prepare a 20 wt% solid slurry which would be recycled to the HLW melter.

The glass-filled canisters would be plugged and welded closed before being decontaminated to remove exterior contamination. The spent decontamination liquids would be accumulated and recycled to the feed preparation system to evaporate excess water and recover contaminants. The decontaminated canisters, filled with monolithic HLW glass, would be placed into an overpack container (four canisters per overpack). The overpack containers would be transferred from the vitrification building to an interim storage building while awaiting eventual shipment to a federal geologic repository.

The high-activity glass vitrification process would be relatively independent of the type of melter employed. Most of the data used for sizing and costing purposes for this document used a combustion melter for high-activity glass vitrification. This would impact the high-activity melter's offgas equipment but would have a small impact on the volume of

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offgas when compared to the low-activity melter's offgas. Also, the masses of components in the offgas stream and the volume of glass produced would not change with the melter type.

Rather than the oxide-specific glass composition limits given above and used in this report, a simpler composition limit of 25 wt% total waste oxides is specified in the *TWRS Process Flowsheet* (Orme 1994). This limit is the so-called low temperature glass limit which is used in the *Tri-Party Agreement Alternative Data Package for the Tank Waste Remediation System Environmental Impact Statement* (Slaathaug 1995). This limit may be more conservative and reduce the likelihood of devitrification by reducing the glass temperature. The results of an Extensive Separations case using this limit is shown in Table A-5 for comparison purposes.

### D16.3 HIGH-LEVEL WASTE MELTER OFFGAS PROCESSING

The HLW MOG systems would receive the hot gases from the melter and subject them to a water quench, a Venturi scrubber/separator, a demister, and a high-efficiency particulate air (HEPA) filtering system. Quenching these offgases by countercurrent contact with cool scrub water (75 °C) would remove most of the entrained particulates and water-soluble contaminants and would condense much of the water vapor. Excess condensates from the HLW MOG system would recycle to the HLW feed preparation system for re-evaporation. Excess process condensates would be continuously purged to the process liquid waste system while quenched offgas would pass through a mist eliminator and multiple stages of high-efficiency filtration where most of the remaining radionuclides would be captured. The scrubbed MOGs would flow to a catalytic de-NO<sub>x</sub> reactor before final discharge to the heating, ventilating and air conditioning (HVAC) system. The treated MOG would combine with the vessel offgas before passing through NO<sub>x</sub> abatement equipment enroute to the HVAC system.

Overall separation of gas from condensed phases achieved by these processing steps was assumed to be 99.999+ percent solid oxide removal, 100 percent Hg removal, 50 percent NH<sub>3</sub> removal, 100 percent TcO<sub>2</sub> removal, with conversion to TcO<sub>4</sub><sup>-</sup>(aq), and 99+ percent water removal. Condensed stream recycle would send the solid slurry to the LLW pretreatment evaporator and the liquid to HLW pretreatment evaporator.

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## D17.0 INITIAL TREATMENT FOR ACIDIC LOW-LEVEL WASTE TREATMENT

### D17.1 LOW-LEVEL WASTE CONCENTRATION

The LLW concentration portion of the flowsheet is shown in Figure C-1, sheets 25 and 26. The LLW streams from the separation steps described in the previous sections (solvent wash waste from uranium purification, solvent wash waste from CMPO extraction, Am/La band ion exchange elution waste, solvent wash waste from crown ether extraction, and raffinate from base-side ion exchanges) would be combined (TK-139) and concentrated (EV-15) by evaporation of water to a 7M NaOH solution. This concentrated LLW stream from evaporation would be fed to the LLW calcination process while the basic evaporator overheads would be allocated to dilution water for the salt grout process and/or combined (TK-145) with slightly acidic evaporator overheads from the separation steps described in the previous sections (evaporator overheads from crown ether extraction, from CMPO extraction, from uranium purification, from TBP extraction, from organic destruction, and from acid recovery) and recycled to the wash operations for the former separation processes. The LLW waste would be processed through an evaporator at 76 liters per minute to provide an evaporated feed of 15.1 liters per minute to the calciner.

### D17.2 LOW-LEVEL WASTE CALCINATION (T-THERMAL PROCESS)

The LLW calcination portion of the flowsheet is shown in Figure C-1, sheet 34. Low-activity waste feedstreams to the calcination process (TK-4-PL) would consist of the liquid from the basic LLW evaporator (sheet 26, EV-15) and the denitrated liquid from the acidic LLW evaporation (sheet 25, EV-20) of the raffinate streams from Am/Ln band ion exchange and APM cesium ion exchange. These streams would be fed to the calcination/T-Thermal process for primarily nitrate destruction.

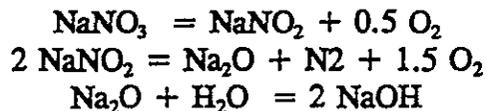
Destruction of nitrate/nitrite salts, organics, and ferrocyanide, in addition to solubilizing aluminum compounds would be accomplished using a modified plasma arc calcination (Hendrickson 1994) process known as the T-Thermal process. The main modification is related to using ammonia as the combustion fuel. The amount of ammonia required would be based on calculating the sensible heat required to raise the reactants to a temperature of ~800 °C. The major process steps would consist of waste calcination, product handling, and offgas treatment. Preliminary tests at the Westinghouse Science and Technology Center indicated an operating temperature range of 750 - 850 °C for the plasma arc calciner operation. However, further temperature optimization must be done together with kinetics studies.

The flowsheet equipment used assumptions of 0.1 second residence time and an 800 °C operating temperature, under atmospheric pressure to provide 91 to 86 percent nitrate/nitrite decomposition and 99 percent organic decomposition. The percent decomposition, not the residence time, is based on engineering judgment of the of the Westinghouse Science and

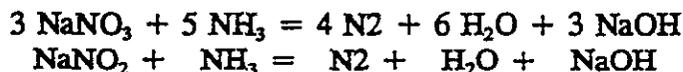
Technology Center test data (Hendrickson 1994, Colby et al. 1994, and McLaughlin 1994). The residence time assumption is engineering judgment that would require further test data to optimize. It was assumed that 100 percent excess  $\text{NH}_3$  was supplied, where unreacted  $\text{NH}_3$  was oxidized to nitrogen and water.

The high operating temperatures would oxidize the waste resulting in the destruction of organics and oxidation of metals which would produce a fluid molten salt, primarily sodium hydroxide (NaOH). The oxidation produces gaseous decomposition products that would consist of primarily  $\text{N}_2$ ,  $\text{O}_2$ , and water vapor with small amounts of  $\text{NO}_2$ ,  $\text{NO}$ ,  $\text{CO}_2$ , and  $\text{CO}$ . The molten salt, carried by the gas stream, would be redissolved in an integral water quench.

The reaction stoichiometry for conversion to sodium hydroxide would be described by the following reactions, where 91 and 86 percent conversion of nitrate and nitrite respectively would be assumed:

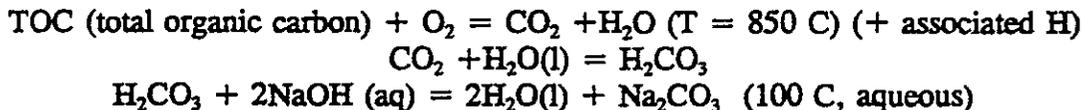


The remaining sodium nitrate and sodium nitrite would decompose to  $\text{NO}_x$ , which would be destroyed by addition of ammonia. The overall calcination/ $\text{NO}_x$  destruction stoichiometry was described as follows:



It was assumed that 100 percent excess  $\text{NH}_3$  was supplied, where unreacted  $\text{NH}_3$  is oxidized to nitrogen and water.

After condensing the reactions products, nitrogen gas would be separated from the sodium hydroxide slurry (7M NaOH). The calcination would also convert the NaOH to  $\text{Na}_2\text{CO}_3$  by action of  $\text{CO}_2$  gas generated by organic decomposition shown by the reactions below.



The equipment for the offgas system would consist of a liquid/vapor separator, Venturi scrubber,  $\text{NO}_x$  reactor, condensers, heaters, two stages of HEPA filters, and stack. To be conservative, a 5 percent entrainment of cesium and technetium was assumed for the quench tank offgas.

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### D17.3 DISSOLUTION/CONCENTRATION

The calciner molten salt stream, carried by the gas stream, would be redissolved in an integral water quench. Originally, the water-insoluble stream was expected to contain a majority of the transuranic (TRU) isotopes, the multivalent metal oxides and hydroxides as well as the aluminates. However, literature data suggest that the TRUs may become soluble under process conditions (Delegard et al. 1993).

The quench solution would be about 85 °C and would remove particles two microns or larger from the gas stream, some of the soluble gases, and 95 percent of the cesium and technetium. Solubility data from testing performed at Los Alamos National Laboratories indicate 25 to 30 wt% of the calcine product is insoluble (Hendrickson 1994). Thirty wt% insoluble aluminum was assumed for the flowsheet.

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## D18.0 GLASS IN SULFUR (LOW-LEVEL WASTE OPTION A)

### D18.1 LOW-LEVEL WASTE VITRIFICATION

The type of melter selected would have its largest impact on the volume of offgas produced but generally would not impact the masses of the components in it. The melter models have not been developed to such an extent that individual component carryover in the offgas streams are based upon the melter type. For example, every melter would have the same percent of sodium fed to it in its offgas regardless of its (combustion, joule, etc.). Also, the glass produced would meet the same criteria regardless of the melter type.

The melter used in the LLW vitrification process was represented as a combustion (kerosene fuel) driven design that would operate at a temperature of 1200 °C. The feed slurry and the dry glass formers would be added to the melter simultaneously. The composition and flowrate of glass formers (containing primarily SiO<sub>2</sub>, Al<sub>2</sub>O<sub>3</sub>, and CaO) would be adjusted to minimize total LLW glass weight subject to the following glass constraints:

Na <sub>2</sub> O content	≤ 25 wt%
Al <sub>2</sub> O <sub>3</sub> content	= 5 wt%
CaO content	= 10 wt%
SiO <sub>2</sub> content	≥ 50 wt%

The separation of glass/offgas was assumed to send 100 percent gas and 1 percent solids to LLW offgas treatment with 99 percent of solids to quench. The glass quench water would be 2,000 wt% of glass. The glass screening was assumed to use screen wash water at 12 wt% of the glass with the screened solid containing 2 wt% liquid where the liquid from screening operation would be recycled to quench. The glass cullet was assumed to be saturated at 30 °C when loaded in the drying bin. The inlet air was assumed to be water saturated at 15.5 °C with the cullet carryover to drying air assumed to be 0.1 percent. Air would be passed through cullet bins to remove 99 percent of moisture from the cullet. Outlet air would be water saturated at 30 °C after contacting cullet. Water would be condensed from the outlet air and the air would be sent to offgas treatment.

Cullet transfer to sulfur processing would require 7.16 kilogram of air per kilogram solid to fluidize the cullet. The cyclone was assumed to recover 99.999 percent of fluidized cullet for cask loading with the air from the cyclone sent through filters in offgas treatment. Glass rework of off-specification material would be processed through a separate cyclone/roll crusher/catch tank to prepare a 20 wt% solid slurry which would be recycled to the LLW melter. Glass in the sulfur product mixture was assumed to be 67 wt% oxides and 33 wt% sulfur (60/40 by percent volume). The glass/sulfur molten stream recycle from molten vault to vault surge tank would aid in control of vault surge tank temperature. Oligimer and Dicyclopentadiene each would constitute 2.5 vol% of sulfur feed stream.

The sulfur polymer cement would be composed principally of elemental sulfur with additions of plasticizing agents to inhibit crystal growth and to control polymerization and viscosity.

### D18.2 LOW-LEVEL WASTE MELTER OFFGAS PROCESSING

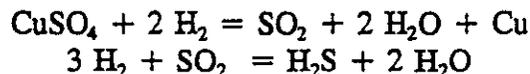
Low-level waste melter offgas would be sequentially processed through the following steps to remove condensables and solid carryover: water quenching (75 °C), Venturi scrubber/separator (75 °C) processing, demisting, and HEPA filtering. Overall separation of gas from condensed phases achieved by these processing steps was assumed to be 99.999+ percent solid oxide removal, 100 percent Hg removal, 50 percent NH<sub>3</sub> removal, 100 percent TcO<sub>2</sub> removal, with conversion to TcO<sub>4</sub><sup>-</sup>(aq), and 99+ percent water removal. Condensed stream recycle would send the solid slurry to the LLW pretreatment evaporator and the liquid to the HLW pretreatment evaporator. Remaining offgas would be sent to SO<sub>2</sub> removal/NOx destruction processing.

### D18.3 SO<sub>2</sub> REMOVAL AND SULFUR RECOVERY

Gas from melter offgas processing would undergo SO<sub>2</sub> removal by adsorption on CuO beds before NOx destruction would take place as described in the next section. The SO<sub>2</sub> would be desorbed, converted to H<sub>2</sub>S, and then sent to Claus units for sulfur recovery. The CuO bed was assumed to have an operating temperature of 400 °C with 90 percent absorption of SO<sub>2</sub> by the reaction:



A 10 percent excess O<sub>2</sub> feed was assumed along with 100 percent excess CuO in bed. The NH<sub>3</sub> cracker that supplies H<sub>2</sub> for bed desorption of SO<sub>2</sub> was assumed to have a 99 percent cracking efficiency of NH<sub>3</sub>, where no excess NH<sub>3</sub> beyond stoichiometric H<sub>2</sub> requirement was assumed. The CuO bed operation for SO<sub>2</sub> desorption was assumed to have a 100 percent conversion for the following reactions:

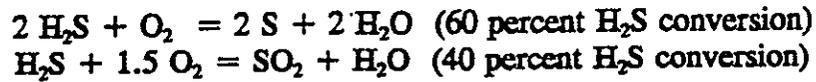


The CuO bed regeneration was assumed 100 percent regeneration of Cu to CuO per reaction:

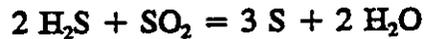


The air for regeneration was based on three times the stoichiometric O<sub>2</sub> requirement.

The sulfur recovery process would mix the H<sub>2</sub>S gas mixture from SO<sub>2</sub> desorption with air and heats the mixture to 800 °C where the following reactions achieve 100 percent H<sub>2</sub>S conversion:



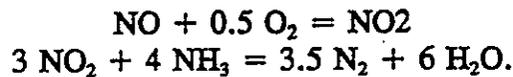
The remaining gas mixture would be cooled to 300 °C and sent through two Claus units in series with liquid sulfur dropout before and after each Claus unit. Each Claus unit achieves 95 percent conversion of H<sub>2</sub>S to sulfur by the reaction:



A 100 percent separation efficiency was assumed for sulfur removal from gas.

#### D18.4 NO<sub>x</sub> DESTRUCTION PROCESSING

The reaction conditions for NO<sub>x</sub> destruction was assumed to have an operating temperature of 500 °C with no excess NH<sub>3</sub>. The process was modeled by the following reactions assuming 99 percent conversion:



#### D18.5 OFFGAS TREATMENT SYSTEM

A set of HEPA filters (99.95 percent fines removal) would treat offgas from the following portions of the flowsheet before gas would be sent to the stack: plasma arc calciner (Stream 1543, HLW process sheet 35A and B of Figure C-1), salt grout process (Stream 15, grout process sheet 1 of Figure C-3), LLW melter offgas (Stream 907, LLW process sheet 9 of Figure C-2), HLW melter offgas (Streams 1371 and 1374, HLW process sheet 31 of Figure C-1).

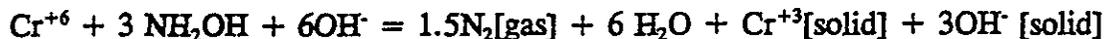
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## D19.0 LOW-LEVEL WASTE STREAM GROUT TREATMENT (LOW-LEVEL WASTE OPTION B)

### D19.1 CHROMIUM REDUCTION

To address the issues of chromium reduction and removal (see discussion in Section 4.7.5), the ESP flowsheet calculations included a grout product calculation with and without a prior chromium reduction and removal process. Based on the CLEAN flowsheet (Swanson 1993), a 1.5M  $\text{NH}_2\text{OH}$  reductant would be used to reduce soluble  $\text{Cr}(+6)$  to insoluble  $\text{Cr}(+3)$ . The proposed reaction was as follows:



where all species above would be in solution unless denoted otherwise. Since the flowsheet mass balance calculations did not distinguish between the two different valence states of chromium, it was assumed that all chromium to this reduction step is  $\text{Cr}^{+6}$ .

The reaction extent was assumed to achieve 99 percent Cr reduction. A subsequent venting step would be included to allow for separation of evolved nitrogen gas.

Insoluble chromium would be removed after this reduction step by a centrifuge system and sent to a separate waste processing step. Approximately 90 percent of the centrate (7M NaOH) would be recycled to the sodium hydroxide leach tank (sheet 8, TK-31) to achieve 4M NaOH in the leaching tank. The remainder of the centrate would feed the salt grout LLW process.

