3.1.3.1 Combustion Modifications

The major concerns regarding application of CM on oil-fired boilers are effects on boiler efficiency, load capacity, boiler vibration, flame instability, and steam and tube temperatures. These are summarized in Table 3-10. OSC operation generally increases the minimum excess air requirements of the boiler. In many cases, BOOS operation in oil-fired boilers has been found to be more effective in controlling NO_x than OFA firing (Table 3-11). Under BOOS firing the the fuel flow to the active burners must be increased if load is to remain constant. In some cases, it has been necessary to enlarge the burner tips in order to accommodate these increased flows.

Other potential problems attendant with applying OSC in oil-fired boilers have concerned flame instabilities, boiler vibrations, and excessive convective section tube temperatures. However, in past experience, none of these problems has been significant. Staged operation usually results in hazy flames and obscure flame zones. Thus the new flame scanners and detectors are often required in retrofit applications. In addition, because OSC produces an extended flame zone, flame carryover to the convective section may occur occasionally.

NO_x controls (combustion modifications) currently in use in the SCAQMD on boilers in the 175 to 240 MW range which are predominantly front, wall-fired units are TSC and OSC (Table 3-4). The NO_x emissions range from 210 to 425 ppm (dry, 3% O_2) but are generally in the 225 to 300 ppm range. The amount of NO_x reduction achieved as a result of CM is varied, from none (or what appears to be an anomalous increase at Alamitos 1 and 2 for TSC) to 15% for Redondo Beach units 5 and 6. For off-stoichiometric operation, 18 to 42% reduction was observed, and for combined OSC/TSC the amount was 25%. Data from unidentified boilers in the 130 to 250 MW range are reported in Ref. 3-8 as 25 to 31% reduction for OSC.

TABLE 3-10. NITRIC OXIDE CONTROL TECHNIQUES (Ref. 3-12)

Class of Technique	Method in Practice	Application Considerations
1. Off-stoichiometric firing (fuel-rich flames do not produce NO)	a. Low excess air	Reduced operating cost but may adversely effect steam temperature
	b. Off-stoichiometric optimization	Burner pattern selection to control CO and smoke at minimum nitric oxide
	c. Staged-combustion ports	Allows versatility in patterns
2. Direct flame temperature reductions	a. Flue gas recirculation	Increased fan usage and boiler pressure, changes heat transfer and may be more effective on gas than on oil or coal.
	b. Water injection in air	Enthalpy effect not concentrated in the primary combustion zone, lower boiler efficiency, and increased plume appearance
	c. Reduced air preheat	Enthalpy effects same as water injection but with increased plume size and decreased visibility and equipment fouling.
3. Indirect flame temperature reduction	a. Increased flame zone heat transfer area	Equipment redesign; heat transfer area is not effective when combustion takes place faster than heat transfer
	b. Reduced burner flow	Greatly increased equipment and control complications
	c. Increased burner spacing	More efficient use of flame zone heat transfer
	d. Flame zone cleanliness	Constantly varies in practice
	e. Division walls	Increased heat transfer may cause flame quenching and with stratification may cause CO and smoke problems

TABLE 3-11. COMPARISON OF AVERAGE NO_x REDUCTION IN UTILITY BOILERS FROM COMBUSTION MODIFICATIONS INCLUDING LOW NO_x BURNERS

Control Method	Emissions, 175-250 MW Boilers, U.S.			Japan[c]	
	ppm	Overall Reduction, %[b]	Δ %	Overall Reduction, %	Δ %
2-stage (TSC)	288[a]	20	--	27	--
OSC (BOOS)	238[a]	34	14[d]	18	-9[d]
OSC + 2-stage	271[a]	25	5[d]	--	--
Low NO_x Burner (LNB)	--	--	--	27	--
TSC + LNB	263(25)[g]	27	7[f]	34	7[d]
OSC + LNB	213(25)[g]	41	7[f]	25	7[e]
Flue Gas Recirc (FGR)	259	28[h]	--	32	--
TSC + FGR	223(36)	38	10[f]	42	10[j]
OSC + FGR	198(61)	45[h]	17[h]	--	--
TSC + LNB + FGR	227(36)	37	10[l]	44	10[l]

[a] Ref: Table 3-4.

[b] Relative to uncontrolled emissions

[c] Ref. 3-11.

[d] Relative to TSC

[e] Relative to OSC

[f] Assumed same as Japan

[g] ppm reduction equal to 7% overall (360 ppm baseline, uncontrolled)

[h] Ref. 3-7.

[i] ppm reduction equal to 10% increment

[j] Relative to FGR

[k] ppm reduction equal to 17% increment

[l] Relative to LNB + TSC (i.e., FGR increment)

3.1.3.2 Low NO_x Burners

The average performance of CM, LNB, and combinations is summarized in Table 3-11, based on experience in the Los Angeles area and in Japan. It is recognized, especially in view of the wide range of reduction attained (Table 3-4), that the average is not indicative of the control potential on individual units. Also, it should be noted that actual experience cited on utility boilers by users and manufacturers of LNB in Japan (Table 3-12) exceeds the average value by approximately 100%, i.e., 54 to 57% reduction vs 27% average reduction. However, it is nonetheless informative in several ways to examine average performance in terms of (1) a realistic expectation on retrofitting utility boilers with LNB, (2) the potential incremental improvement that may be achieved by combining LNB with existing CM, and (3) an examination of other potential CM alternatives.

On the basis of experience in Japan, the incremental reduction in NO_x for use of LNB in combination with TSC or OSC is 7% (Table 3-11). Assuming that the same characteristics can be achieved here, a reduction of 25 ppm (55 lb/hr) can be expected at full load by retrofitting LNB on 175 MW units with existing TSC or OSC modifications.

In addition to the potential for significant NO_x reduction (up to 55%) by the use of LNB without other modifications, LNB has the advantage of eliminating or decreasing the need for reducing or near-reducing conditions near furnace walls (Ref. 3-10). Therefore, corrosion problems associated with reducing atmospheres should not arise with these systems. Although the LNB flames can be expected to be less turbulent and hence longer than flames from normal burners, the combustion zone will probably not extend any further up the furnace than with OSC. Potential changes in heat absorption profile and excessive steam and tube temperatures are, therefore, less likely to occur.

As fuel and airflows are controlled more closely in LNB-equipped systems, nonuniform distribution of fuel-air ratios leading to excessive carbon monoxide generation or high excess air requirements should be eliminated. Boiler efficiencies should, therefore, not be affected by installation of LNB. In combination with existing NO_x control measures, LNB combined with OSC are estimated to reduce emissions from 238 to 213 ppm (Table 3-13).

TABLE 3-12. EXAMPLE OF SPECIFIC LOW NO_x BURNER PERFORMANCE IN JAPANESE UTILITY BOILERS

Site	Utility Boiler Size, MW	Type of LNB Installation	Fuel	Fuel N, %	Emissions, ppm		% Reduction	Reference	No. of Burners, Firing Configuration
					Uncontrolled	Controlled			
A	350	R**	Crude	0.04	300	130[a]	57[a]	3-13	-
B	600	R	LNG	-	(90)[e]	60	-	3-13	-
Chubu Elec, Chita Station									
No. 5 & 6	700	N**	LNG	-	-	38[b]	-	3-14	32 Opposed
2	375	R	Crude Oil	~0.1	~400	140	c	3-14	18 Front
3	500	R	Heavy Oil	~0.2	~400	100	c	3-14	40 Corner
4	700	R	Crude Oil	~0.1	~400	100	c	3-14	40 Corner
TEPCO, Yokosuka 1B	265	R	Heavy Oil	0.18	300-350	61	81[d]	3-15	Corner
TEPCO, Kashima 1B	600	R	Heavy Oil	0.14	300-350	80	75[d]	3-15	Corner
TEPCO, Ooi 1B	350	R	Crude Oil	0.11	300-350	78	76[d]	3-15	Corner
TEPCO, Chiba 1B	125	R	Crude Oil	0.14	300-350	62	81[d]	3-15	Corner
	350	R	Heavy Oil	0.20	300-350	80	75[d]	3-15	Corner
	265	R	Heavy Oil	0.18	300-350	62	81[d]	3-15	Corner
C*	125	R	Heavy Oil	0.15	325	150[a]	54[a]	3-16	Front
C*	125	R	Heavy Oil	0.15	325	175	f	3-16	Front
C*	125	R	Heavy Oil	0.15	325	112	66[g]	3-16	Front
C*	125	R	Heavy Oil	0.15	325	63	81[h]	3-16	Front

TABLE 3-12. EXAMPLE OF SPECIFIC LOW NO_x BURNER PERFORMANCE IN JAPANESE UTILITY BOILERS (Continued)

Site	Utility Boiler Size, MW	Type of LNB Installation	Fuel	Fuel N, %	Emissions, ppm		% Reduction	Reference	No. of Burners, Firing Configuration
					Uncontrolled	Controlled			
Test Burner	100×10^4 kcal/hr (4.0×10^6 Btu/hr)	-	Heavy Oil A	~0.1	117[i]	32	72	3-17	-
Test Burner	100×10^4 kcal/hr (4.0×10^6 Btu/hr)	-	Kerosene	-	98[i]	25	74	3-17	-
Test Burner	60×10^4 kcal/hr 2.4×10^6 Btu/hr	-	C_3H_8	-	375[j]	100	73	3-17	-

[a] LNB only
[b] LNB + overfire air (OFA) + FGR
[c] OFA + FGR only
[d] 4% O_2 includes OFA +FGR
[e] Emissions with OFA + FGR
[f] TSC only
[g] TSC + LNB
[h] TSC + LNB + FGR
[i] 3% O_2, No air preheat
[j] Combustion air preheated to 400°F, 3% O_2
[*] U.S. installation: Tests by Foster Wheeler
[**] R = retrofit
N = new

TABLE 3-13. POTENTIAL NO_x REDUCTION IN 175 MW UTILITY BOILERS RELATIVE TO EMISSIONS WITH COMBUSTION MODIFICATIONS[a]

CONTROL METHOD	EMISSIONS		OVERALL REDUCTION[c] %	REDUCTION IN EMISSIONS[d]		
	ppm[b]	lb/hr		ppm	Percent	
					TSC	OSC
2-stage (TSC)	228	691	20	-	-	-
Off-Stoich (OSC)	238	571	34	-	-	-
OSC + TSC	271	650	25	-	-	-
TSC + LNB[e]	263	631	27	25	8.7(60[f])	-
OSC + LNB	213	511	41	25	-	10.5(60)
TSC + FGR	223	535	38	65	22.6(156)	-
OSC + FGR	198	475	45	40	-	16.8(96)

[a]Ref. Table 3-5.

[b]Dry, 3% O_2.

[c]Relative to uncontrolled levels.

[d]Relative to current levels of control.

[e]LNB: Low NO_x burner.

[f]Parens denote lb/hr.

3.1.3.3 Other Considerations

A further examination of Table 3-11 illustrates the effectiveness of FGR as a NO_x control measure. Although recirculation of flue gas in the windbox is not effective as a fuel NO_x countermeasure (Table 2-3), it can be adapted to the boiler and is highly effective for thermal NO_x abatement.

A number of potential problems can occur in retrofit FGR applications. The most common, such as FGR fan and duct vibrations, can usually be avoided by good design. Other problems, such as flame instability, which can lead to furnace vibrations, are caused by the increased gas velocity at the burner throats. Modifications to the burner geometry and design, such as enlarging the throat, altering the burner tips, or adding diffuser plates or flame retainers, may then be required. These modifications are usually made by trial and error for each boiler and are often very time consuming. If the problems of excessive boiler vibration and flame instabilities persist at high loads, the boiler may require derating to operate stably.

Another potential problem associated with FGR is high tube and steam temperatures in the convective section. The increased mass velocities which occur with FGR cause the convective heat transfer coefficient to rise. This, coupled with reduced furnace heat absorption, can give rise to high convective section temperatures leading to tube failures, exceeding attemperator spray flow limits, or loss in cycle efficiency due to excessive reheat steam attemperation. Increased mass flowrates in the furnace may also cause furnace pressures to increase beyond safe limits. However, FGR usually has the advantage of not increasing minimum excess air levels. Boiler efficiency is therefore relatively unaffected except for the power consumed by the FGR or booster fans.

Another effective method is the use of windbox FGR in combination with existing CM. It has the estimated potential, if it can be incorporated into the boiler, of reducing NO_x emissions from 238 to 198 ppm with OSC and 288 to 223 ppm with TSC. SCE has reported an overall 42% reduction, to 203 ppm for an uncontrolled level of 350 in tangential boilers using 15% windbox FGR and OSC (viz., BOOS) (Ref. 3-7).

The problems associated with combining OSC and FGR are also combined. Tube and steam temperature problems in the upper furnaces are particularly exacerbated, as both OSC and FGR tend to increase upper furnace temperatures and convective section heat transfer rates. In addition, boiler efficiencies usually decline slightly with combined OSC and FGR firing because of higher excess air requirements and greater fan power consumption (Ref. 3-10).

Before low NO_x techniques are applied to oil-fired boilers, it is important to assure that the unit is in good operating condition. Uniform burner air and fuel flows are essential for optimal NO_x control. Retrofit NO_x control systems must be designed and installed properly to minimize potential adverse effects. Despite these precautions, problems may arise, such as flame instability or high tube temperatures. In some instances hardware modifications will be required to resolve the problems. In others, increased vigilance will be needed on the part of the boiler operator, and an accelerated schedule of maintenance and overhaul may be required. Many of the problems experienced in the past can now be avoided because of hindsight and experience. Thus, retrofit systems can now be designed and installed with care to avoid any potential adverse effects. New units with builtin LNB, OFA, or FGR systems should function without problems (Ref. 3-10).

The operational impacts of the various modifications are summarized in Tables 2-3 and 3-10. The comparison of the costs for the different alternatives is examined in Section 4.1.1.

3.2 STATIONARY GAS TURBINES

Denitrification technologies for reducing NO_x emissions 90% were examined for application to both simple cycle and combined cycle gas turbines used for electric power generation. An example of a facility using simple cycle gas turbines, i.e., conventional turbines without heat recovery is unit 5 at the SCE Etiwanda Generating Station, Fontana, California. Units 3 and 4 at the SCE Cool Water Generating Station, Daggett, California, are examples of a combined cycle facility.

3.2.1 Simple Cycle Gas Turbine

Simple cycle gas turbines are typically used by electric utilities as peaking units. The normal and projected operation of unit 5 at the Etiwanda Station is 1 to 2% at full load (50 to 150 hr/yr). It operates at 121 MW rated load (actually between 90 and 132 MW, depending on ambient conditions). Projected lifetime is approximately 30 years, and it has been on line since 1969 for a total of approximately 2000 hr.

Unit 5 at the SCE Etiwanda Facility consists of eight aircraft-type turbojet engines that act as gas generators, with the exhaust gases expanding through a final set of turbine stages that drive electric generators. The exhaust gas stream is discharged through a silencer and a single stack.