### D19.2 SALT GROUT

The concept for the salt grout process is to mix a waste slurry with cementitious materials to form a grout that can be pumped into removable tubes placed retrievably onsite in underground concrete vaults (5,300  $\text{m}^3$  volume). The grout formulation is 1,100 grams dry mix per 1 liter waste. The dry mix used in the grout flowsheet consists of 40 wt% limestone, 28 wt% fly ash, 28 wt% slag, and 4 wt% Portland cement (Hendrickson 1992 and Serny et al. 1989). Grout formulations to date have been limited to 3M total nitrate and nitrite in the feed along with a 5M sodium content. These two constraints are present in the grout flowsheet. Further discussion of grout formulation issues is provided in Section 4.7.3.

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## D20.0 SODIUM HYDROXIDE RECYCLING

The sodium hydroxide slurry produced in the calcination operation would be concentrated by evaporation (sheet 35, EV-OH) to 7M sodium hydroxide, which can be recycled to meet sodium hydroxide dissolution process needs and thereby minimize LLW disposal volume. To eliminate chromium from recycled sodium hydroxide, soluble Cr(VI) is reduced to insoluble Cr(III) with hydroxylamine. After solid-liquid separation, approximately 90 percent of this sodium hydroxide stream must be recycled to meet flowsheet requirements for sodium hydroxide (e.g., the sodium hydroxide leach step of the solid dissolution operation). The remaining portion of the sodium hydroxide slurry is incorporated into a LLW form for disposal.

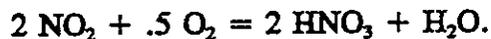
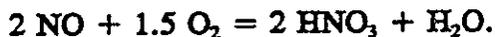
In the Extensive Separations base case (see Table A-1), a fraction of the sodium hydroxide product from the LLW calciner would be recycled without further treatment. The recycle stream contains part of the solids that were formed in the calciner but which had previously been decontaminated to NRC Class A levels. To determine the effect of not recycling these solids on the quantity of HLW glass, a centrifuge was placed downstream from the calciner to remove 99.9 percent of the solids. The net result of the calculation (see Table A-2), is to greatly reduce the volume of HLW glass by reducing the volume of solids including the volume limiting  $P_{205}$  that had come from the recycled sodium hydroxide.

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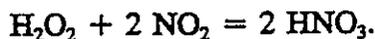
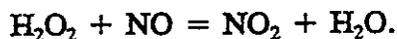
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## D21.0 NITRIC ACID RECYCLING

Acidic evaporator overheads (from HLW treatment, strontium ion exchange, and acidic LLW treatment) are contacted with NO<sub>x</sub> streams (from denitration and calcination steps) with the addition of hydrogen peroxide and air to absorb and convert NO<sub>x</sub> to nitric acid. Conversion of 80 percent NO<sub>x</sub> to acid is accomplished by oxidation by oxygen.



The remaining 20 percent of the NO<sub>x</sub> is oxidized by hydrogen peroxide.



The resulting dilute acid liquid is concentrated by acid fractionation to recover 99.5 percent of the nitric acid as 12M HNO<sub>3</sub>. The weak acid overhead is used as water recycle.

## **D22.0 WATER RECYCLING**

Water from the numerous evaporators throughout the flowsheet would ultimately be routed to wash water tank TK-145, sheet 26. The dilution water requirements of the various separations processes would be supplied from this wash water tank.

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## D23.0 CLEAN SALT PROCESS

The clean salt process (CSP) is an optional process shown in Figure 4-2. The results of the preliminary assessment of the clean salt process documented in the *Clean Salt Integrated Flow Sheet* (Lunsford 1994), estimate that the process has the potential to reduce the LLW volume by 90 percent. The primary salts produced by the process are sodium nitrate and aluminum nitrate. It is estimated that a total of 65,000 to 75,000 metric tons of sodium nitrate and 1,800 to 3,200 metric tons of aluminum nitrate will be produced by CSP (Herting 1993, Lunsford 1994). The salts are expected to be extracted and decontaminated to very low levels of radioactivity. The primary constituent of concern is cesium-137. The degrees to which these salts are decontaminated is dependent upon the design and operations of the CSP. Varying levels of decontamination can be achieved by increasing the number of recrystallization stages applied to the waste steam. Current efforts indicate that CSP is a viable method to achieve volume reduction, and therefore provide significant cost savings, especially in the LLW stream.

A number of potential disposition options for the product material have been evaluated, and the issues arising from the regulatory framework that affect each disposition option have been defined (Herting 1995).

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**APPENDIX E**

**FACILITY LAYOUTS**

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**APPENDIX E****FACILITY LAYOUTS**

This appendix includes the following facility layouts.

Figure E-1. Extensive Separations Pretreatment and High-Level Waste Treatment Facility Layout.

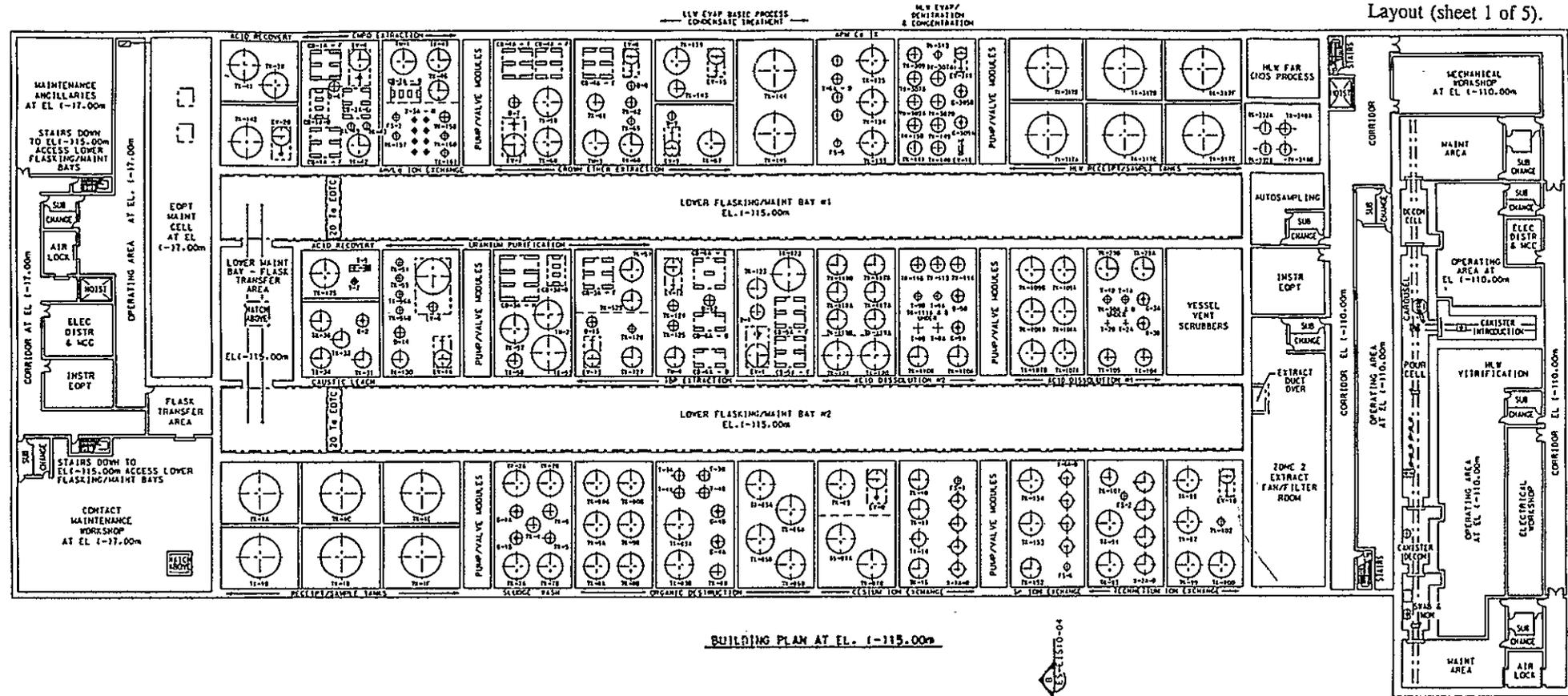
Figure E-2. Low-Level Waste Vitrification Facility Layout.

Figure E-3. Low-Level Waste Grout Facility Layout.

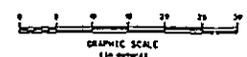
Figure E-4. Cesium and Strontium Cut-Up Cell Layout.

Figures E-1 through E-4 provide facility layouts for the integrated extensive separations pretreatment and high-level waste vitrification treatment facility, low-level waste vitrification treatment facility, low-level waste grout treatment facility, and cesium and strontium capsule pretreatment, disassembly, and metathesis cell, respectively. The equipment lists for these facilities/cell are in Appendix B.

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BUILDING PLAN AT EL. 1-115.00m



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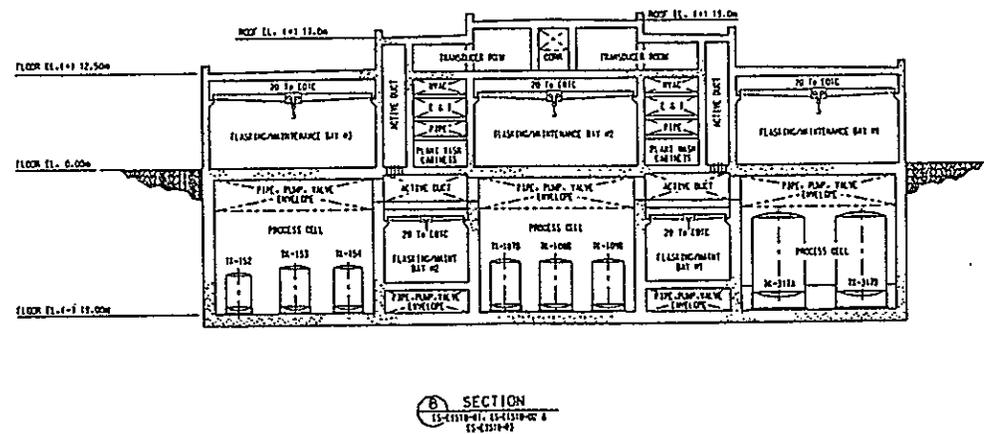
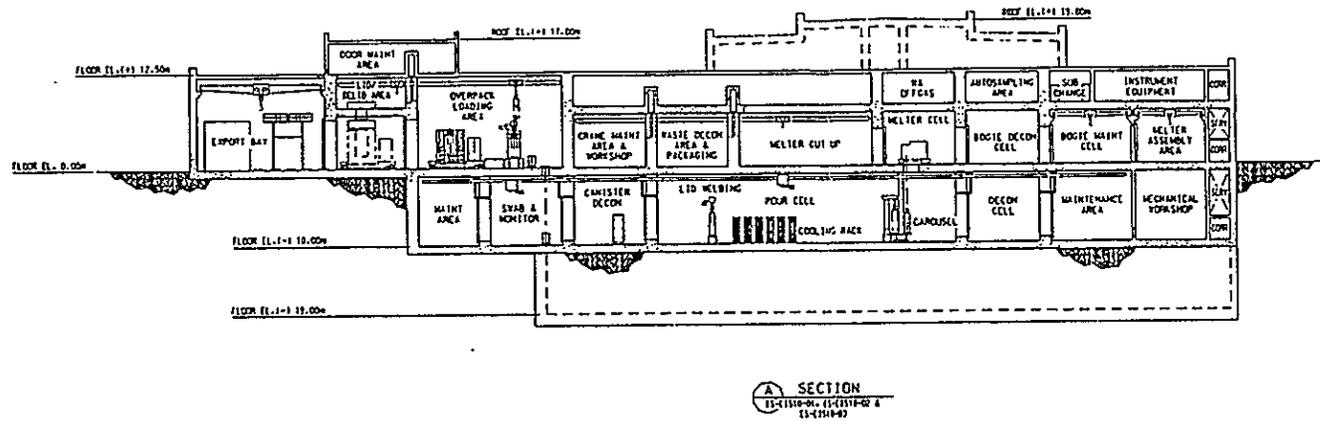
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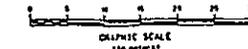
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Figure E-1. Extensive Separations Pretreatment and High-Level Waste Treatment Facility Layout (sheet 4 of 5)



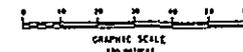
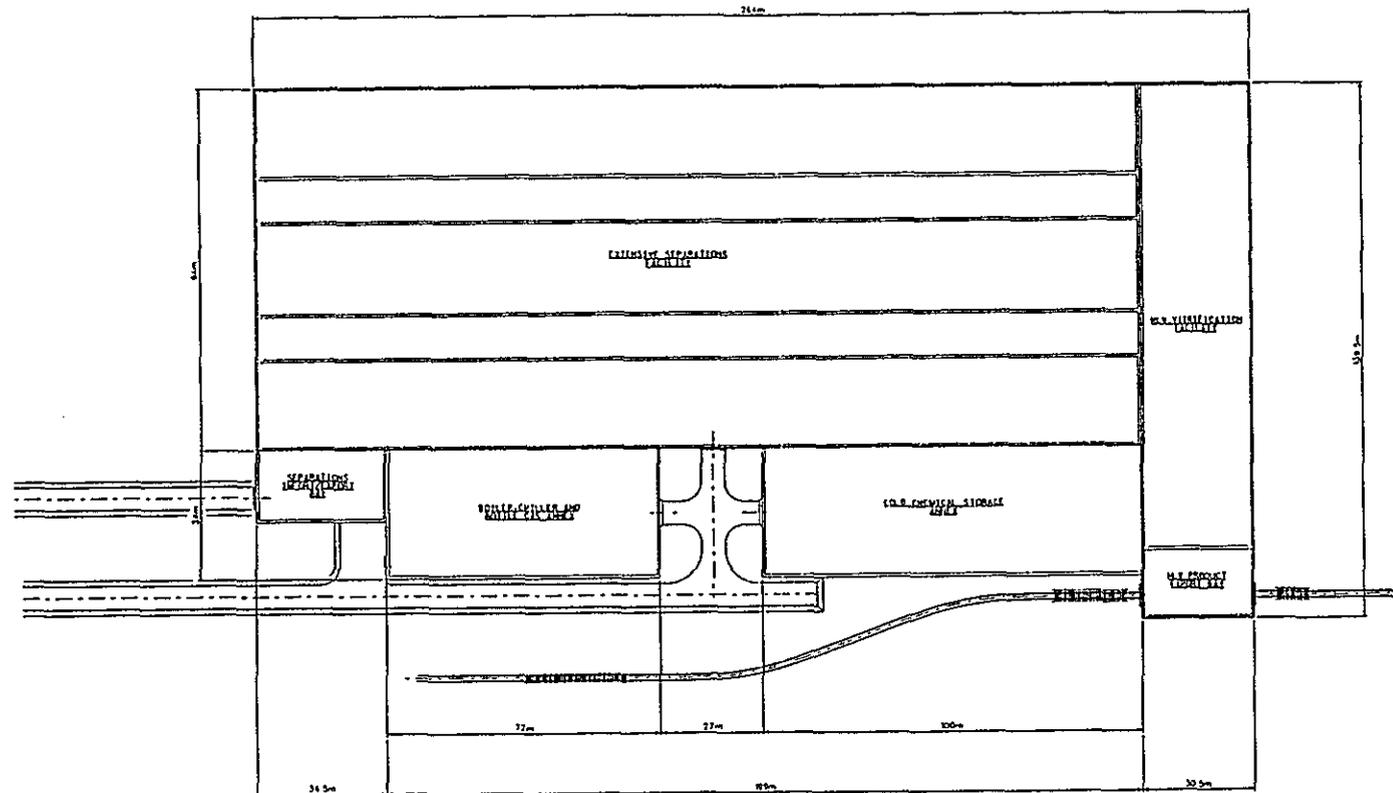
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Figure E-1. Extensive Separations Pretreatment and High-Level Waste Treatment Facility Layout (sheet 5 of 5).



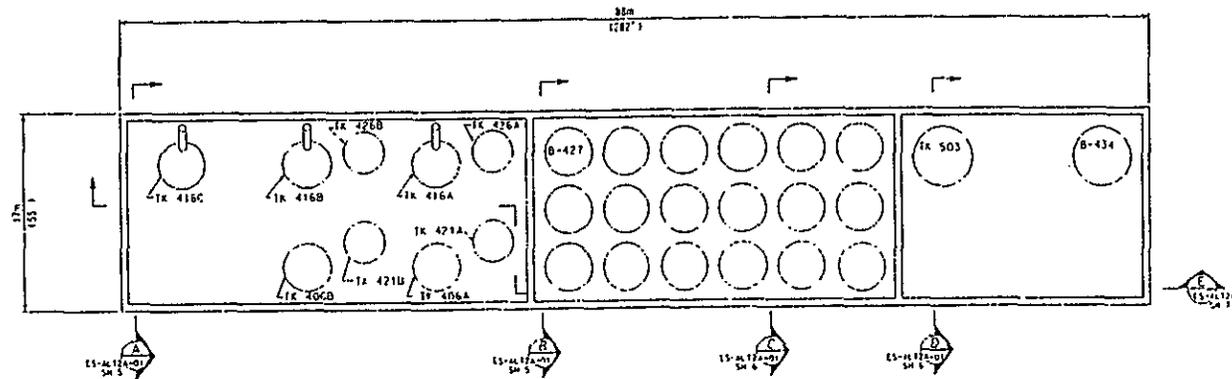
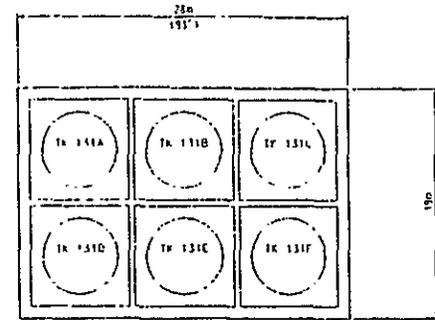
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Figure E-2. Low-Level Waste Vitrification Facility Layout (sheet 1 of 10).