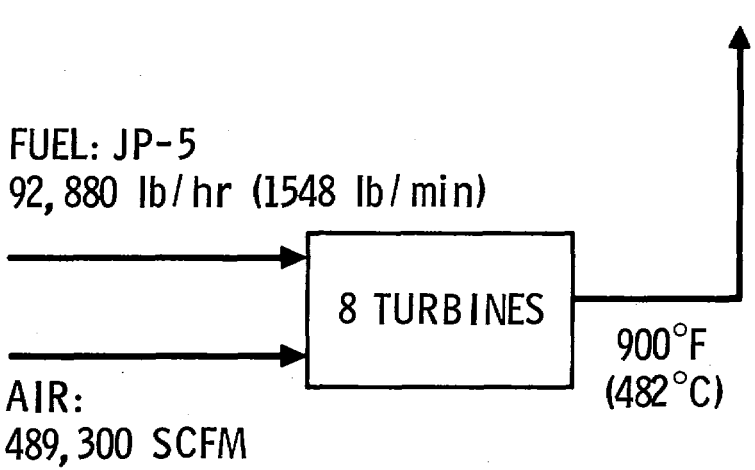
The gas turbines are manufactured by the Pratt and Whitney Division of United Technologies and the overall system, by Turbo Power and Marine Systems, Inc. Natural gas or distillate fuel (JP-5) are burned, the latter at a rate of 327 bbl/hr. JP-5 was considered for this study because of its slightly higher NO_x emissions. The unit heat rate is approximately 15,000 Btu/kWh. Ambient temperature air is used, the overall excess oxygen being 15%. Typically aircraft jet engine turbine inlet temperatures are about 1800 to 2000°F (980 to 1090°C) with the turbine exit temperature approximately 900°F (482°C). The temperature of the gases exiting the stack is 300°F (150°C). Figure 3-1 summarizes the unit characteristics tabulated from questionnaire responses from SCE (Appendix A).

Typical of high performance aircraft engines, the unit cannot be operated for extensive periods, perhaps 200 hr, without major maintenance.

3.2.1.1 NO_x Emissions

Each module of unit 5, comprising two turbine units, emits a total of 147 lb/hr of NO_x emissions (Ref. 3-2), with a total of 588 lb/hr for the entire unit. Emissions are 415 ppm NO_x (dry, 3% O_2). No specific NO_x control measures, such as water injection, are used. The volumetric flow rates of the exhaust gases and NO_x emissions referenced to wet and dry conditions are summarized in Figure 3-1.

SIMPLE TURBINE



	WET	DRY
ACFM (300°F)	785,500	432,000
SCFM	508,300	279,600

	WET	DRY
ACFM (900°F)	1,405,700	773,100
SCFM	508,300	279,600
NO_x (15% O_2), ppm	150	273
NO_x (3% O_2), ppm	228	415

Figure 3-1. Etiwanda unit No. 5 characteristics

3.2.1.2 NO_x Control Alternatives

The NO_x control alternatives reviewed for retrofitting to achieve 90% reduction with simple turbines included CM, including water injection and thermal and selective catalytic $DeNO_x$ with the use of ammonia.

Combustion Modifications

The design of gas turbine combustion systems is such that there is no provision for field adjustments or CM (Ref. 3-18). Work is being conducted to develop combustors that achieve significant reductions in thermal and peak NO_x for designs representative of those employed in current and future stationary gas turbine engines (Ref. 3-19).

Although some units of this type are equipped with water injection systems for NO_x control, the Etiwanda installation was not configured with this capability. Examples of the potential for NO_x reduction by water injection are shown in Figures 3-2 and 3-3 from Refs. 3-20 and 3-15, respectively, with approximately 50% reduction at a water-fuel weight ratio of 0.4. By modifying the distribution of primary, secondary, diluting, and cooling air, Mitsubishi Heavy Industries (MHI) (Ref. 3-15) has achieved significant reductions in NO_x emissions (Figure 3-4). However, this is not considered to have retrofit potential. Further reduction was attained with water injection.

Thermal Denitrification

Typically aircraft jet engine inlet temperatures are in the range of 1800 to 2000°F (980 to 1090°C). The turbine exit temperature is 900°F (482°C). Thus, the optimum temperature for thermal $DeNO_x$ ammonia injection systems, which is 1750°F (954°C), occurs upstream of and within the turbine stage. Therefore, the exhaust gas temperature must be raised to that level by reheating. However, this would require 42,700 lb/hr additional fuel to increase the exhaust gas 418°C. This is an additional amount of fuel equal to 46% of that burned in the turbines. Considering that the potential reduction using this process is about 50% and that the additional NO_x, estimated as 270 to 350 lb/hr, created by the reheat process would further decrease its effectiveness, further consideration of this approach was dropped.

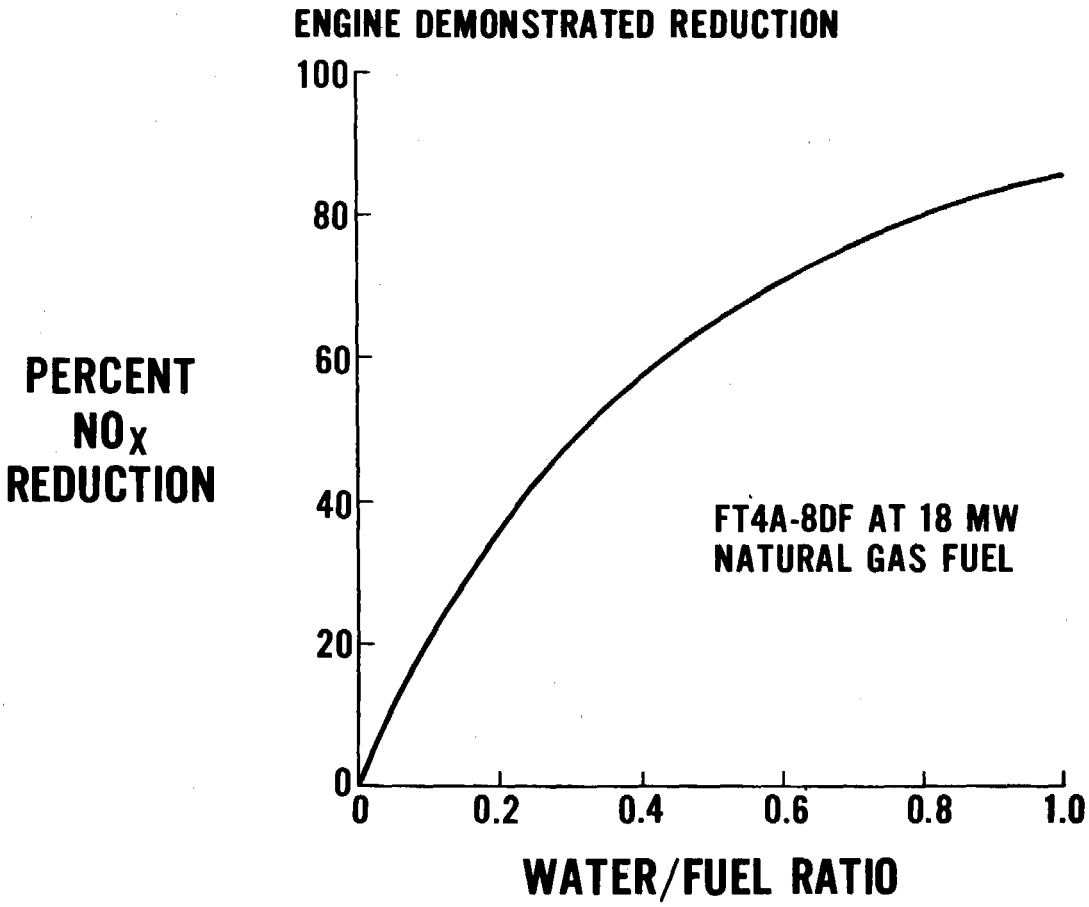


Figure 3-2. Effect of water injection on NO_x emissions from a gas turbine (Ref. 3-19)

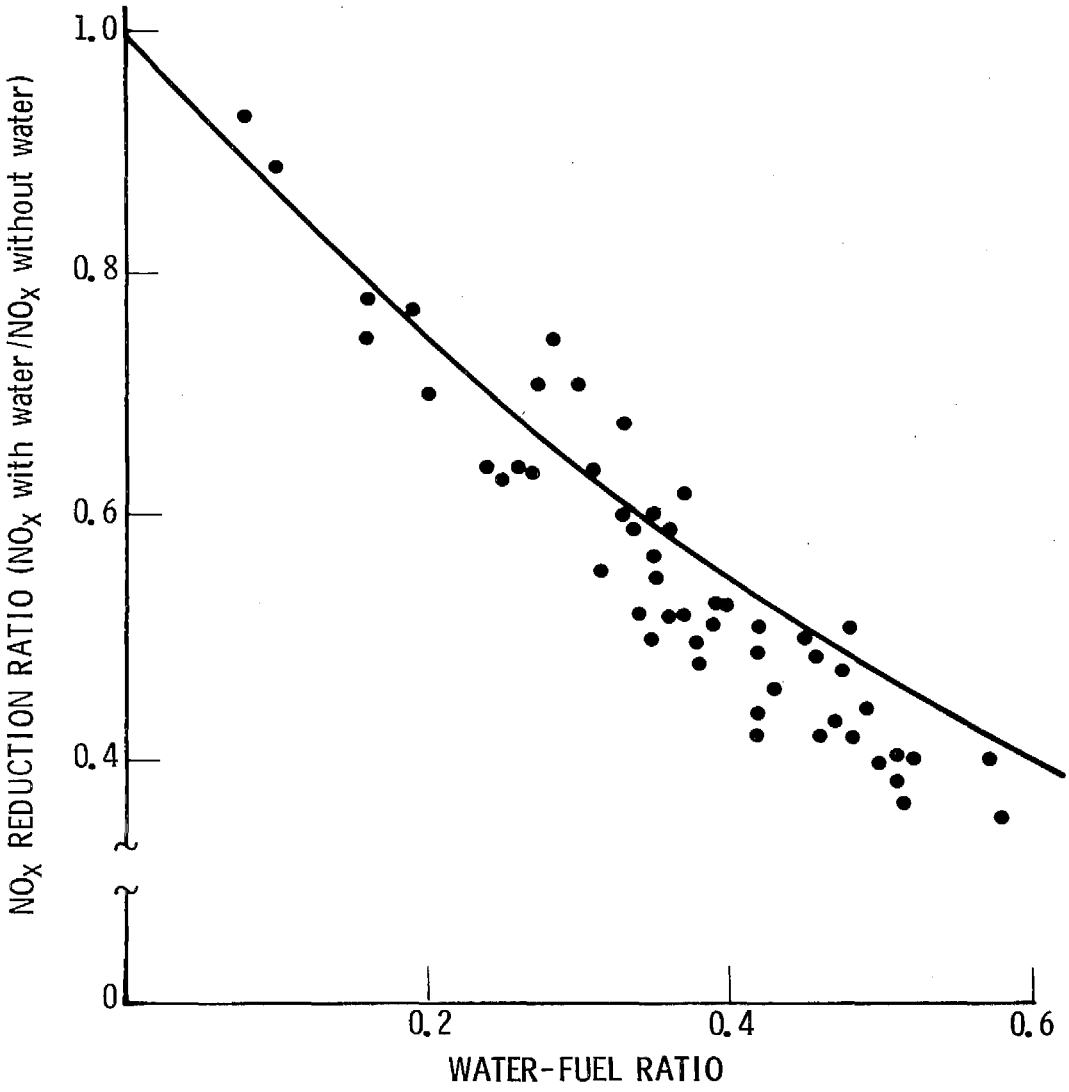


Figure 3-3. Effect of water injection on NO_x emissions (Ref. 3-15)

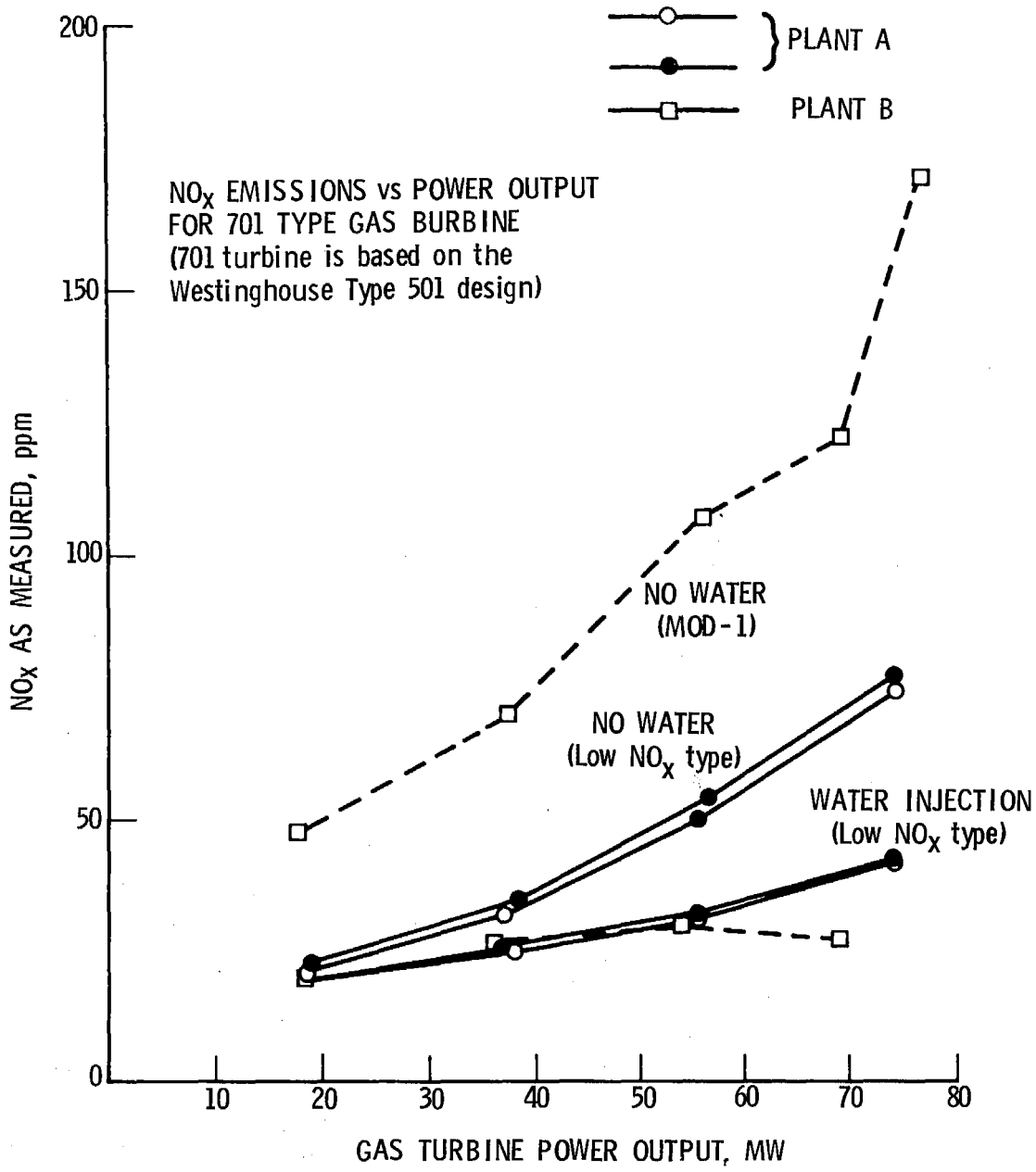


Figure 3-4. NO_x emissions versus power output for 701 type gas turbine (based on the Westinghouse type 501 design) (Ref. 3-15)

Selective Catalytic Reduction

Figure 3-5 illustrates the conditions for use of SCR, with the potential of achieving 90% NO_x reduction. However, site specific factors may make it more or less practical as a retrofit, depending on the space available in the vicinity of the gas turbines. Approximately 313,000 ACFM of ambient air (70°F) must be mixed with the 1,406,000 ACFM turbine exhaust gas (900°F), resulting in 1,564,000 ACFM of gas at 751°F (400°C). The temperature conditions would then be conducive to application of SCR with ammonia, using existing catalysts. Catalyst bed dimensions for each of four parallel reactors with a space velocity of 10,000 hr^{-1} and a superficial velocity of 26 ft/sec (8 m/sec) are approximately 9 × 9 × 9 ft.