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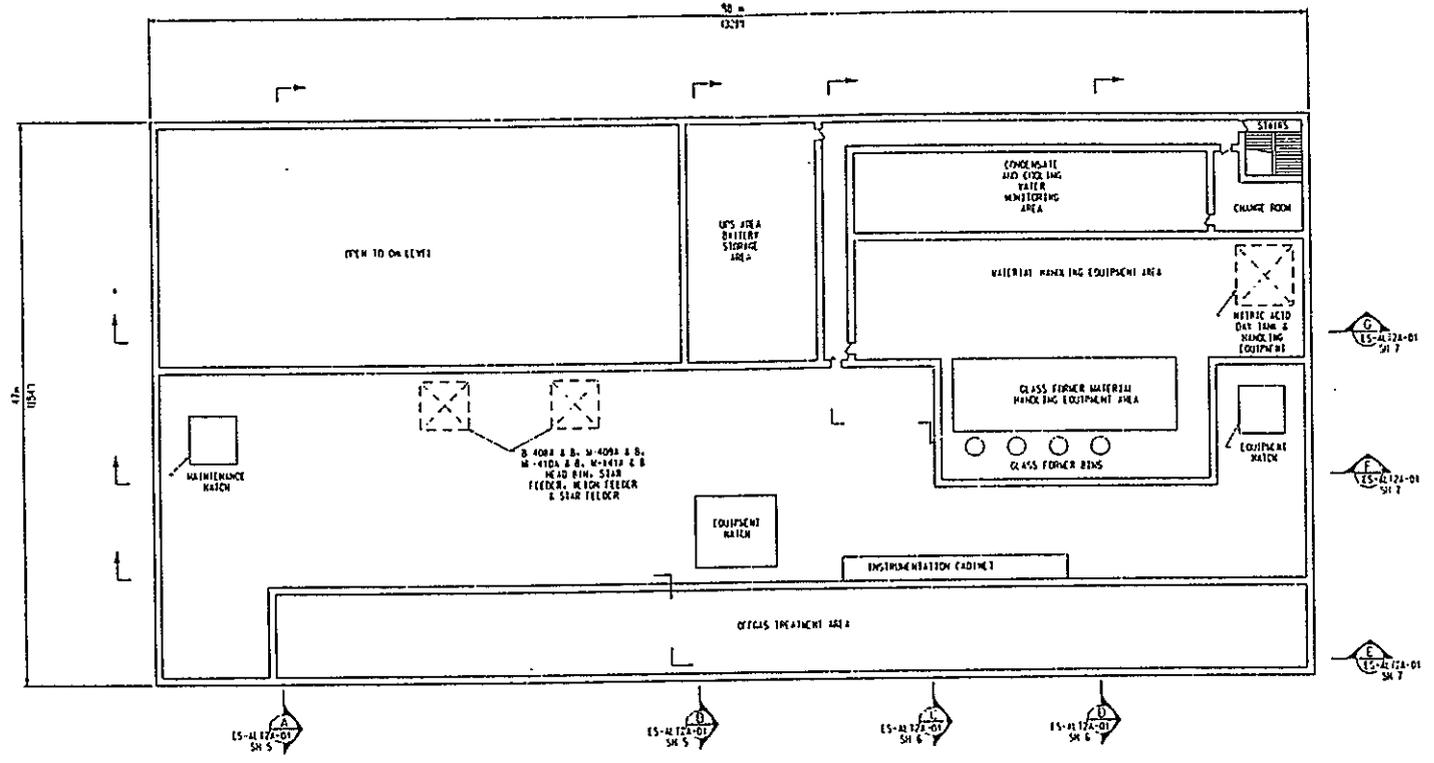
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Figure E-2. Low-Level Waste Vitrification Facility Layout (sheet 3 of 10).



PLAN VIEW @ 12.6m

DATE: 05/12/94



NUMBER	REFERENCE OR DRAWING	DATE	BY	CHKD	APP'D

NO.	DATE	DESCRIPTION

STATE OF \_\_\_\_\_

REGISTERED PE \_\_\_\_\_

U.S. DEPARTMENT OF ENERGY  
 Health, Safety & Environment Office  
 EM-100-010

PLAN AT E-19.13.6m  
 ATTACHED LLW TREATMENT  
 LOW LEVEL WASTE

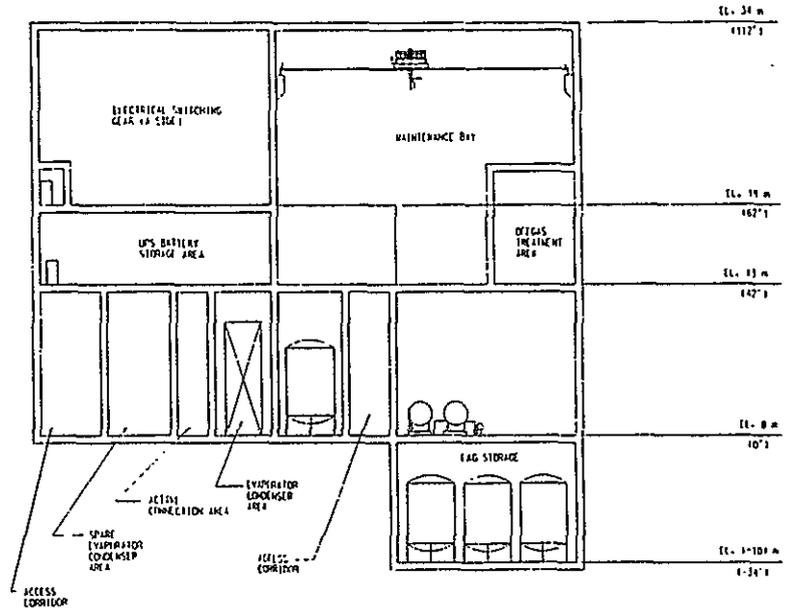
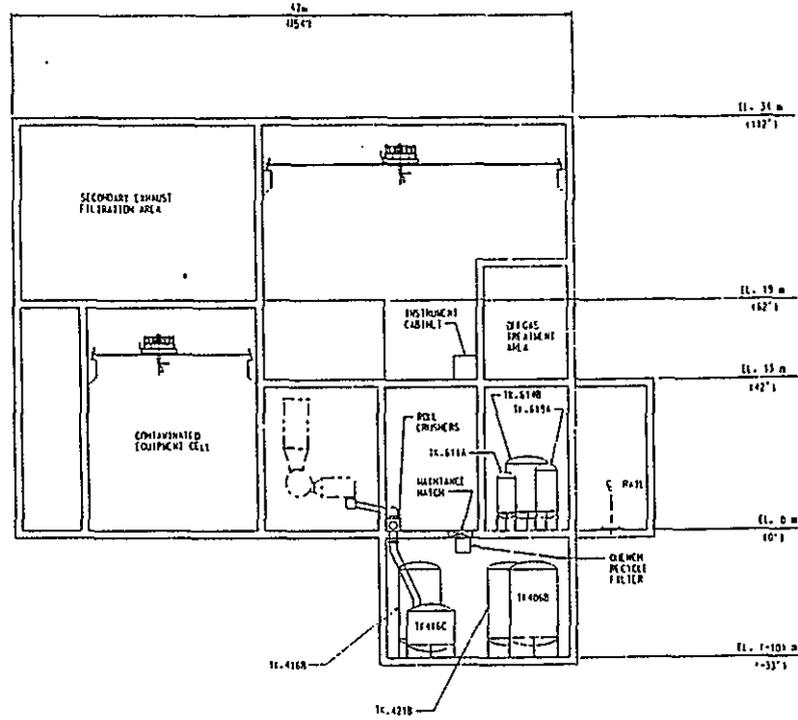
TYPE FACILITY OPTIONS ENGINEERING STUDY

**ES-AL12A-01A**

E-19/E-20 8/11/94



Figure E-2. Low-Level Waste Vitrification Facility Layout (sheet 5 of 10).

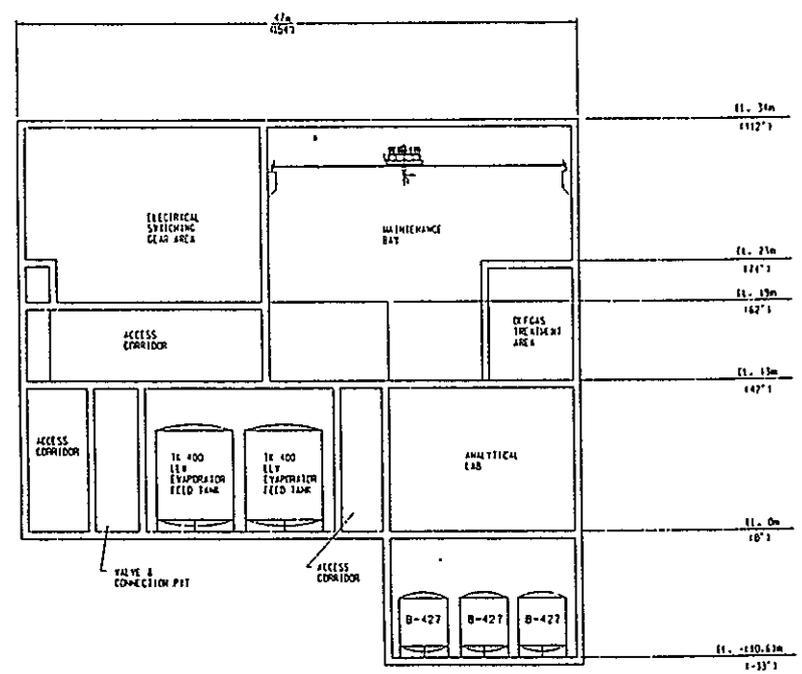


DATE: 05/12/94

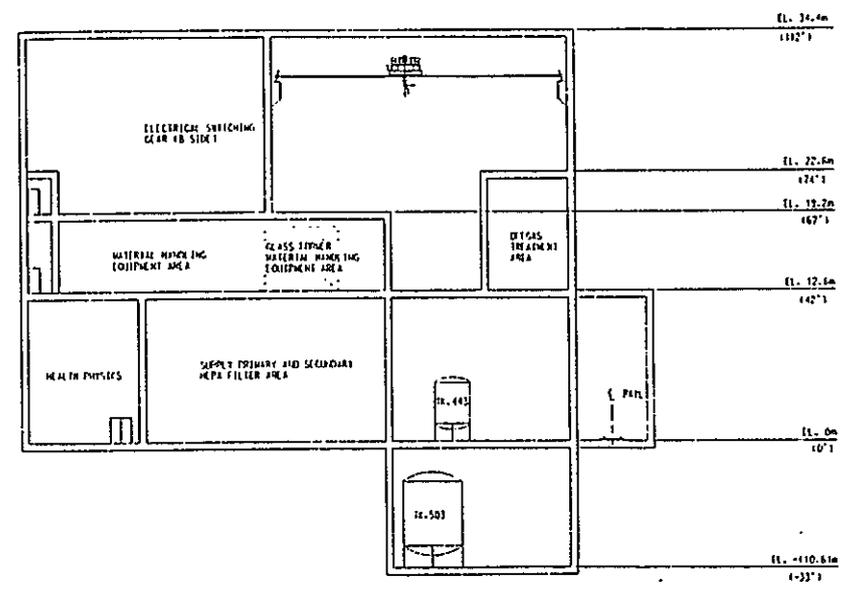
STATE OF	
REGISTERED PE	
U.S. DEPARTMENT OF ENERGY Brookhaven Operations Office EASCO/EMPT	
SECTION A & B DETACHED LLW TREATMENT LOW LEVEL WASTE	
THIS FACILITY DESIGN ENGINEERING STUDY	
ES-ALT2A-01.A	

NO.	DATE	DESCRIPTION
1		ISSUED FOR CONSTRUCTION
2		REVISED
3		REVISED
4		REVISED
5		REVISED
6		REVISED
7		REVISED
8		REVISED
9		REVISED
10		REVISED

Figure E-2. Low-Level Waste Vitrification Facility Layout (sheet 6 of 10).



C SECTION  
ES-AL22A-01 SH 1,2,3,4



D SECTION  
ES-AL22A-01 SH 1,2,3,4

DATE: 05/12/84

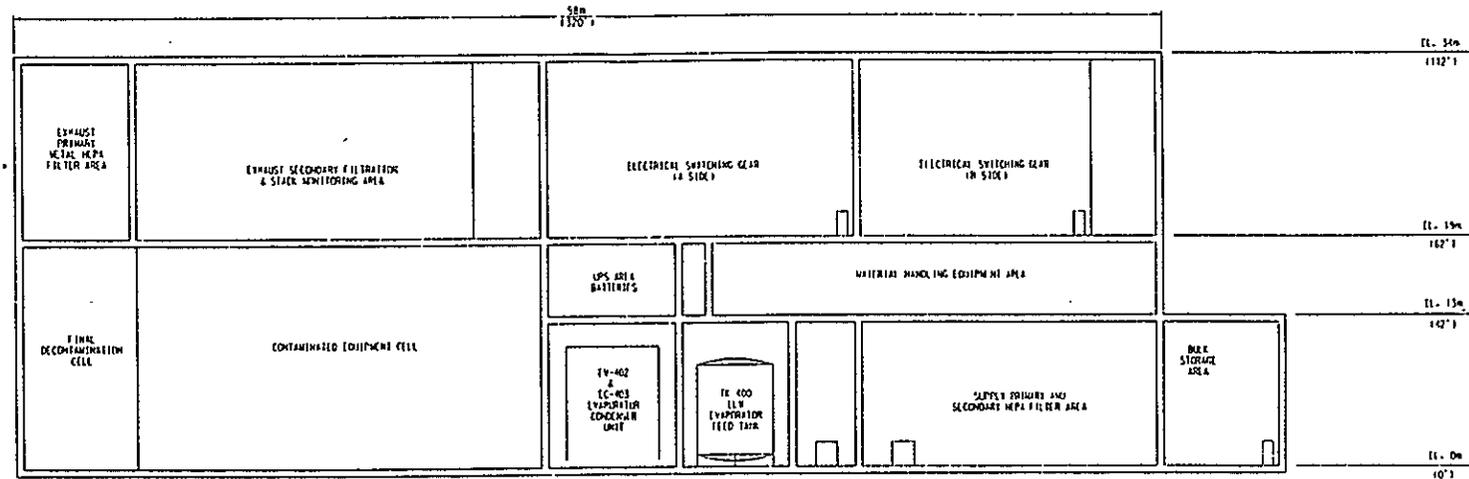


STATE OF	
REGISTERED PROFESSIONAL ENGINEER	
NAME	
FEDERAL BUREAU OF SURVEY	
SECTION C & D	
DETACHED LLL TREATMENT	
LOW LEVEL WASTE	
THIS FACILITY OPTIONS ENGINEERING STUDY	
ES-AL22A-01, A	

NO.	REVISION	DATE	BY	CHKD
1	ISSUED FOR CONSTRUCTION	05/12/84		

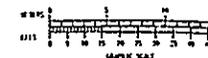


Figure E-2. Low-Level Waste Vitrification Facility Layout (sheet 8 of 10).



**G SECTION**  
 15-AL12A-01 SN 1+2,3+4

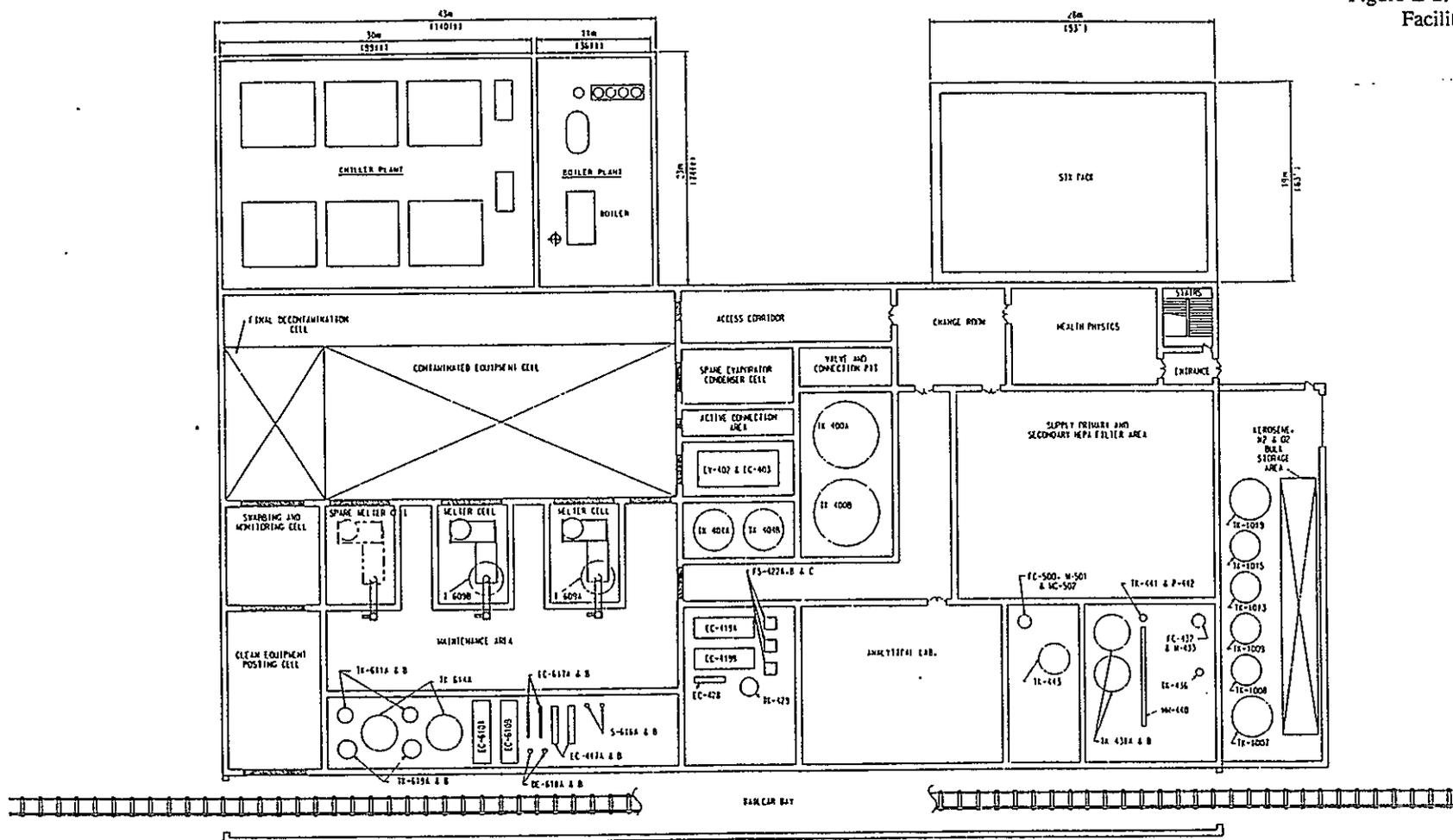
DATE: 05/12/94



NO.	REVISION	DATE	BY	CHKD BY
1	ISSUED FOR CONSTRUCTION	05/12/94		

U.S. DEPARTMENT OF ENERGY National Operations Office E-BASCO, OHIO <b>SECTION G</b> <b>WET-AGED LLW TREATMENT</b> <b>LOW SOURCE WASTE</b> FINAL FACILITY OPTING ENGINEERING STUDY		STATE OF  REGISTERED PE (Signature)
PROJECT NO. 15-AL12A-01 DRAWING NO. 01A SHEET NO. 8 OF 10		DATE: 05/12/94

Figure E-2. Low-Level Waste Vitrification Facility Layout (sheet 9 of 10).



PLAN VIEW @ EL. 0m

DATE: 05/12/94



REVISION	DATE	BY	DESCRIPTION

U.S. DEPARTMENT OF ENERGY National Operations Office EMERGENCY RESPONSE BENCH PLAN DETACHED LLW TREATMENT LOW SOURCE WASTE THIS FACILITY OPTIMUM ENGINEERING STUDY	STATE OF REGISTERED PE DATE: 05/12/94
PROJECT NO. ES-ALT2A-01A SHEET NO. B-31/E-32 OF 10	DATE: 05/12/94



EQUIPMENT LIST

Figure E-2. Low-Level Waste Vitrification Facility Layout (sheet 10 of 10).

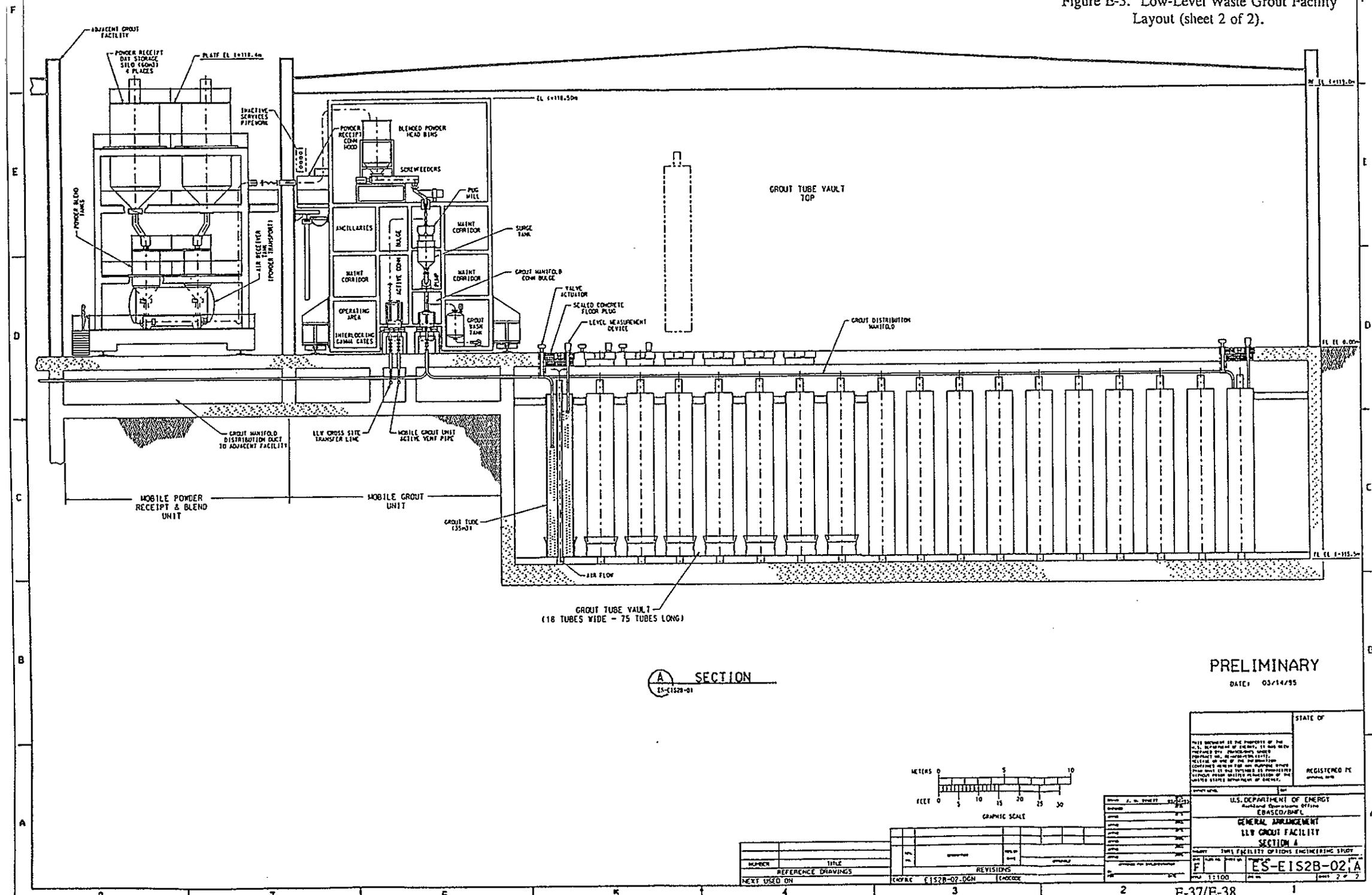
1K-131A	RECEIPT/SAMPLE TANKS
1K-131B	RECEIPT/SAMPLE TANKS
1K-131C	RECEIPT/SAMPLE TANKS
1K-131D	RECEIPT/SAMPLE TANKS
1K-131E	RECEIPT/SAMPLE TANKS
1K-131F	RECEIPT/SAMPLE TANKS
1J-400A	LLW EVAPORATOR FEED TANKS
1K-400B	LLW EVAPORATOR FEED TANKS
EV-402	LLW FEED EVAPORATOR
EC-405	CONDENSER
1K-404A	LLW MELTER FEED ADJUSTMENT TANK
1K-404B	LLW MELTER FEED ADJUSTMENT TANK
1K-406A	LLW MELTER FEED TANK
1K-406B	LLW MELTER FEED TANK
EM-412A	LLW MELTERS
EM-412B	LLW MELTERS
MC-415A	ROLL CRUSHERS
MC-415D	ROLL CRUSHERS
1K-416A	LLW CULLET CATCH TANK
1K-416B	LLW CULLET CATCH TANK
1K-416C	LLW CULLET CATCH TANK
EC-419A	CHILLER
EC-419B	CHILLER
1K-421A	LLW QUENCH WATER RECYCLE TANKS
1K-421B	LLW QUENCH WATER RECYCLE TANKS
FS-422A	SCREEN
FS-422B	SCREEN
FS-422C	SCREEN
1K-426A	LLW FILTER CATCH TANK
1K-426B	LLW FILTER CATCH TANK
B-427	LLW CULLET LAG STORAGE
EC-428	CONDENSER
1F-429	LLW CONDENSATE CATCH TANK
FC-432	CYCLONE w/ SINTERED METAL
M-433	ROTARY STAR FEEDER
B-434	OAT BIN
1K-436	RECOVERED SULFUR STORAGE TANK
1K-438A	SULFUR CEMENT MIXING TANK
1K-438B	SULFUR CEMENT MIXING TANK
MM-440	MIXER
1F-441	SURGE TANK
P-442	SULFUR TANK PUMP
1L-443	DECAHEDRON TREN SULFUR TANK
FC-500	CYCLONE w/ SINTERED METAL FILTERS
M-501	ROTARY STAR FEEDER
MC-502	ROLL CRUSHER
1K-503	RECYCLE CULLET CATCH TANK
1-609A	LLW QUENCH TOWER
1-609B	LLW QUENCH TOWER
RC-610A	CHILLER
RC-610B	CHILLER
1K-611A	LLW SCRUB FILTER TANK
1K-611B	LLW SCRUB FILTER TANK
1K-614A	LLW SCRUB SOLUTION TANK
1K-614B	LLW SCRUB SOLUTION TANK
S-616A	LLW SEPARATOR
S-616B	LLW SEPARATOR
EC-617A	CHILLER
EC-617B	CHILLER
DL-618A	LLW DEMISTER
DL-618B	LLW DEMISTER
1K-619A	SCRUB SOLUTION MAKEUP TANK
1K-619B	SCRUB SOLUTION MAKEUP TANK
1K-1007	BULK SULFUR LIQUID STORAGE TANK
1K-1008	BULK DCPD STORAGE TANK
1A-1009	BULK CPD OLIGOMER STORAGE TANK
1K-1013	SO <sub>2</sub> NOSH STORAGE TANK
1K-1015	BULK ANHYDROUS NH <sub>3</sub> STORAGE TANK
1K-1019	BULK WATER STORAGE TANK

DATE: 05/12/94

U.S. DEPARTMENT OF ENERGY National Operations Office EM-500/WVPL EQUIPMENT LIST RELATED LLW TREATMENT LOW LEVEL WASTE THIS FACILITY OPTION ENGINEERING STUDY		STATE OF  REGISTERED PE
DRAWN BY: [ ] CHECKED BY: [ ] DATE: [ ]	REVISIONS NO.   DESCRIPTION   DATE	PROJECT NO.: E-33/E-34 SHEET NO.: 10 of 10 ES-ALT2A-01A

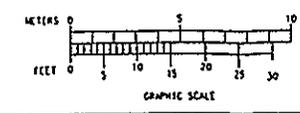


Figure E-3. Low-Level Waste Grout Facility Layout (sheet 2 of 2).



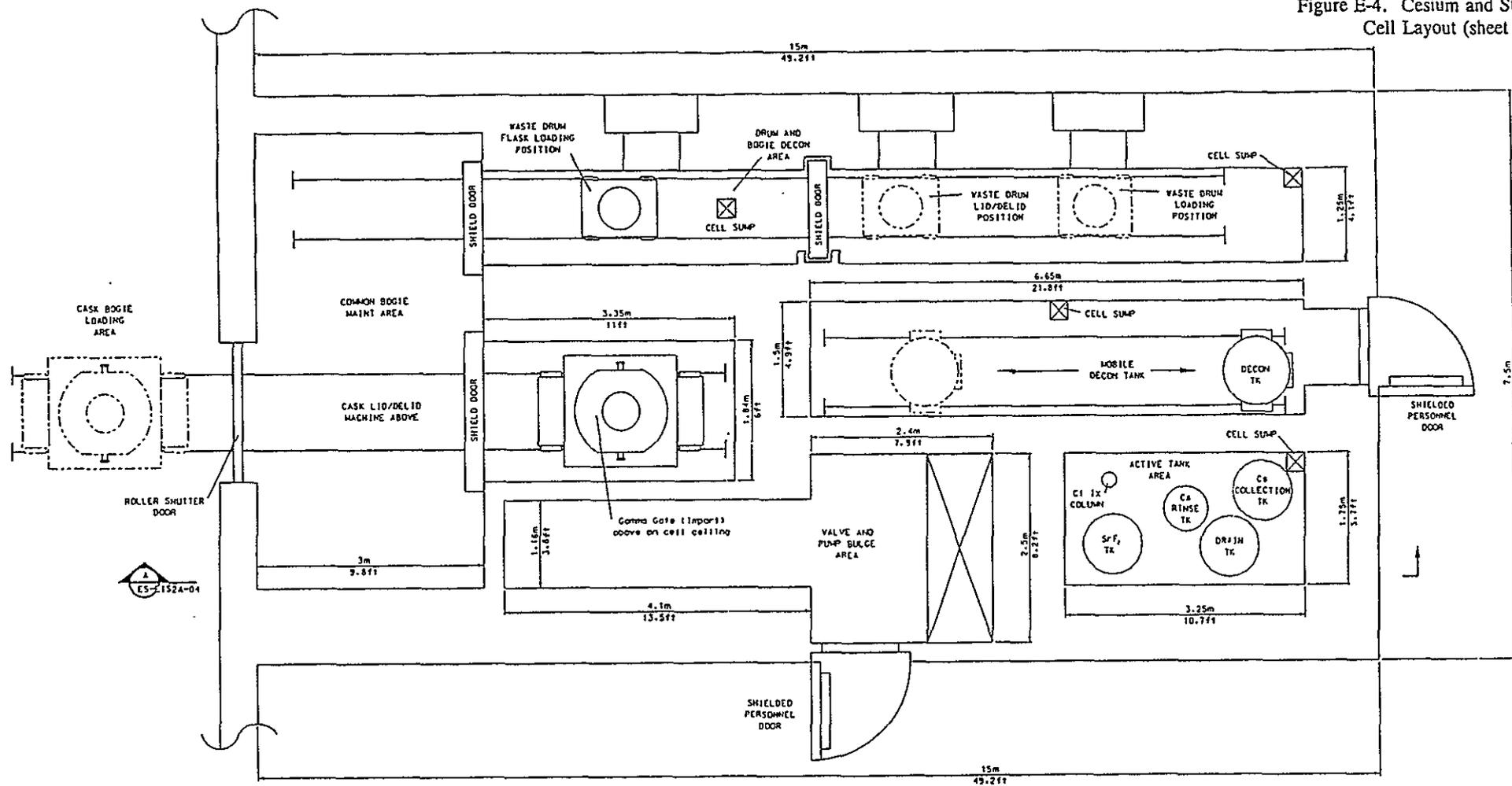
**A SECTION**  
E5-1128-01

**PRELIMINARY**  
DATE: 03/14/95



U.S. DEPARTMENT OF ENERGY Marketing Operations Office EASCO/BWF		STATE OF
GENERAL ARRANGEMENT LLW GROUT FACILITY SECTION A		REGISTERED PE
THIS FACILITY OPTIONS ENGINEERING STUDY		
NO. 15100	DATE 11/90	2 of 2