For unit 5 at Etiwanda, limited space is available in the building housing the turbine to install the reactors between the turbine exits and the existing stack, which also contains a silencer. However, it may be possible to install the reactors outside and above the building.

Capital and annual control costs for application of SCR to the turbine are presented in Section 4.

3.2.2 Combined Cycle Gas Turbine

Cool Water Generating Station units 3 and 4 are both Westinghouse-packaged systems, designated as "Pace 260". Each of the two units consist of two Westinghouse 501 industrial gas turbines and two waste heat boilers, which drive a single steam turbine and which are manifolded into a single stack.

Both units 3 and 4 can operate at a maximum continuous rating of 472 MW, with each of the four turbines generating about 61 MW. Each gas turbine with its associated waste heat boiler and steam turbine operates in either a full capacity "on" or "off" mode, but the four gas turbines can operate independently to provide output in 25% increments. SCE schedules operate the units at 100% of maximum capacity for approximately 15% of the time, approximately 1300 hr/yr. The projected capacity factor expected is between 15 to 20% until 1983; thereafter it is unknown.

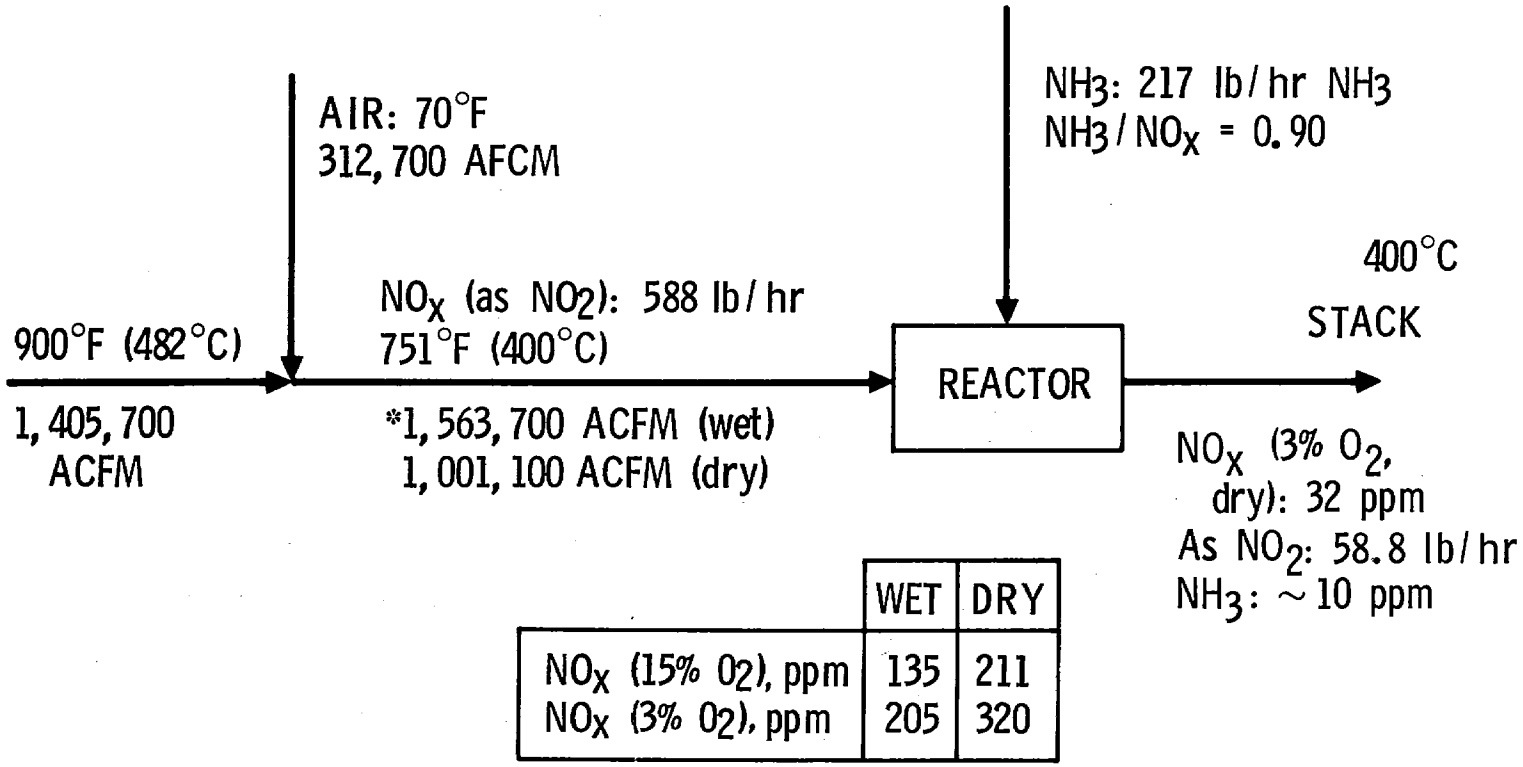


Figure 3-5. Selective catalytic reduction NO_x control schematic

Number 2 fuel oil is burned in the turbines, which consume approximately 80,000 lb/hr per pair. The unit heat rate is reported to be 8890 Btu/kWh. All combustion air enters through the gas turbine inlet at ambient temperature or below (through an evaporative cooler). Oxygen content at the stack is approximately 14%. Turbine exit temperatures are approximately 1096°F (591°C), and the flue gas temperature at the stack is about 300°F (149°C) (Figure 3-6).