NUMBER	TITLE	DATE	BY	CHKD	APP'D
	REFERENCE DRAWINGS				
	REVISIONS				
	CHGRC E1528-02.DGN				
	CHOICE				



CELL PLAN AT EL. (+)0.00m

EQUIPMENT LIST		
EQPT NO	DESCRIPTION	QTY
GG-1	GAMMA GATE (IMPORT)	1
CC-2	BUSS CASK BOGIE	1
CC-14	...	1
GG-18	WASTE DRUM EXPORT SHIELD DOORS	2
GG-19	BUSS CASK IMPORT SHIELD DOOR	1
TK-20	SF <sub>6</sub> SLURRY TANK	1
TK-21	C <sub>1</sub> RINSE TANK	1
TK-22	C <sub>1</sub> COLLECTION TANK	1
TK-23	DRAIN TANK	1
TK-24	DECONTAMINATION TANK	1
TK-25	C <sub>1</sub> ION EXCHANGE COLUMN	1
GG-37	SHIELDED PERSONNEL DOORS	2

PRELIMINARY  
DATE: 03/14/85

E-39/40

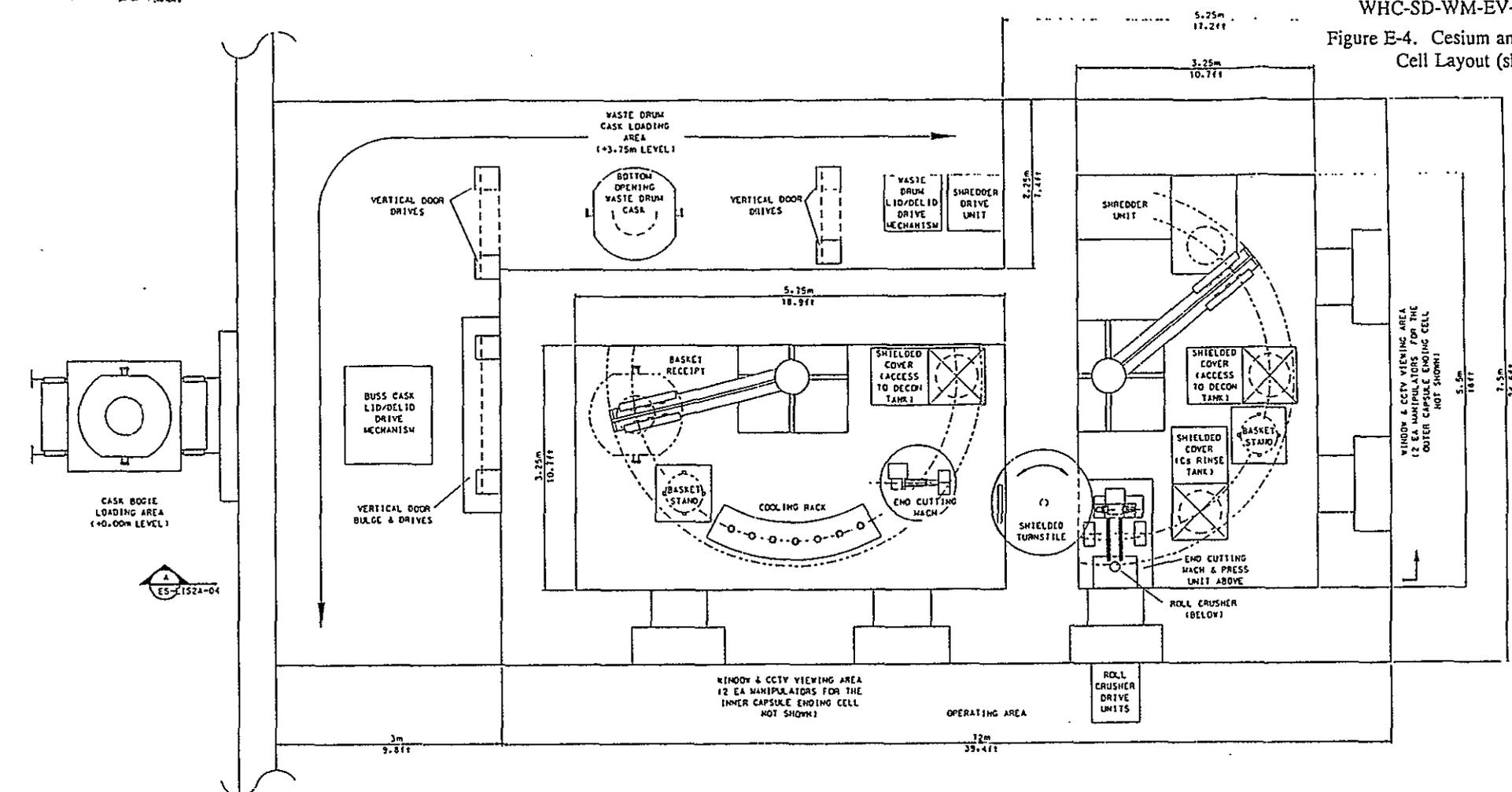
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<p>U.S. DEPARTMENT OF ENERGY Nuclear Operations Office EBASCO/BWPL</p>		REGISTERED PE
<p>GENERAL ARRANGEMENT C<sub>1</sub> - S<sub>1</sub> CAPSULE DISPOSAL CELL PLAN EL. (+)0.00m</p>		
<p>ES-E152A-01A</p>		

NO.	DATE	DESCRIPTION	BY	CHKD

NO.	DATE	DESCRIPTION	BY	CHKD



CELL PLAN AT EL. (+13.75m)

EQUIP NO	DESCRIPTION	QTY
ST-3	BASKET STAND	2
RK-4	COOLING RACK	1
FX-5	OUTER CAPSULE END CUTTING MACHINE	1
CG-6	SHIELDED TURNSTILE (CAPSULE TRANSFER)	1
FX-7	INNER CAPSULE END CUTTING MACHINE	1
FX-8	ROLL CRUSHER	1
BB-9	HULL AND END DECON BASKET	2
CG-10	DECON TANK SHIELDED TANK COVER	2
CG-11	CS RINSE TANK SHIELDED TANK COVER	1
CM-12	JIB HOIST (.75 T)	2
FX-13	CAPSULE SHREDDER	1
CM-15	BUSS CASK LID/DELID MACHINE	1
WM-16	SHIELDED WINDOWS	5
TV-17	CCTV CAMERA COPY	2
CG-36	VERTICAL SHIELD DOOR MAINT BULGE	1
CM-38	WASTE DRUM LID/DELID MACHINE	1
FX-39	SRF, PRESS	1
FL-40	WASTE DRUM CASK	2
CG-41	VERTICAL SHIELD DOOR MAINT BULGE	2
CM-42	MANIPULATORS	4

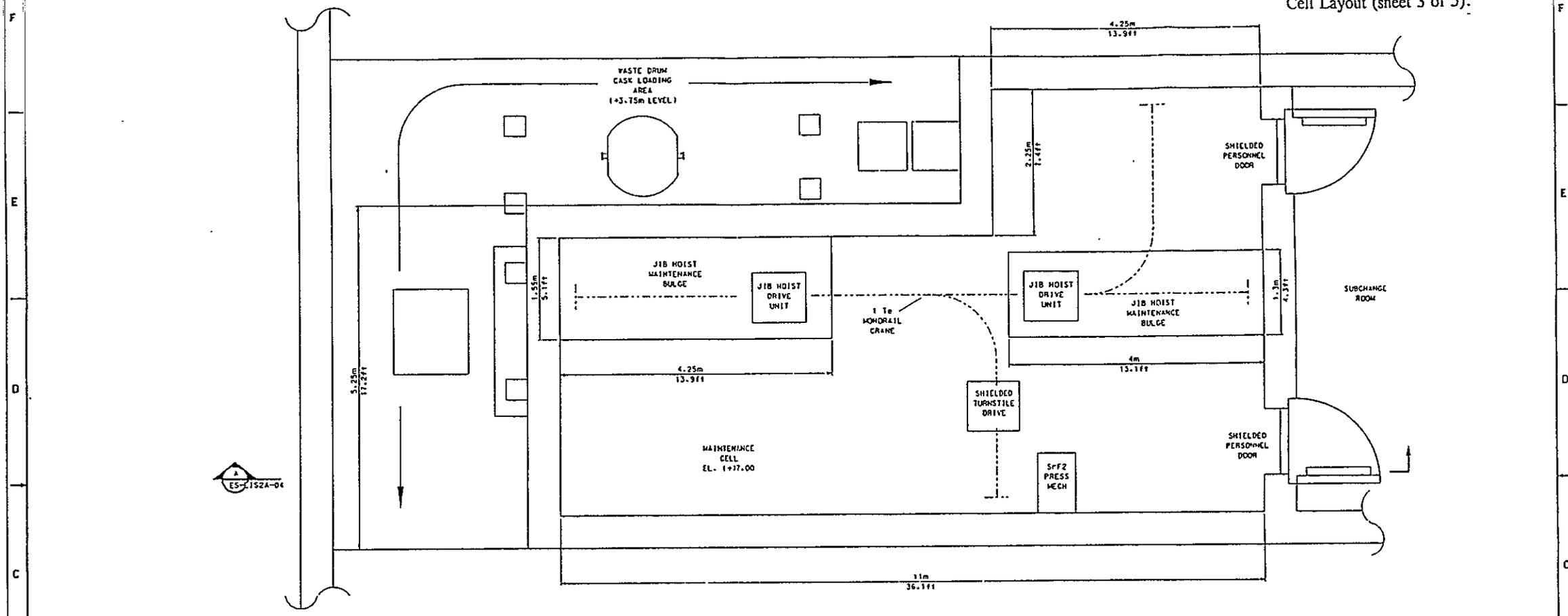
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 DATE: 03/14/95

STATE OF	REGISTERED PE
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U.S. DEPARTMENT OF ENERGY Research and Development Office EBS&CO/DMFL GENERAL ARRANGEMENTS Cs - Sr CAPSULE DISPOSAL CELL PLAN #1, (+13.75m) DATE: 03/14/95 DRAWN BY: [Signature] CHECKED BY: [Signature] APPROVED BY: [Signature]	
ES-E1S2A-02.A SHEET 2 OF 5	DATE: 03/14/95

NUMBER	REFERENCE DRAWINGS	TITLE	REVISIONS

F  
E  
D  
C  
B  
A



CELL PLAN AT EL. (+)7.50m

EQUIPMENT LIST		
EQPT NO	DESCRIPTION	QTY
GN-32	CRANE, MONORAIL (1 Tc)	1
GG-35	JIB HOIST MAINT BULGES	2
GG-37	SHIELDED PERSONNEL DOOR	2

PRELIMINARY  
DATE: 03/14/95

E-43/E-44

NO.	DATE	BY	DESCRIPTION

U.S. DEPARTMENT OF ENERGY National Operations Office EBASCO/BNL <b>GENERAL ARRANGEMENT</b> Cs - Sr CAPSULE DISPOSAL CELL PLAN EL. (+)7.50m		STATE OF  REGISTERED PE <small>NAME</small> <small>NO.</small>
PROJECT: TMS FACILITY OFFICE ENGINEERING STUDY SHEET NO: F DRAWING NO: ES-E1S2A-03.A DATE: 03/14/95		U.S. DEPARTMENT OF ENERGY NATIONAL OPERATIONS OFFICE EBASCO/BNL

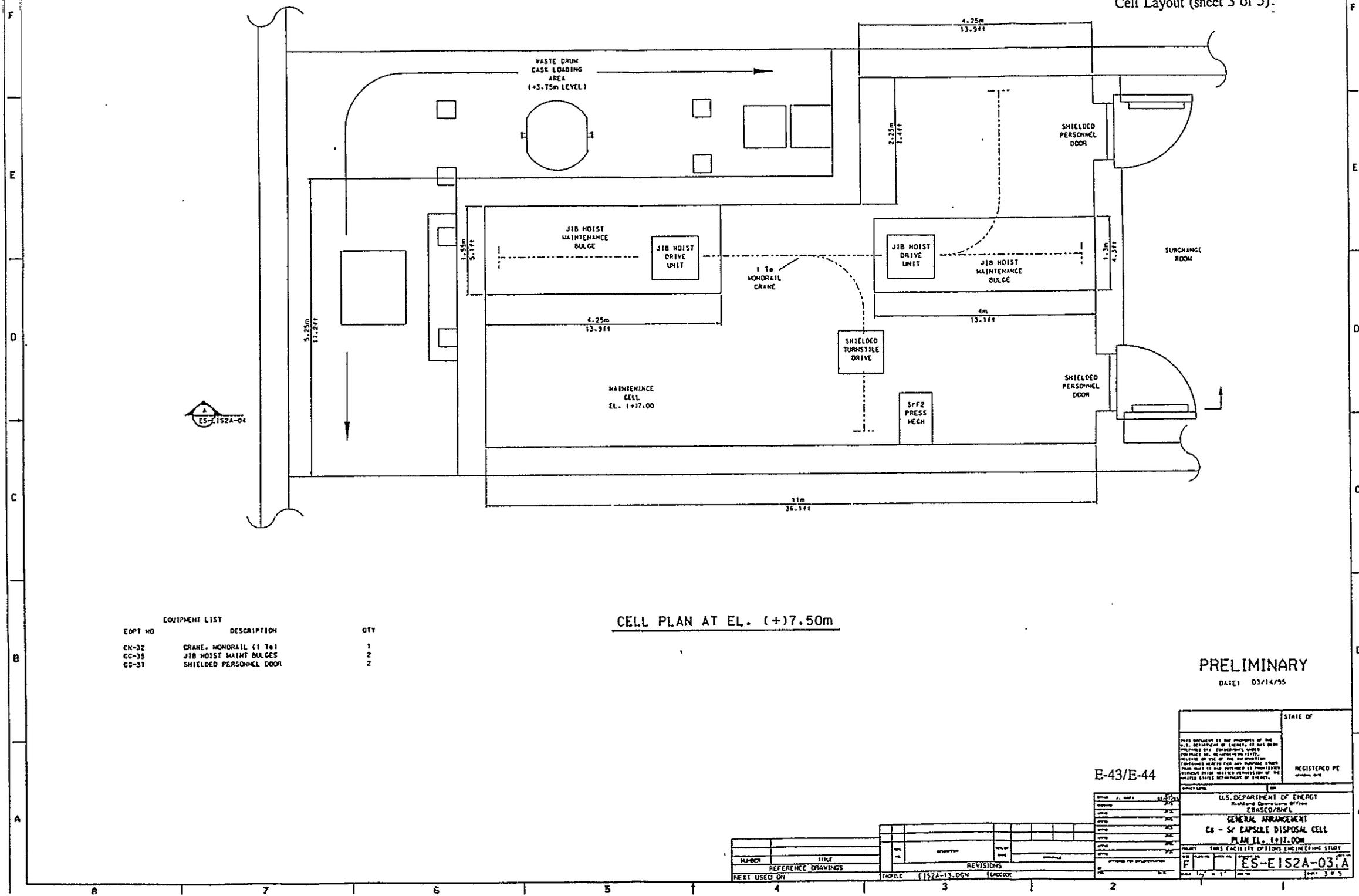
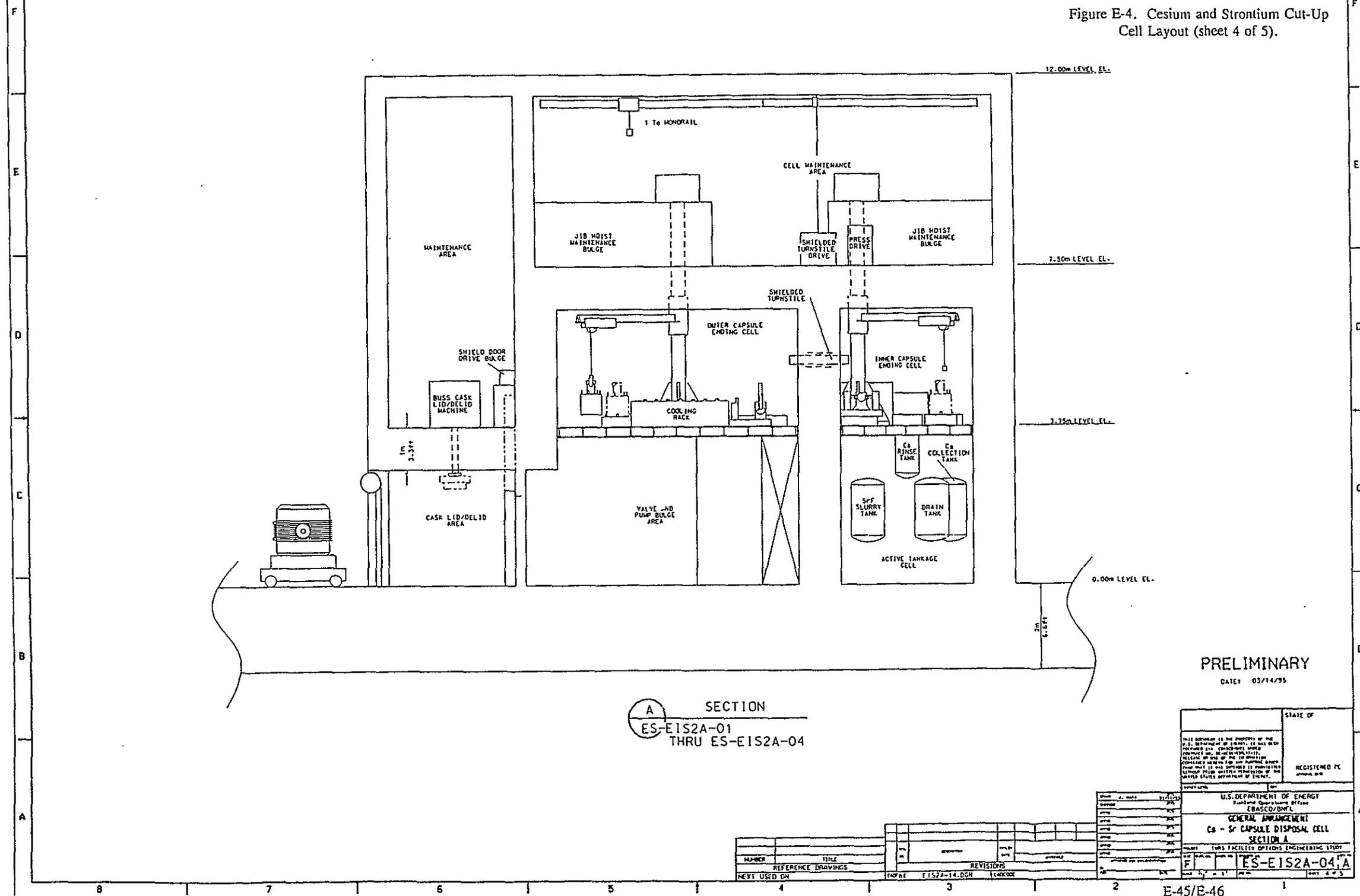


Figure E-4. Cesium and Strontium Cut-Up Cell Layout (sheet 4 of 5).



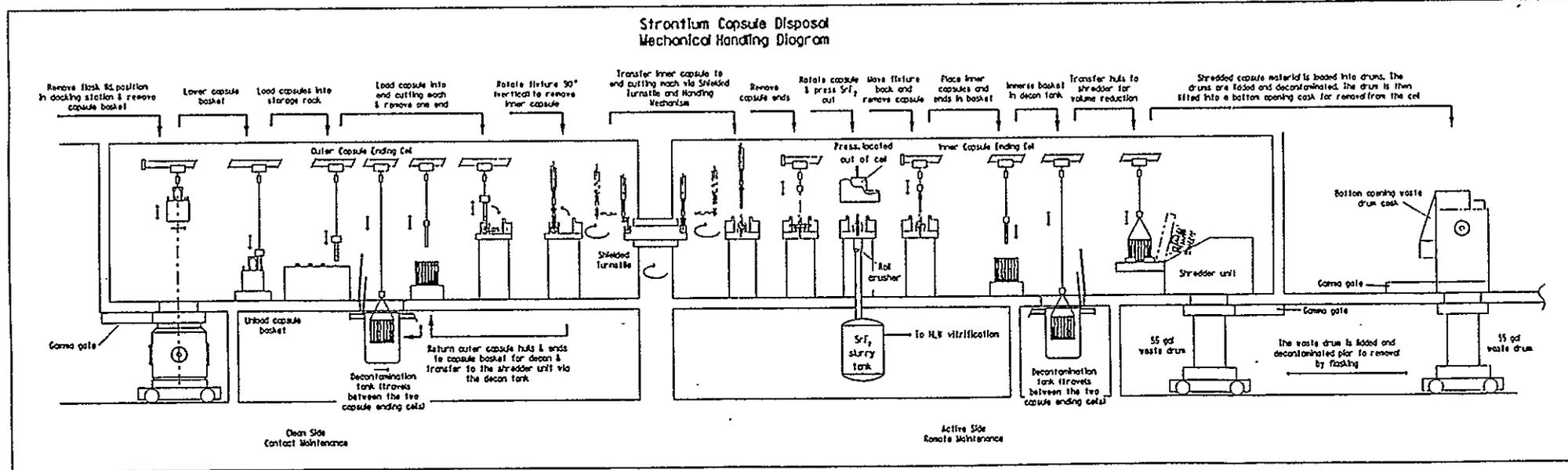
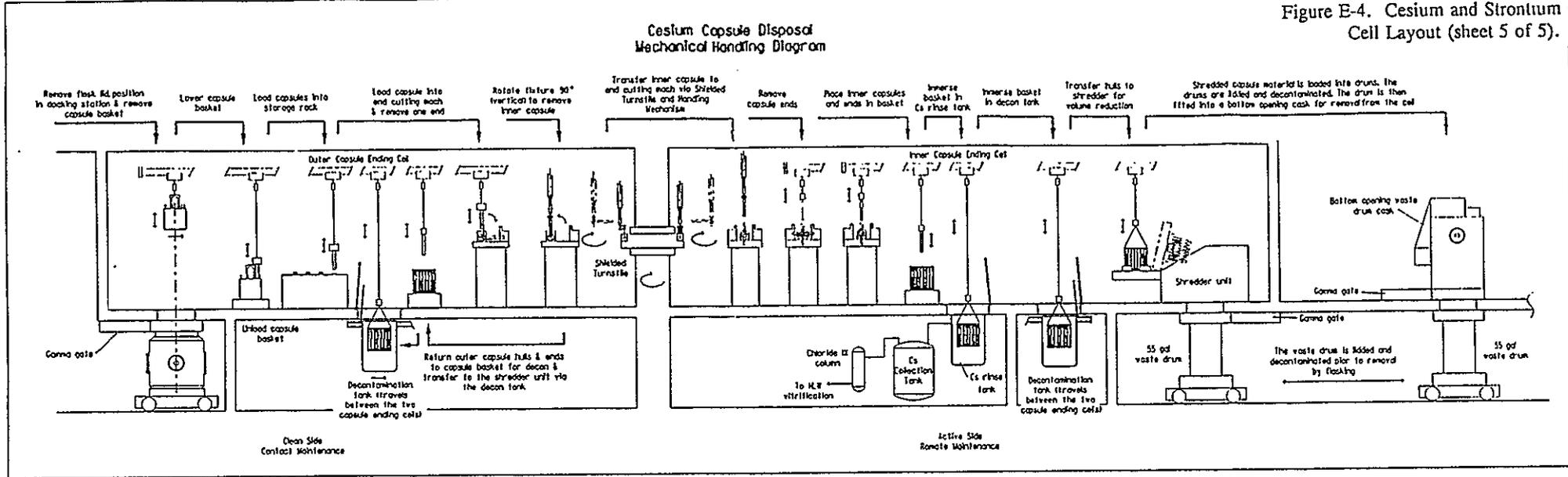
A SECTION  
 ES-E1S2A-01  
 THRU ES-E1S2A-04

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U.S. DEPARTMENT OF ENERGY Washington, D.C. 20545-0001 <b>GENERAL ARRANGEMENT</b> <b>Cs - Sr CAPSULE DISPOSAL CELL</b> <b>SECTION A</b>	
THIS FACILITY OPTIONS ENGINEERING STUDY <b>ES-E1S2A-04, A</b>	

NO.	DATE	BY	DESCRIPTION

Figure E-4. Cesium and Strontium Cut-Up Cell Layout (sheet 5 of 5).



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U.S. DEPARTMENT OF ENERGY Environmental Restoration Office EBASCO/WHC	
MECHANICAL HANDLING DIAGRAM Cs - Sr CAPSULE DISPOSAL CELL LAYOUT STUDY	
PROJECT: TMS FACILITY OPTIONS ENGINEERING STUDY	
DATE: 03/14/95	ES-E1S2A-051A
REVISED	REVISED
BY: _____	BY: _____
DATE: _____	DATE: _____
APPROVED FOR SUBMITTAL	DATE: _____
BY: _____	DATE: _____

**APPENDIX F**

**BACKUP TABLES**

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**APPENDIX F****BACKUP TABLES**

The tables in this appendix provide backup information for the data tables in Section 9.0.

Table F-1. Backup to Table 9-2. Unit Process: Operating Personnel Requirements (Staff Hours).

Staffing Category	Years	Staff Per Year	Total Staff Years	Total Staff Hours
Extensive separations	15	620	9,300	1.7E+07
LLW vitrification	15	160	2,400	4.3E+06
HLW vitrification	15	160	2,400	4.3E+06
Indirect staffing	19	29	551	1.0E+06
Pretreatment start-up	1.5	626	939	1.7E+06
LLW vitrification start-up	1.5	157	236	4.3E+05
HLW start-up	1.5	157	236	4.3E+05
Pretreatment decontamination and decommissioning	2	626	1,252	2.3E+06
LLW vitrification decontamination and decommissioning	2	157	314	5.7E+05
HLW vitrification decontamination and decommissioning	2	157	314	5.7E+05
HLW monitoring and maintenance	12	10	120	2.2E+05
HLW transportation	—	30	30	5.4E+04
			Total	3.3E+07

Note:

HLW = high-level waste

LLW = low-level waste

All staff hours are based on a staff-year of 1,812 hours.

Table F-1A. Backup to Table 9-2. Unit Process: Operating Personnel Requirements (Staff Hours).

Staffing Category	Cesium Removal	LLW Vitrification	LLW Disposal	HLW Vitrification	HLW Transportation/ Disposal	Total
Extensive separations	16,851,600					1.7E+07
LLW vitrification		4,348,800				4.3E+06
HLW vitrification				4,348,800		4.3E+06
Indirect staffing	249,603	249,603	249,603	249,603		1.0E+06
Pretreatment start-up	1,701,468					1.7E+06
LLW vitrification start-up		426,726				4.3E+05
HLW start-up				426,726		4.3E+05
Pretreatment decontamination and decommissioning	2,268,624					2.3E+06
LLW vitrification decontamination and decommissioning		568,968				5.7E+05
HLW decontamination and decommissioning				568,968		5.7E+05
HLW monitoring and maintenance				217,440		2.2E+05
HLW transportation					54,360	5.4E+04
<b>Total</b>	<b>2.1E+07</b>	<b>5.6E+06</b>	<b>2.5E+05</b>	<b>5.8E+06</b>	<b>5.4E+04</b>	<b>3.3E+07</b>

Notes:

HLW = high-level waste  
 LLW = low-level waste

Table F-2. Backup to Table 9-3. Operating Personnel Requirements (Staff Hours/Years).

Operation Mode	Exempt	Nonexempt	Bargaining Unit	Total	Exempt	Nonexempt	Bargaining Unit
Pretreatment	202	40	378	620	33%	6%	61%
LLW vitrification	52	10	98	160	33%	6%	61%
HLW vitrification	52	10	98	160	33%	6%	61%
Total	306	60	574	940	98%	19%	183%

Notes:

HLW = high-level waste

LLW = low-level waste

Total staffing of 940 was based on an estimate by K. D. Boomer.

Table F-2A. Backup to Table 9-3. Operating Personnel Requirements  
(Staff-Hours/Year).

Staffing Category	Radiation Worker <sup>1</sup>	Nonradiation Worker <sup>1</sup>	Total
Extensive separations	1.6E+07	1.1E+06	1.7E+07
Low-level waste vitrification	4.1E+06	2.7E+05	4.3E+06
High-level waste vitrification	4.1E+06	2.7E+05	4.3E+06
Indirect staffing <sup>2</sup>		1.0E+06	1.0E+06
Pretreatment start-up	1.6E+06	1.1E+05	1.7E+06
Low-level waste vitrification start-up	4.0E+05	1.4E+05	5.4E+05
High-level waste start-up	4.0E+05	2.7E+04	4.3E+05
Pretreatment decontamination and decommissioning	2.1E+06	1.5E+05	2.3E+06
Low-level waste vitrification decontamination and decommissioning	5.3E+05	3.6E+04	5.7E+05
High-level waste decontamination and decommissioning	5.3E+05	3.6E+04	5.7E+05
High-level waste monitoring and maintenance <sup>3</sup>	4.3E+04	1.7E+05	2.2E+05
High-level waste transportation <sup>4</sup>	3.9E+04	1.5E+04	5.4E+04
Total	3.0E+07	3.3E+06	3.3E+07

## Notes:

<sup>1</sup>All exempt and bargaining unit employees are assumed to be radiation workers. All nonexempt employees are nonradiation workers.

<sup>2</sup>For indirect staffing, it was assumed that all workers would be nonexempt employees.

<sup>3</sup>For monitoring and maintenance, it was assumed that exempt employees would be 10 percent, bargaining unit employees would be 10 percent, and nonexempt employees would be 80 percent.

<sup>4</sup>For high-level waste transportation, it was assumed that exempt employees would be 36 percent, bargaining unit employees would be 36 percent, and nonexempt employees would be 28 percent.

Table F-3A. Backup to Table 9-7. Transportation in Support of Processing<sup>1</sup>.

Material	Tri-Party Agreement (Metric tons)	Clean option (Metric tons)
Glass former	290,000	354,814
Bulk frit	52	949,937
Nitric acid	4,200	
Sodium hydroxide	26,000	
Sodium nitrite	54	
Flocculent	85	
Glycolic acid	3,500	
Decontamination chemicals	5,300	
Ammonia	8,800	
Kerosene	57,800	
Sulfur	130,000	
Dicyclopentadiene (DCPD)	3,400	
Oligomer	3,400	
Total	532,591	1,304,751

## Notes:

<sup>1</sup>Information is taken from Table 9-4 in the *Tri-Party Agreement Alternative Engineering Data Package for the Tank Waste Remediation System Environmental Impact Statement* (Slaathaug 1995).

Slaathaug, E. J., 1995, *Tri-Party Agreement Alternative Engineering Data Package for the Tank Waste Remediation System Environmental Impact Statement*, WHC-SD-WM-EV-104, Rev. 0, Westinghouse Hanford Company, Richland, Washington.

Table F-3B. Backup to Table 9-7. Transportation in Support of Processing<sup>1,2</sup>.

Transportation	Trips	Load
Train	637	2,048 metric tons per train <sup>3</sup> 102 metric tons per car <sup>3</sup>
Truck	513	4,910 cubic meters <sup>4</sup>

## Notes:

<sup>1</sup>Assumes ion exchange media will be delivered by truck and all other materials will be delivered by train.

<sup>2</sup>Information is taken from Table 9-2 and 9-7 in the *Tri-Party Agreement Alternative Engineering Data Package for the Tank Waste Remediation System Environmental Impact Statement*, (Slaathaug 1995).

<sup>3</sup>Metric tons of cold chemicals.

<sup>4</sup>Ion exchange media.

Slaathaug, E. J., 1995, *Tri-Party Agreement Alternative Engineering Data Package for the Tank Waste Remediation System Environmental Impact Statement*, WHC-SD-WM-EV-104, Rev. 0, Westinghouse Hanford Company, Richland, Washington.

Table F-4. Backup to Tables 9-8 and 9-9. Construction Personnel Requirements (Staff-Hours).

Staff	Extensive Pretreatment Facility	Centralized Facilities	LLW Vitrification Facility <sup>1</sup>	HLW Vitrification Facility <sup>2</sup>	Total All Facilities
Design and engineering	7,924,000	1,822,000	2,135,000	1,916,000	13,797,000
Construction					
Radiation worker	0	0	0	0	0
Nonradiation construction worker	17,012,000	4,559,000	3,984,000	3,737,000	29,292,000
Supervision	4,953,000	1,118,000	1,335,000	1,197,000	8,603,000
Total	29,889,000	7,499,000	7,454,000	6,850,000	51,692,000

## Notes:

HLW = high-level waste

LLW = low-level waste

<sup>1</sup>Low-level waste vitrification (low source), Option 2A.<sup>2</sup>High-level waste vitrification facility is combined with extensive pretreatment.

Table F-6A. Backup to Table 9-10. Construction Resource Requirements.

Account	Central Facility	Canister Storage Building	Extensive Pretreatment	HLW Vitrification	LLW Vault	LLW Vitrification (Option 2A)	Total
Structural backfill (m <sup>3</sup> )	250,000	44,200	274,200	74,600	153,500	53,600	850,100
Excavation (m <sup>3</sup> c)	350,000	71,300	770,600	81,600	436,000	83,000	1,792,800
Clear and grubbing (m <sup>2</sup> )	1,498,100		54,660	10,400			1,563,160
Concrete (m <sup>3</sup> )	27,750	19,300	186,200	33,520	138,400	37,400	442,570
Carbon steel (metric ton)	15,713	7,011	67,632	12,855	16,200	12,391	131,802
Reinforcing bar	1,685	3,433	36,671	6,939		7,570	56,298
Piping	1,890	22	1,839	112		465	4,328
Equipment	7,526	694	2,399	359		465	11,443
Structural steel	3,864	2,528	23,502	4,833	16,200	3,301	54,228
Misc @ 5%	748	334	3,221	612		590	5,505
Stainless steel (metric ton)	1,776	0	10,508	1,530	0	3,157	16,971
Piping	473		2,470	243		835	4,021
Equipment	1,218		5,105	550		1,452	8,325
Wall boxes			821	349		440	1,610
Liner plate			1,612	315		280	2,207
Misc. at 5%	85		500	73		150	808

Table F-6A. Backup to Table 9-10. Construction Resource Requirements.