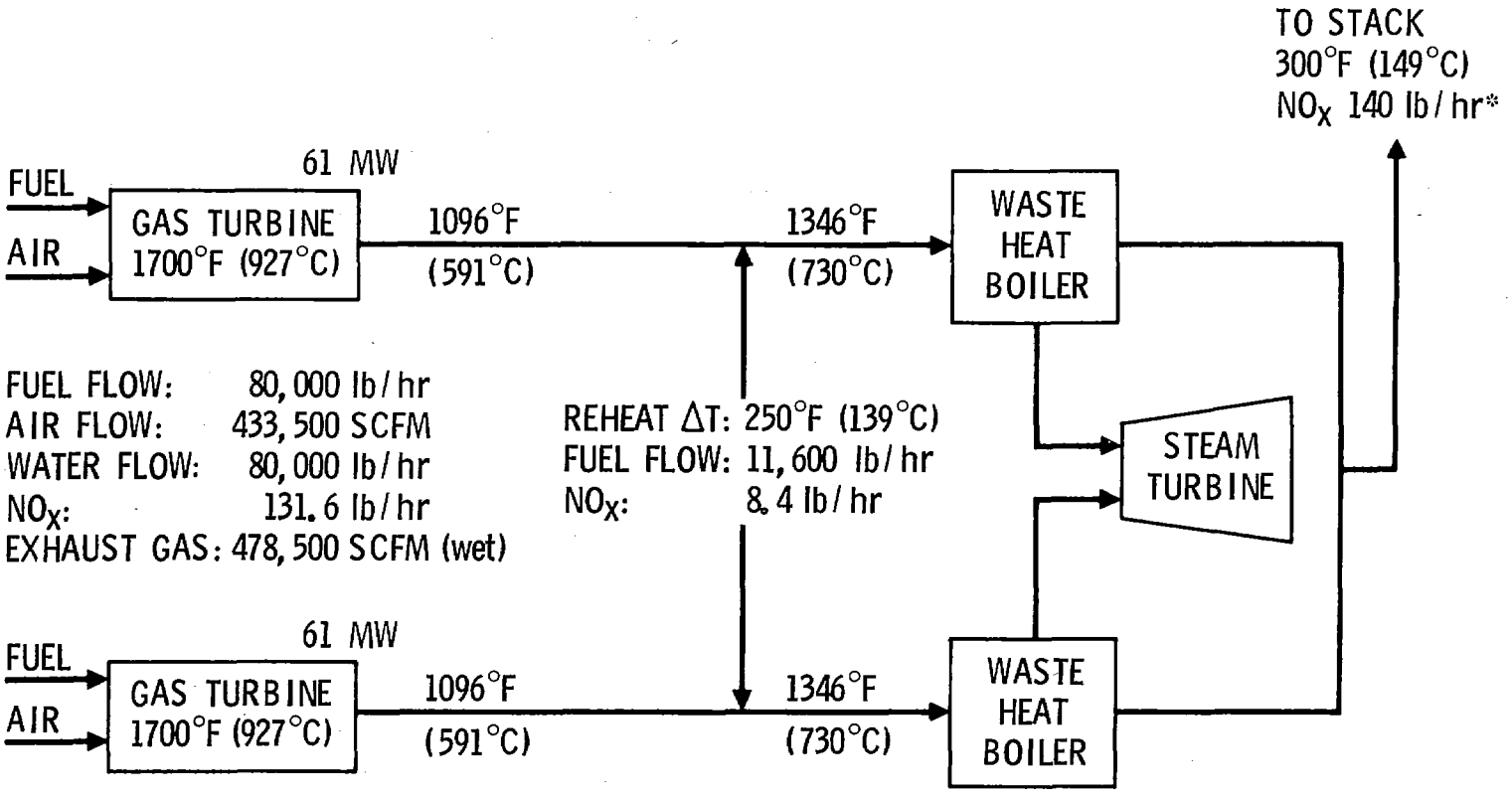
About 5800 lb/hr of fuel per unit is burned in duct heaters located between the gas turbine exit and the waste heat boiler inlet. The duct heater is a series of small burners which also use No. 2 oil (Ref. 3-21) and provide an increment of 250°F (139°C) in heating the turbine exit gas stream to provide a suitable temperature for the waste heat boiler. Figure 3-6 summarizes the unit characteristics. Other information is provided in Appendix A.

3.2.2.1 NO_x Emissions

Measured emission data were not available when the information was requested from SCE since the plant was completed in August 1978 and had not secured all permits. However, SCE indicated that the plant was designed to meet the requirements of SCAQMD Rule 67, i.e., 140 lb/hr NO_x (as NO_2). The only NO_x control technology being used at this time is water injection into the gas turbines. Figure 3-6 is a flow sheet of unit 3 or 4 and summarizes the system volumetric flow rates and emission concentrations.

3.2.2.2 NO_x Control Alternatives

Cool Water units 3 and 4 are currently outfitted with water injection systems for NO_x reduction in order to meet the 140 lb/hr emission requirement. At the injection ratio of one pound water per pound of fuel, approximately 80 to 85% reduction of the NO_x generated in the turbine results in the 140 lb/hr emission limit (Figures 3-2 and 3-3) (Refs. 3-15 and 3-20). Control alternatives considered for the 90% reduction of NO_x emissions from combined cycle turbines include thermal and selective catalytic $DeNO_x$.



	ACFM	SCFM	ppm NO_x 3% O_2	NO_x 14% O_2
WET	810,500	524,300	35	52
DRY	369,500	239,000	76	112

Figure 3-6. Combined cycle turbine system (unit 3, Cool Water, 236 MW design rating)

Combustion Modifications

Combustion modifications are not applicable as previously discussed for simple turbines in Section 3.2.1.2.

Thermal Denitrification

The turbine exit temperature is approximately 1096°F (591°C) and is at the low end of the temperature window for which the application of the thermal $DeNO_x$ process can be considered, ie., 1652°F (900°C). This temperature occurs just upstream of, or within, the turbine stage. There is no information available on the reduction in NO_x that may be attained by modifying the turbine and injecting ammonia at that location. Furthermore, there are unknowns such as whether the residence time is long enough for ammonia to react with the NO_x. Finally, based on experience in other applications, reduction of 50% or less could be anticipated, thereby being considerably less than 90% reduction criterion.

An alternative method to reach the temperature window for thermal $DeNO_x$ involves reheating the turbine exhaust gas an additional 306°F (170°C), which would require 14,200 lb/hr of fuel in addition to the 11,600 lb/hr consumed by the heaters. This would also require the injection of hydrogen in a 1 to 2 mole ratio relative to the NH_3. Since 11,600 lb/hr are consumed by the duct heaters to heat the exhaust gas for waste heat recovery, additional heating of the flue gas stream requiring 14,200 lb/hr fuel would add approximately 10 lb/hr of NO_x (as NO_2), effectively reducing the estimated 50% reduction to the system as well as significantly reducing the thermal efficiency of the system. Consequently, this approach is not considered practical and was not evaluated further.

Selective Catalytic Reduction

If SCR were applied downstream of the turbine but upstream of the duct heaters (Figure 3-7) and considering a 90% NO_x removal in the reactor, then the overall NO_x reduction for the system relative to current emissions is 76% because of the estimated 19.9 lb/hr of NO_x produced in the burner, which combined with 13.2 lb/hr exiting the catalytic reactors would result in a