Hastelloy	0	0	1,344	396	0	97	1,837
Piping			50	19		53	122
Equipment			1,214	336		13	1,563
Wall boxes			16	22		26	64
Misc. @ 5%			64	19		5	88

Notes:

HLW = high-level waste

LLW = low level waste

m<sup>2</sup> = square meters

m<sup>3</sup> = cubic meters

Table F-6B. Backup to Table 9-10. Construction Resource Requirements.

<u>Surface Areas</u>			
Surface committed temporarily (hectares/acres)	125		309
Surface committed permanently (hectares/acres)	12		30
Clearing and Grubbing Area (acres)			309
<u>Excavation</u>			
Total excavation (cubic yards [yd <sup>3</sup> ])			2,344,894
Total excavation (cubic meters [m <sup>3</sup> ])			1,792,800
<u>Excavation for buildings</u>			
0.00 to - 35.0 0 (yd <sup>3</sup> )			1,887,732
haul to waste pile			1,233,005
haul to stockpile			654,727
Elevation -35 to -50 (yd <sup>3</sup> ) (hailed to stockpile)			457,162
Structural backfill (must match cubic yards to stockpile)			1,111,889
<u>Concrete Total (m<sup>3</sup>)</u>			442,570
<u>Steel Total (metric tons)</u>			150,61
<u>Summaries for total contaminated material</u>			
Total concrete quantity (m <sup>3</sup> ) (includes concrete for facilities that will not be contaminated)			442,570
Concrete associated with facilities that are not expected to be contaminated (m <sup>3</sup> )			429,714
Total concrete expected to be contaminated (m <sup>3</sup> )			12,856
<u>Total Steel Quantities (metric tons)</u>			
Carbon Steel	Total: 131,802	Contaminated: 4,644 tons.	Uncontaminated: 127,158
Stainless	Total: 16,971	Contaminated: 12,916 tons.	Uncontaminated: 4,055
Hastl/Incnl	Total: <u>1,837</u>	Contaminated: 1,837 tons.	Uncontaminated: 0
	Total: 150,610		

Table F-6C. Backup to Table 9-10. Construction Resource Requirements.

Construction Resource	Extensive Pretreatment
<b>Land (hectares)</b> Surface committed Temporarily Permanently	  125 12
<b>Water (cubic meters [m<sup>3</sup>])</b>	48
<b>Energy</b> Electrical (gigawatt hours) Propane (m <sup>3</sup> )  Diesel fuel (liters) Gasoline (liters)	 82.5 0  2.6E+07 9.6E+06
<b>Materials</b>  Concrete (m <sup>3</sup> )  Steel (metric tons) Carbon Steel Stainless Steel Hastelloy/Incnl  Excavation (m <sup>3</sup> ) Riprap (m <sup>3</sup> )  Structural backfill (m <sup>3</sup> )  Total contaminated material (m <sup>3</sup> )	  4.43E+05  1.32E+05 1.70E+04 1.84E+03  1.79E+06 0  8.50E+05  6.00E+03

Table F-7. Backup for Table 9-11. Nonradiological Construction Emission.  
Summary of PM10 Fugitive Emissions in Metric Tons.

Source of Emission	Metric Tons
Clear and Grub <sup>1</sup>	
Bulldozer grading <sup>2</sup>	7.82
Paved road traffic <sup>3</sup>	11.79
Construction Site Excavation <sup>4</sup>	
Unpaved road traffic <sup>5</sup>	348
Scraper unloading <sup>6</sup>	5
Dozer grading <sup>7</sup>	13.74
Wind erosion (storage and waste files) <sup>8</sup>	113.08
Aggregate Borrow Pit Excavation <sup>9</sup>	
Dump truck loading, truck unloading, crusher loading <sup>10</sup>	3.51
Unpaved road traffic <sup>11</sup>	4.1
Batch Plant Operation	
Dump truck unloading <sup>12</sup>	see above
Batch plant operation <sup>13</sup>	22
Wind erosion (aggregate and sand pile) <sup>14</sup>	0.28
Paved road traffic <sup>15</sup>	213.8
Unpaved road traffic <sup>16</sup>	1.84
Low-level waste vaults	284.12
<b>Total</b>	<b>1,029.08</b>

## Notes:

ER = emission rate	EF = emission factor
ft = foot/feet	ft <sup>2</sup> = square feet
ft <sup>3</sup> = cubic feet	g = grams
h = hour	in. = inch
kg = kilogram	km = kilometer
lb = pound	m = meter
m <sup>3</sup> = cubic meters	PM = particulate matter
sec = seconds	yd = yard
yd <sup>3</sup> = cubic yard	yr = year

<sup>1</sup>Emission sources would include dust from the clearing and grubbing by bulldozer and travel by dump truck on a paved road to dump collected vegetation.

<sup>2</sup>Assuming this operation is the same as soil grading, the ER, which factors in the silt and moisture content is estimated at 1.2g/sec. The time estimate for clearing and grubbing is 1.2g/sec x 1,810 x 3,600 sec/h = 7.82E+06g (7.82 metric tons).

Table F-7. Backup for Table 9-11. Nonradiological Construction Emission.  
Summary of PM10 Fugitive Emissions in Metric Tons.

Notes: (continued)

<sup>3</sup>The particulate emission rate, or PM 10 factor caused by traffic on paved roads is estimated at EF = 236g/km = EF. The total emission because of paved roads is as follows: 235g/km x 5 miles/trip x 6,237 x 1.609 kms/mile = 1.179E+07g (11.79 metric tons).

<sup>4</sup>Excavation would be done with push loaded scrapers. Once filled, the scrapers are assumed to travel 2,000 ft on a gravel road before dumping. A dozer with a vibratory roller would grade and compact the dumped soil. Scrapers also would be used for backfill operations. Emission sources would include the following: vehicular travel on unpaved roads, scraper unloading, dozer grading, and wind erosion of storage pits.

<sup>5</sup>The EF for travel on unpaved roads is calculated to be EF = 657g/km. It is assumed that a scraper would travel an average of 2,000 ft from the excavation pit to storage and waste piles and that it would take 100 ft for the scraper to fill its bowl and another 100 ft for the scraper to get to the road leading to the storage pile. Thus a one-way trip would be approximately 2,200 ft (0.6706 km). To complete the excavation, the number of one-way trips required would be 268,000. For the backfill operation, 127,100 one-way trips are required. The PM10 emission is as follows: 657g/km x 0.6706 km/trip x 2 x (268,000 + 127,100) trips = 3.48E+08g (348 metric tons).

<sup>6</sup>This is a transfer operation where particulate emissions would be affected by the particle size, mean wind speed, moisture content, the quantity of soil handled, etc.

The amount of soil unloaded by scraper would be equal to the soil excavated (2,344,894 yd<sup>3</sup>) and the soil used for backfill operation (1,111,889 yd<sup>3</sup>), that is 3,456,783 yd<sup>3</sup>.

The conversion from yd<sup>3</sup> of soil to kilograms is based on a soil density of 108 lbs per cubic foot (ft<sup>3</sup>) and is as follows: 3,456,783 yd<sup>3</sup> x 108 lbs per ft<sup>3</sup> x 27 ft<sup>3</sup> per yd x 0.4536 kilograms (kg) per lb = 4.572E+09 kg.

The PM emissions from scraper unloading are estimated as follows:

$$E = (.35)(0.0016)(4.572E+09)(3.431652879)/(1.764118534) = 5.0E+06 \text{ g (5.0 metric tons).}$$

<sup>7</sup>PM10 emissions also would be created during dozer grading of the storage and waste piles. The particulate emission rate would be a function of the silt content of the soil and the percent moisture content. The ER is calculated as 1.2 g/sec.

It is estimated that grading for the storage and waste piles would take approximately 3,180 hours. The PM10 emissions for pile grading would be as follows: Pile grading emissions = 1.2 grams per second x 3,180h x 3,600 sec/h = 1.374E+07 (13.74 metric tons).

<sup>8</sup>Storage and waste piles would be continuously active. The total suspended particulates caused by wind erosion of contiguously active piles would be a function of the percent silt in the aggregate, the number of days with precipitation above 0.01 in. during a year, and the fraction of time the wind is above 5.4 m/sec at the mean pile height. Using parameters applicable for the Hanford Site and Extensive Pretreatment alternative specifics, the EF is calculated to be: EF = 0.060 grams per m<sup>2</sup> per day.

Table F-7. Backup for Table 9-11. Nonradiological Construction Emission.  
Summary of PM10 Fugitive Emissions in Metric Tons.

Notes: (continued)

The storage pile area is estimated at 23 acres with a surface area of 1,002,000 ft<sup>2</sup> and about 30 ft high. The pile would be assumed to exist for a 2-year period. Assuming that the fraction of total suspended particulates that is PM10 is 50 percent, the tonnage of PM10 particulate is calculated as follows:

$$\text{PM10} = 0.060 \text{ gr}/(\text{m}^2 \text{ day}) \times 1,002,000 \text{ ft}^2 \times 2 \text{ yrs} \times 365 \text{ days/yr} \times 0.5 = 2.194\text{E}+07 \text{ g (21.94 metric tons)}.$$

<sup>9</sup>Excavation would be done with a front end loader that excavates soil and loads it into a dump truck. The dump truck would travel 500 ft to a crusher and dump the load. Another front end loader would feed the crusher which would produce aggregate and reject material. Both materials, now wet, would be moved on a separate conveyor belt to form separate storage piles. A front end loader would pick up the reject material and put it into a dump truck which returns it to the borrow pit. A front end loader would pick up the crushed material and put it into a dump truck which would travel 5 miles over paved roads to the batch plant, located on the construction site, and dump the load. Emission sources include material transfers and unpaved road travel. No emissions are assumed to be generated at the conveyor belt since all materials are wet.

<sup>10</sup>Transfer operations include the following: unprocessed aggregate from a front end loader to a dump truck, dumping, feeding material into the crusher, loading processed aggregate and reject into dump truck with front end loaders, and dumping. Emissions are calculated as follows:

$$\text{Mass of material excavated is } 486,510 \text{ yd}^3 \times 108 \text{ lbs/ft}^3 \times 27 \text{ ft}^3/\text{yd}^3 \times 0.4536 \text{ kg/lb} \times 5 = 3.22\text{E}+09 \text{ kgs}.$$

The emissions are then calculated as follows:

$$E = (0.35) \times (0.0016) \times (3.22\text{E}+09) \times (3.431652879)/(1.764118534) = 3.51\text{E}+06 \text{ g (3.51 metric tons)}.$$

<sup>11</sup>Based on the model used, the EF would be 659 g/km. The EF would be used to calculate the total emissions for the 16,000 round trips required to excavate the quantity of aggregate. An additional 4,600 round trips would be required to dump the reject back into the borrow pit. Once loaded, the dump truck would carry the excavated raw aggregate an average of 500 ft (.152 km) to the crusher. The same distance is assumed to truck the reject back to the pit.

<sup>12</sup>Unpaved road emissions are calculated as follows:  $659 \text{ g/km} \times 0.1524 \text{ km/trip} \times 41000 = 4.1\text{E}+06 \text{ g (4.1 metric tons)}$ .

<sup>12</sup>Material transfers have already been considered.

<sup>13</sup>Fugitive emissions from batch plant operations is estimated at  $2.20\text{E}+01$  metric tons.

<sup>14</sup>Wind erosion for aggregate and sand storage piles is estimated as follows: aggregate pile, 0.09 metric tons ( $9.0\text{E}-02$ ) sand pile, 0.19 metric tons ( $1.9\text{E}-01$ ).

<sup>15</sup>Total emissions from aggregate transfer from the borrow pit to the batch plant is estimated at 213.8 metric tons ( $2.1\text{E}+02$ ).

<sup>16</sup>Unpaved road emissions from concrete transfer to the construction site is estimated at 1.84 metric tons ( $1.8\text{E}+00$ ).

Table F-8. Backup to Table 9-16. Process Modules: Overall Cost.

	Construction	Operating Labor	Operating Equipment	Materials and Supplies			LLW Vaults	HLW Canisters	R&D	Repository Fee	Total
				Start-up <sup>1</sup>	D&D <sup>2</sup>	Operation					
Radionuclide removal	\$2,778	\$1,159	\$126	\$92	\$728	\$569			\$282		\$5,733
Central facilities	\$638										\$638
LLW vitrification	\$749	\$308	\$224	\$24	\$193	\$151			\$268		\$1917
LLW disposal		\$14	\$14	\$1	\$9	\$7	\$190		\$14		\$248
HLW vitrification	\$672	\$320	\$28	\$25	\$201	\$157			\$282		\$1,685
HLW transportation		\$3		\$0	\$2	\$1					\$6
HLW disposal								\$11		\$491	\$502
Total	\$4,837	\$1,803	\$392	\$143	\$1,132	\$885	\$190	\$11	\$846	\$491	\$10,730

## Note:

D&D = decontamination and decommissioning

HLW = high-level waste

LLW = low-level waste

R&D = Research and Development

<sup>1</sup>Start-up and operation materials and supplies are allocated based on operation labor.

<sup>2</sup>Decontamination and decommissioning materials and supplies are allocated based on construction cost.

Table F-9. Backup to Table 9-18. Capital Costs (1995 Dollars).

Components	Extensive Pretreatment Facility	Centralized Facilities	LLW Vitrification Facility <sup>1</sup>	HLW Vitrification Facility <sup>2</sup>	Total All Facilities
Labor	1,210,606,000	318,524,000	2,678,586,000	272,903,000	2,778,350,000
Materials and supplies	479,276,000	202,194,000	98,837,000	147,518,000	748,665,000
Equipment	1,088,477,000	117,735,000	381,972,000	251,358,000	671,779,000
Total	2,778,359,000	638,453,000	748,665,000	671,779,000	4,837,256,000

## Notes:

HLW = high-level waste

LLW = low-level waste

<sup>1</sup>Low-level waste vitrification (low source), Option 2A.

<sup>2</sup>High-level waste vitrification facility is combined with extensive pretreatment.

Table F-10a. Backup to Table 9-19. Operating Cost Component (Millions of 1995 Dollars)  
 Extensive Pretreatment, Annual Consumables for Extensive Pretreatment with High-level  
 Vitriification Detached Low-Level Waste Vitriification--Option 2A. (2 sheets).

Utility/Chemical description	Unit Cost (\$/each)	Quantity Units	Cost
<b>Glass Former</b>			
B <sup>2</sup> O <sup>3</sup>	\$1,000	57,500	\$57,500,000
Fe <sup>2</sup> O <sup>3</sup>	\$500	14	\$7,000
LiO <sup>2</sup>	\$5,000	2,100	\$10,500,00
SiO <sup>2</sup>	\$40	287,000	\$11,480,000
MgO	\$500	4,100	\$2,050,000
CaO	\$60	4,100	\$246,000
Nitric acid 50 percent (Mg)	\$160	465,000	\$74,400,000
NaOH, 505 (Mg)	\$250	150,000	\$37,500,000
Flocculant (Mg)	\$1,100	150	\$165,000
Glycolic acid, 50 percent (Mg)	\$1,740	7,700	\$13,398,000
Oxalic acid (Mg)	\$860	11,900	\$10,234,000
Ammonia (Mg)	\$350	9,240	\$3,234,000
Ion exchange media (m <sup>3</sup> )	\$10,600	4,910	\$52,046,000
FeSA (Mg)	\$660	4,720	\$3,115,200
Aluminum nitrate (Mg)	\$460	112	\$51,520
CMPO (Mg)	\$6,530	5	\$32,650
TBP (Mg)	\$6,530	755	\$4,930,150
NPH (Mg)	\$440	703	\$309,320
Sodium Carbonate (Mg)	\$170	180	\$30,600
Crown ether (Mg)	\$6,530	284	\$1,854,520
Formic acid, 96 percent (Mg)	\$1,210	1,470	\$1,778,700
HF (Mg)	\$500	740	\$370,000
Sodium oxalate (Mg)	\$500	20,900	\$10,450,000
Sodium bicarbonate (Mg)	\$500	12,500	\$6,250,000
Zinc nitrate (Mg)	\$500	47	\$23,500
Na <sup>3</sup> DTPA (Mg)	\$500	196	\$98,000
APM (Mg)	\$500	10	\$5,000
Hydroxylamine nitrate (Mg)	\$500	95	\$47,500
DCPD (Mg)	\$150	4,480	\$672,000
CPD (Mg)	\$150	4,480	\$672,000

Table F-10a. Backup to Table 9-19. Operating Cost Component (Millions of 1995 Dollars)  
 Extensive Pretreatment, Annual Consumables for Extensive Pretreatment with High-level  
 Vitrification Detached Low-Level Waste Vitrification—Option 2A. (2 sheets).

Utility/Chemical description	Unit Cost (\$/each)	Quantity Units	Cost
Decontamination chemicals (Mg)	\$401	14,000	\$5,614,000
Sulfur (Mg)	\$370	170,100	\$62,937,000
Grout powders (Mg)	\$100	0	\$0
Kerosene (Mg)	\$150	70,170	\$10,525,500
Raw water (m <sup>3</sup> )	\$0.03	30,000,000	\$900,000
Sanitary water (m <sup>3</sup> )	\$0.03	1,100,000	\$33,000
Electricity (MWh)	\$30	16,600,000	\$498,000,000
Subtotal			\$881,460,160
Solid waste (m <sup>3</sup> )	\$1,000	4,000	\$4,000,000
Equipment (per year x 14 years)	\$28,000,000	14	\$392,000,000
Vaults	\$5,000,000	38	\$190,000,000
Canisters	\$10,000	448	\$4,480,000
Containers	\$25,000	0	\$0
Overpacks	\$60,000	112	\$6,720,000
Subtotal			\$597,200,00
Total			\$1,478,660,160
Total minus equipment			\$1,086,660,160
Start-up costs (materials)			\$143,000,000
Decontamination and decommissioning (material)			\$1,132,000,000
Total			\$2,753,660,160
Total minus equipment			\$2,361,660,160

## Notes:

- HLW = high-level waste  
 LLW = low-level waste  
 m<sup>3</sup> = cubic meters  
 Mg = megagrams

Annual equipment purchase estimated by K. D. Boomer.

- Pretreatment = \$ 9 million per year (size increase)  
 HLW vitrification = \$ 2 million per year (melter)  
 LLW vitrification = \$ 16 million per year (melter)  
 Sub-total = \$ 27 mm per year  
 Miscellaneous = \$ 1 million per year  
 Total = \$28 million per year

Table F-10B. Backup to Table 9-19. Operating Cost Component  
(Millions of 1995 Dollars).

Staffing Category	Years	Staff Per Year	Total Staff Years	Total Staff Hours
Extensive separations	15	620	9,300	1.7E+07
LLW vitrification	15	160	2,400	4.3E+06
HLW vitrification	15	160	2,400	4.3E+06
Indirect staffing	19	29	551	1.0E+06
Pretreatment start-up	1.5	626	939	1.7E+06
LLW vitrification start-up	1.5	157	236	4.3E+05
HLW start-up	1.5	157	236	4.3E+05
Pretreatment decontamination and decommissioning	2	626	1,252	2.3E+06
LLW vitrification decontamination and decommissioning	2	157	314	5.7E+05
HLW decontamination and decommissioning	2	157	314	5.7E+05
HLW monitoring and maintenance	12	10	120	2.2E+05
HLW transportation	--	30	30	5.4E+04
<b>Total</b>				<b>3.3E+07</b>

## Notes:

HLW = high-level waste

LLW = low-level waste

All staff hours are based on a staff-year of 1,812 hours.

Table F-10C. Backup to Table 9-19. Operating Cost Component  
(Millions of 1995 Dollars).

Operation Mode	Exempt	Nonexempt	Bargaining Unit	Total	Cost Per Year
Pretreatment	202	40	378	620	\$63
LLW vitrification	52	10	98	160	\$16
HLW vitrification	52	10	98	160	\$16
Total	306	60	574	940	\$95

## Notes:

HLW = high-level waste

LLW = low-level waste

Table F-10D. Backup to Table 9-19. Operating Cost Component  
(Millions of 1995 Dollars).

Staffing Category	Total cost
Extensive separations	\$940
LLW vitrification	\$243
HLW vitrification	\$243
Indirect staffing <sup>1</sup>	\$33
Pretreatment start-up <sup>2</sup>	\$94
LLW vitrification start-up <sup>2</sup>	\$24
HLW start-up <sup>2</sup>	\$24
Pretreatment decontamination and decommissioning <sup>3</sup>	\$125
LLW vitrification decontamination and decommissioning <sup>3</sup>	\$32
HLW decontamination and decommissioning	\$32
HLW monitoring and maintenance <sup>3,4</sup>	\$8
HLW transportation <sup>5</sup>	\$3
Total	\$1,803

Notes:

<sup>1</sup>For indirect staffing, it was assumed that all workers would be nonexempt.

<sup>2</sup>The total start-up cost for each facility is arbitrarily set to the annual staffing requirements. These start-up requirements have been set to 1 1/2 years of the annual. This leaves 1 1/2 years of staff cost (\$95 million + \$24 million + \$24 million = \$143 million) for start-up materials and supplies.

(\$63 million x 1.5 yrs = \$95 million) (\$16 million x 1.5 yrs = \$24 million) (\$16 million x 1.5 yrs = \$24 million).

<sup>3</sup>Decontamination and decommissioning for each facility is arbitrarily set equal to three years of staff cost plus thirty percent of the total capital cost minus the contingency. The staffing requirements have been set to two years of the annual. This leaves one year of staff cost (\$63 million + \$16 million + \$16 million = \$95 million) plus thirty percent of the capital for material and supplies.

The total capital is equal to the capital plus a forty percent contingency; therefore, the capital value to be used for the decontamination and decommissioning cost for materials and supplies is equal to:  
Capital \$4,837 million per 140 percent = \$3,455 million

The cost based upon the capital would be as follows: capital decontamination and decommissioning = \$3,455 million x 30 percent = \$1,037 million. Thus, the total decontamination and decommissioning cost for materials and supplies is set to \$1,132 million.

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Table F-10D. Backup to Table 9-19. Operating Cost Component  
(Millions of 1995 Dollars).

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Notes (continued):

<sup>4</sup>For monitoring and maintenance, the following was assumed: exempt would be 10 percent, bargaining unit 10 percent, and nonexempt 80 percent.

<sup>5</sup>For HLW transportation, the following was assumed: exempt would be 36 percent, bargaining unit 36 percent, and nonexempt 28 percent.

All exempt and bargaining unit employees are assumed to be radiation workers. All nonexempt employees are nonradiation workers.

Table F-11. Backup to Table 9-20. Overall Schedule.

Construction			
Option	Start	Finish	Facility Configuration Study
Pretreatment + HLW vitrification	July 2001	December 2006	page 114
Standalone LLW vitrification	December 1997	July 2005	page 63
Overall	December 1997	December 2006	
Operation			
Pretreatment	January 2004	January 2018	pages 114, 33
HLW vitrification	March 2005	March 2019	pages 114, 33
LLW vitrification	September 2005	September 2019	pages 63, 33
Overall	January 2004	September 2019	
Decontamination and decommissioning for all facilities is assumed to start after the completion of process duration, that is, 5 years.			
	October 2019	October 2024	
<p>Monitoring and maintenance is for HLW canisters.</p> <p>Monitoring and maintenance is assumed to start with the completion of HLW vitrification.</p> <p>Monitoring and maintenance is assumed to finish when the last multi-purpose canisters is shipped to the repository.</p> <p>Shipments to the repository start in 2035<sup>1</sup>.</p> <p>Based on discussions with K. D. Boomer, 1,000 canisters are assumed, 250 multi-purpose canisters.</p> <p>10 multi-purpose canisters would be transported per week (Slaathaug 1995, Table 9-7, footnote 5).</p> <p>Duration of shipments = 250 multi-purpose canisters/10 = 25 weeks.</p>			
	March 2019	September 2035	<sup>1</sup> page 151
Research and Development			
	1995	2018	

Notes:

HLW = high-level waste  
 LLW = low-level waste

**APPENDIX G**

**SITE LAYOUTS**

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## APPENDIX G

### SITE LAYOUTS

This appendix includes the following figures:

- Figure G-1. Disturbed Area.
- Figure G-2. Temporary Facilities and Laydown Areas.
- Figure G-3. Construction Aggregate Pit.

Site layouts for the Extensive Separations Pretreatment (ESP) alternative with Low-Level Waste (LLW) vitrification treatment and optionally with LLW grout treatment are shown in Section 8.0, Figures 8-2 and 8-2, respectively.

Figures G-1 through G-3, which are associated with the Tri-Party Agreement preferred alternative (Slaathaug 1995), also provide the information required for the ESP alternative. Figure G-1 shows the areas disturbed by construction and operation of the two facilities [i.e., pretreatment/LLW vitrification and high-level waste (HLW) vitrification]. Figure G-2 displays the temporary facilities and laydown areas associated with construction. Figure G-3 gives the location of the construction aggregate pit and the relative proximity of it to the construction site.

Each facility provides most required process support equipment. Common utilities and cold chemical areas provide common headers for services to support individual systems in the plants. Common services include the following:

- Medium pressure steam (consumed) and condensate
- Compressed air (instrument/plant air)
- Cooling tower
- Water (sanitary, process, demineralized, raw, and fire)
- Sanitary sewer
- Non-radioactive liquid waste processing
- Cold chemicals bulk storage and make-up (including glass formers)
- Bulk cold chemical building vent system
- Oxygen
- Electricity.

The medium pressure steam system provides consumed steam (e.g., steam jets) to the facilities as a shared utility. The steam trap condensate collection systems are located in individual facilities. Certain systems within individual processing plants require closed loop steam (to minimize the amount of potentially radioactive material leaving the area). Steam is

provided by packaged process steam generators located within each facility. The closed loop process steam system is independent of the medium pressure steam system.

The compressed air system provides instrument air for pneumatically controlled components, plant air for spargers, jets, and general maintenance use. Plant air is provided preferentially to critical plant air users. This utility source is located in the Mechanical Utilities Building.

The cooling tower system removes heat from the normal processing operations of the cooling water systems, and rejects the heat by evaporative cooling. It includes a cooling tower, cooling tower water circulation pumps, inhibitor addition pump, and distribution piping. Raw water is used for cooling tower make up. Bleedoff is routed to the 200 Area Treated Effluent Disposal Facility.

The sanitary water system supplies water for domestic uses and for heating, ventilating and air conditioning humidifiers; it is also the source for the process water system. The sanitary water system consists of distribution piping installed in a loop around its users/facilities.

The process water system is the source water for low-pressure process water tanks and other users in the separations, vitrification treatment facilities and annexes, the Bulk Cold Chemical Building, the Waste Staging and Handling Facility, and demineralizer and other users in the Mechanical Utilities Building. The process water system is the source for chemical dilution water, priming for pumps, and equipment flushing.

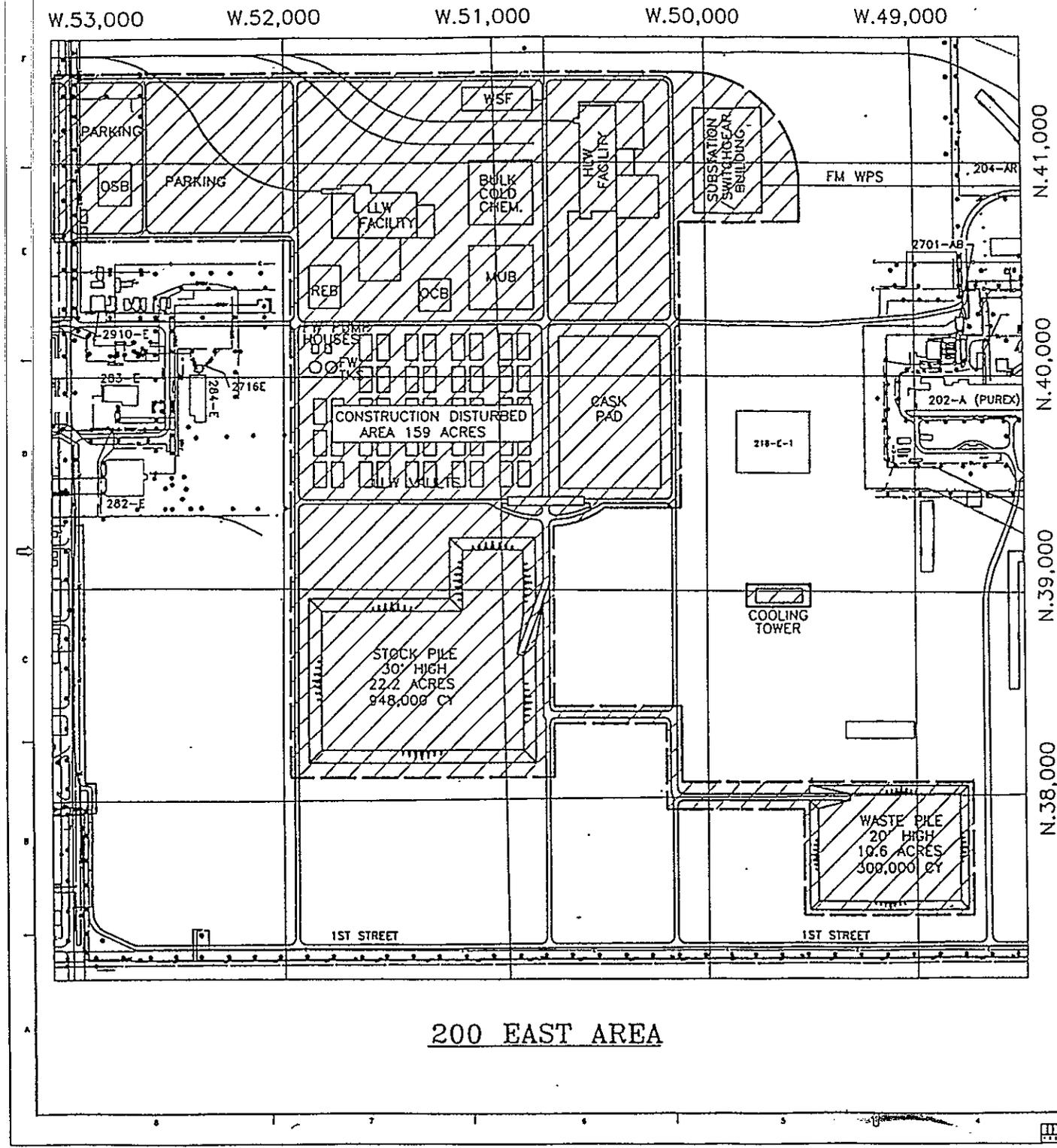
Sanitary water, separated from the process water by a backflow preventer, is the source for the process water system.

The nonradioactive liquid waste processing system receives, collects, stores, and disposes of all nonradioactive liquid waste in a safe and environmentally acceptable manner. Chemically contaminated liquid wastes are sufficiently neutralized prior to transfer to the 200 Area Treated Effluent Disposal Facility. The nonradioactive liquid waste processing system includes hold tanks, neutralization tanks, sumps, and pumps.

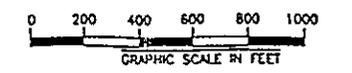
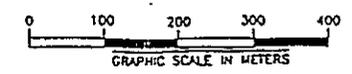
The cold chemicals bulk storage and make-up system includes all facilities required to receive, store, prepare, and feed cold chemicals to process users. It includes storage tanks, make-up tanks, feed tanks, sump tanks, and appurtenances, such as pumps, agitators, heaters, and distribution facilities. Glass former bins and transport equipment are also a part of this system.

The site will require approximately 120 MVA to support operations. A new power line will be provided to the 200 Area. It will provide a new 230 KV power line approximately 2-km long to the TWRS Treatment Complex Switch Yard.

Figure G-1. Disturbed Area.



CASE 8A EIS



NO. 1	NO. 2	NO. 3	NO. 4	NO. 5	NO. 6	NO. 7	NO. 8	NO. 9	NO. 10

PROJECT TITLE	ALTB8A-04	PROJECT NUMBER	28184/ACD:10.02.55
ENGINEERING RELEASE	REV	DATE	
DESIGN			
APPROVED BY			
DATE			
U.S. DEPARTMENT OF ENERGY			
Fluor Daniel, Inc. Advanced Technology Division			
EIS	TPA PREFERRED ALTERNATIVE		
DISTURBED AREA	SITE PLAN		
DATE WHEN PROJECTOR STOPPED			
PROJECT NUMBER	28184/ACD:10.02.55	DATE	2013
DESIGNER	J. ELSON	DATE	2007-07
NO. 1	NO. 2	NO. 3	NO. 4
NO. 5	NO. 6	NO. 7	NO. 8
NO. 9	NO. 10	NO. 11	NO. 12
NO. 13	NO. 14	NO. 15	NO. 16
NO. 17	NO. 18	NO. 19	NO. 20
NO. 21	NO. 22	NO. 23	NO. 24
NO. 25	NO. 26	NO. 27	NO. 28
NO. 29	NO. 30	NO. 31	NO. 32
NO. 33	NO. 34	NO. 35	NO. 36
NO. 37	NO. 38	NO. 39	NO. 40
NO. 41	NO. 42	NO. 43	NO. 44
NO. 45	NO. 46	NO. 47	NO. 48
NO. 49	NO. 50	NO. 51	NO. 52
NO. 53	NO. 54	NO. 55	NO. 56
NO. 57	NO. 58	NO. 59	NO. 60
NO. 61	NO. 62	NO. 63	NO. 64
NO. 65	NO. 66	NO. 67	NO. 68
NO. 69	NO. 70	NO. 71	NO. 72
NO. 73	NO. 74	NO. 75	NO. 76
NO. 77	NO. 78	NO. 79	NO. 80
NO. 81	NO. 82	NO. 83	NO. 84
NO. 85	NO. 86	NO. 87	NO. 88
NO. 89	NO. 90	NO. 91	NO. 92
NO. 93	NO. 94	NO. 95	NO. 96
NO. 97	NO. 98	NO. 99	NO. 100