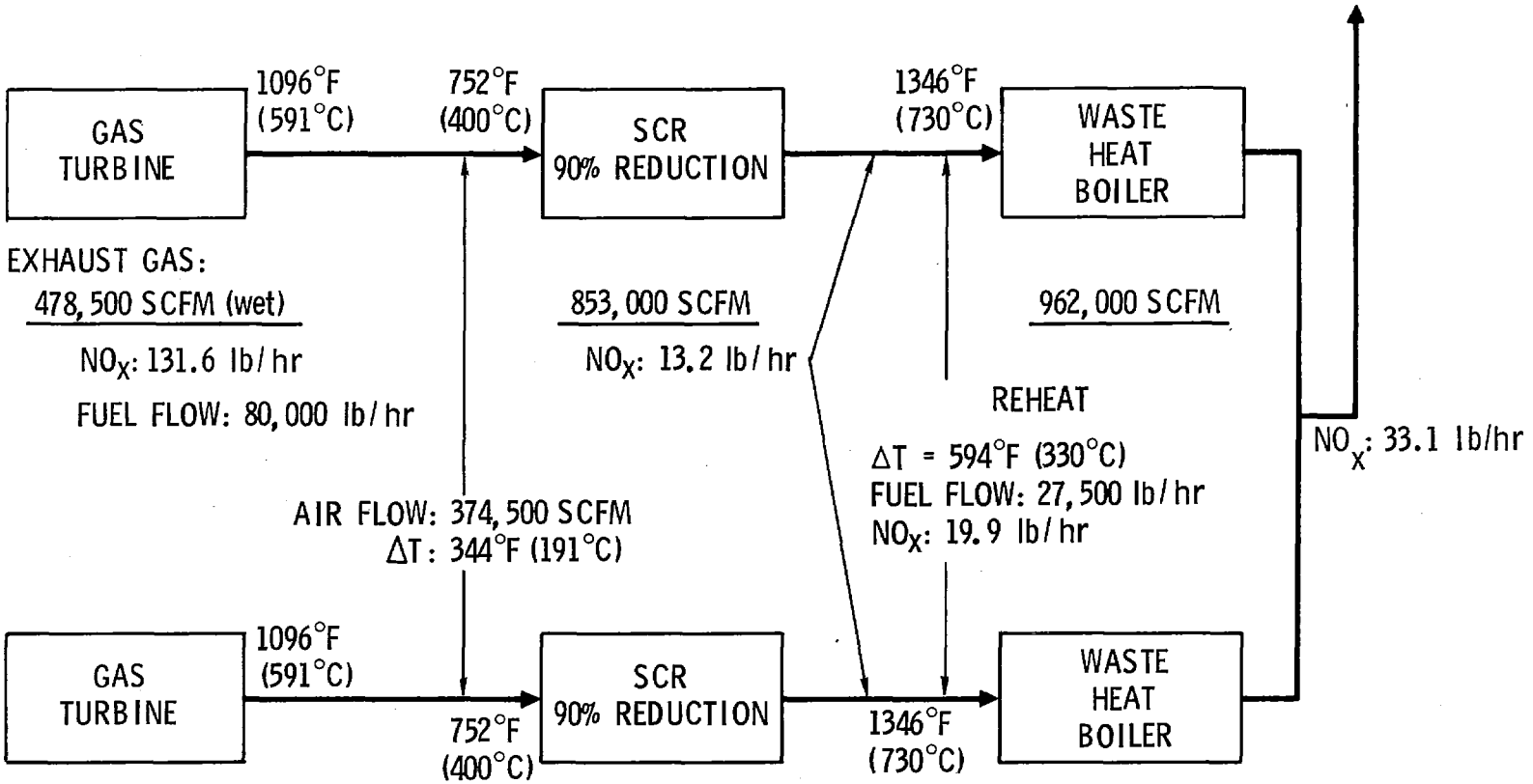


Figure 3-7. Catalytic reactor located between gas turbine and waste heat boiler

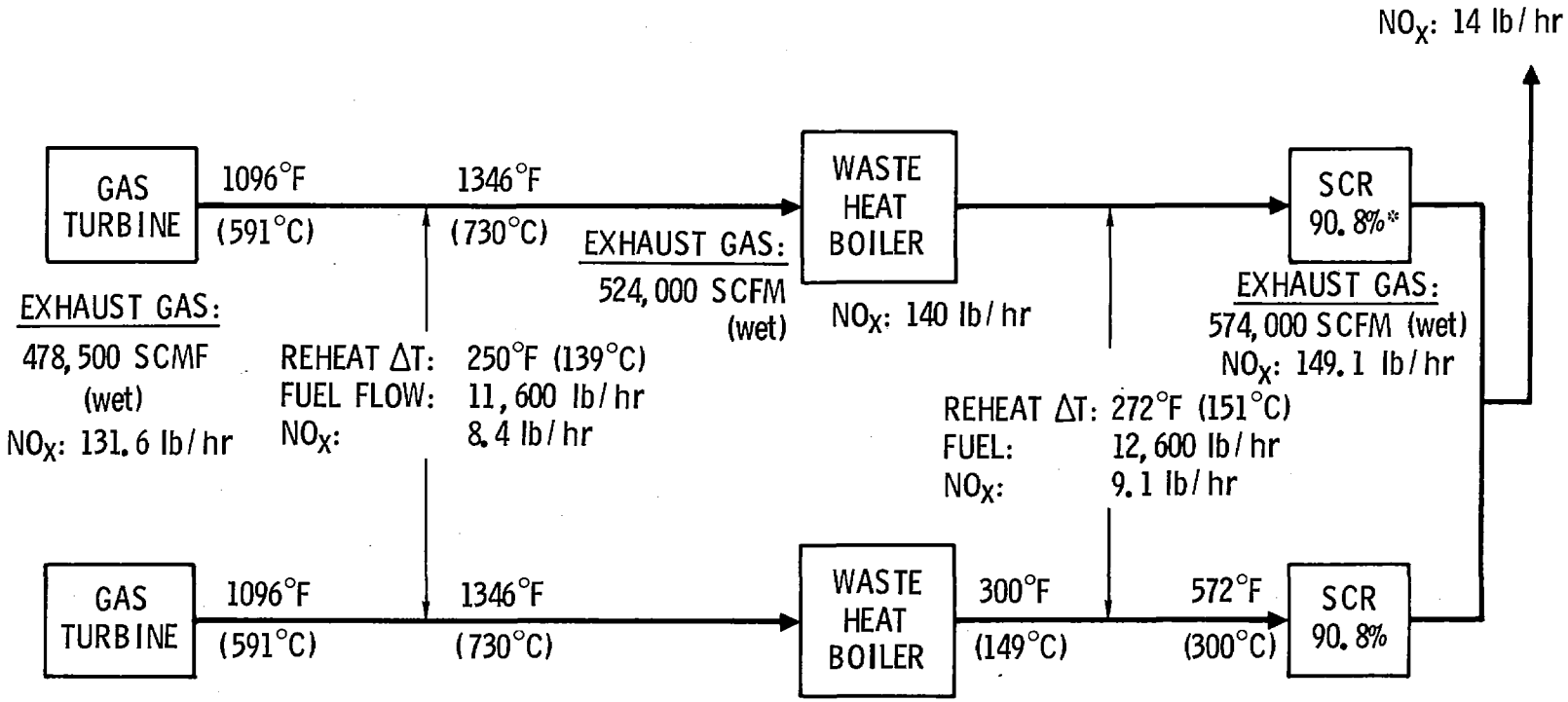
total emission rate of 33.1 lb/hr. Consequently, an overall 90% reduction, or NO_x emissions of 14 lb/hr, relative to the uncontrolled condition is impossible because NO_x removal greater than 100% would be required with the reactor in this location. Therefore, because of the impracticality of such a configuration, locating the reactor downstream of the waste heat boiler was then evaluated.

With the SCR reactor located downstream of the waste heat boiler, 272°F (151°C) of reheat would be required to achieve a minimum reactor bed temperature of 300°C. This would thereby necessitate the use of 12,600 lb/hr of fuel in addition to the 11,600 lb/hr normally consumed in the heater located upstream of the waste heat boiler (Figure 3-8). This results in a 91% NO_x removal requirement on the SCR system. Although there appears to be sufficient room in the vicinity of Cool Water units 3 and 4 for an SCR reactor and ducting between the exit of the waste heat boiler and stack, the application of SCR with catalysts currently being used with utility boilers in Japan achieving 90% reductions will be costly because of the high reheat penalty (108% of the waste heat boiler requirement) coupled with the greater NO_x load on the SCR system (Section 4.2.3).

Mitsubishi Heavy Industries is developing a catalyst for control of NO_x emissions from turbine combined cycle applications capable of operating at 650°C (1200°F) (Ref. 3-15). This will enhance the prospects of locating the catalytic reactor between the gas turbine and waste heat boiler without incurring the cooling and reheating penalty illustrated in Figure 3-8. Although specific data were not available, it was reported that endurance testing of catalysts has been initiated.

3.3 INTERNAL COMBUSTION ENGINE

A 2000 HP two-cycle, turbocharged, spark-ignited IC engine that burns natural gas and is operated by the Southern California Gas Company was examined for potential applications of $DeNO_x$ technology. The particular engine, a Cooper-Bessemer GMVH-10, was placed in service in 1965 and is one of nine engines at the Playa del Rey Station used for pumping natural gas into underground storage reservoirs. The engines are typically in service 25% of



*Reduction required to achieve 14 lb/hr emissions

Figure 3-8. Catalytic reactor located downstream of waste heat boiler

the year (approximately 1865 hr/yr), usually during the summer when the demand for natural gas is relatively low. Because of this infrequent use, the engines are projected to have a lifetime of between 50 and 100 years.

The maximum combustion air inlet manifold temperature is 98°F (87°C), and the maximum heat rate is approximately 6800 Btu/Bhp at 80°F. The air-fuel ratio is reported at 30 to 1, with a 12,580 CFH fuel consumption rate. These and other unit characteristics are summarized in Appendix A.

A typical analysis of the fuel (natural gas) which is burned in the engine is given in Figure 3-9. Southern California Gas reported that the fuel as burned has a 953 to 1080 Btu/SCF calorific value.

One stack is provided for each engine, and the 35,000 lb/hr of exhaust gases exit at 650°F (343°C). Excess O_2 at the stack was computed to be 18%. Gas flow rates are indicated in the flow diagram (Figure 3-9), along with system temperatures.

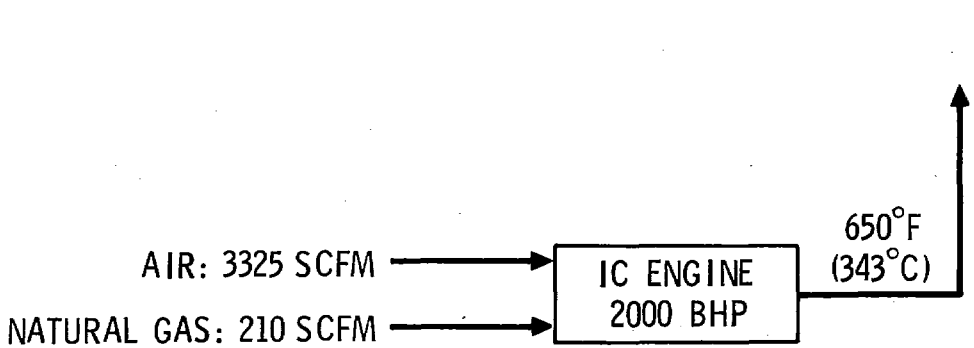
Excess O_2 in the combustion chambers and at the stack differ due to the nature of two-cycle gas engines. In most cases, the fuel is injected directly into the cylinder (premixed with air), and combustion and scavenging air enter through ports in the cylinder wall, which are uncovered as the piston nears the bottom of its stroke. As a result of exhaust scavenging, exhaust pollutants are diluted to about 1/2 to 1/3 of their original concentrations (Ref. 3-22).

3.3.1 NO_x Emissions

The 2000 HP Cooper Bessemer GMVH-10 engine emits approximately 40.2 lb/hr of NO_x as NO_2 (Ref. 3-23) at a concentration of 3365 ppm NO_x (dry, 3% O_2). No specific NO_x control measures are currently in use. Figure 3-9 summarizes the volumetric flow rate of exhaust gases referenced to wet and dry conditions at 3% O_2 and at actual conditions.

3.3.2 NO_x Control Alternatives

The basic objectives of the study were to assess the feasibility of a 90% NO_x reduction for the 2000 HP engine. Thermal $DeNO_x$ and SCR were the NO_x control technologies examined for the IC engine although other strategies have



	WET	DRY
ACFM	15,570	5,800
SCFM	7,340	2,570
NO_X (18% O_2), ppm	713	2,037
NO_X (3% O_2), ppm	1,178	3,365

TYPICAL FUEL ANALYSIS, %

N_2	5.15
CH_4	81.11
C_2H_6	9.66
C_3H_8	3.51
C_4H_{10}	0.43
C_5	0.09
C_6	0.05
	100.00

Figure 3-9. Internal combustion engine gaseous emissions characteristics

been discussed in the literature including combustion chamber design modification, alternative fuels, exhaust gas recirculation, optimization of air-fuel ratio, modification of fuel injection timing, and use of variable compression ratios (Ref. 3-24). Basically, a 50% reduction in the NO_x emitted from the engine can be expected with thermal $DeNO_x$. With the use of SCR alone, it is expected that the NO_x can be reduced by 90%.

For this engine, the exhaust gas temperature encountered is approximately 650°F (343°C); thus, in order for thermal $DeNO_x$ to be applicable, the gases require a reheating increment of 1000°F (557°C) to reach the optimal temperature envelope, which starts at about 1650°F (900°C). Consequently, 754 lb/hr of fuel oil or approximately 222 SCFM of natural gas is necessary to achieve the required process temperature. This does not appear practical in terms of thermal efficiency since the engine only consumes 210 SCFM of natural gas. The heater would generate an estimated 2 lb/sec of NO_x, which would reduce the NO_x removal effectivity of the thermal system from 50 to 45%.

Tests with thermal $DeNO_x$ systems have indicated that the addition of hydrogen in the range of 0.5 to 2.0 mole ratio (relative to NH_3) can be used to improve NO_x conversion at temperatures below optimum. An illustration of this phenomenon is shown in Figure 3-10, based on tests in a coal-fired system. Although NO_x removal can be improved at temperatures lower than optimal, the same degree of NO_x control as at the optimal temperature condition without hydrogen is unlikely (Figure 3-10). For example, if the optimum temperature is 1700°F for a 60% reduction and the actual temperature encountered is 1600°F, then the amount of expected NO reduction without hydrogen is about 25%. However, with hydrogen injection, the NO_x removal rate at 1600°F is on the order of 50%. Consequently, use of hydrogen in the IC engine application does not improve overall NO reduction nor significantly alleviate the high reheat fuel requirement.

The other alternative, SCR, requires no reheat or additional air for dilution to reach the appropriate process temperature. A schematic of this configuration is shown in Figure 3-11. A 90% NO_x reduction is considered feasible with an SCR unit using existing base metal catalysts (technology

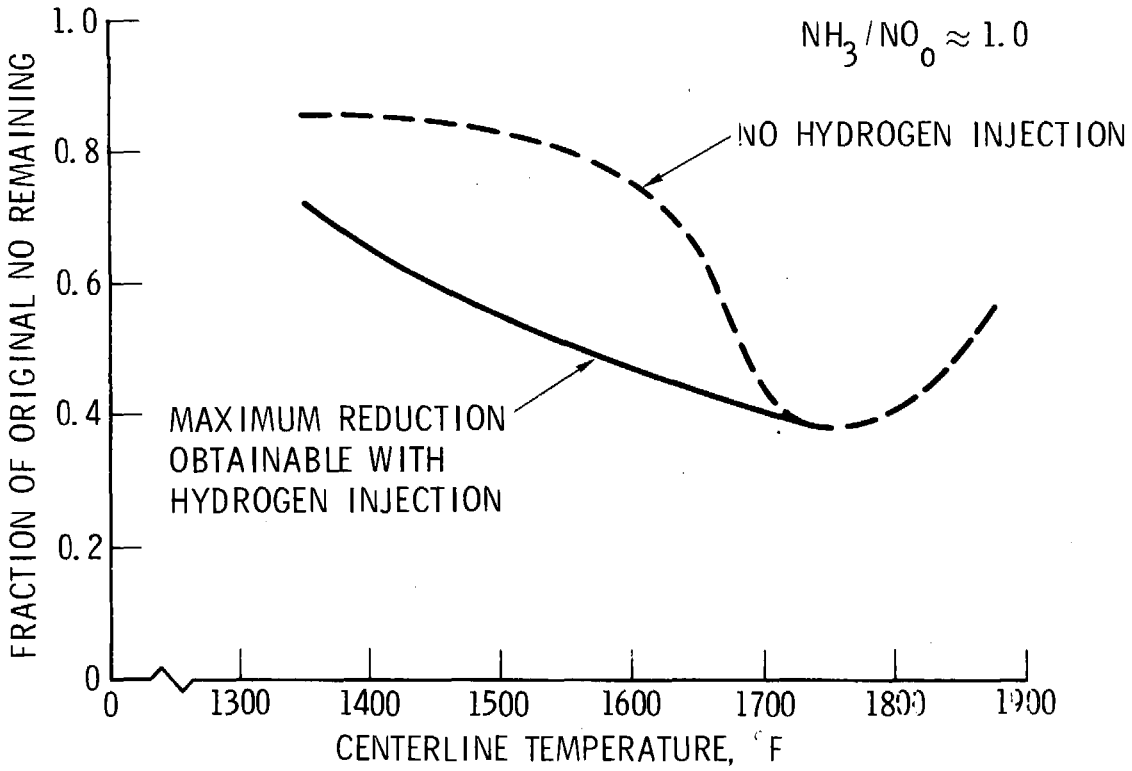
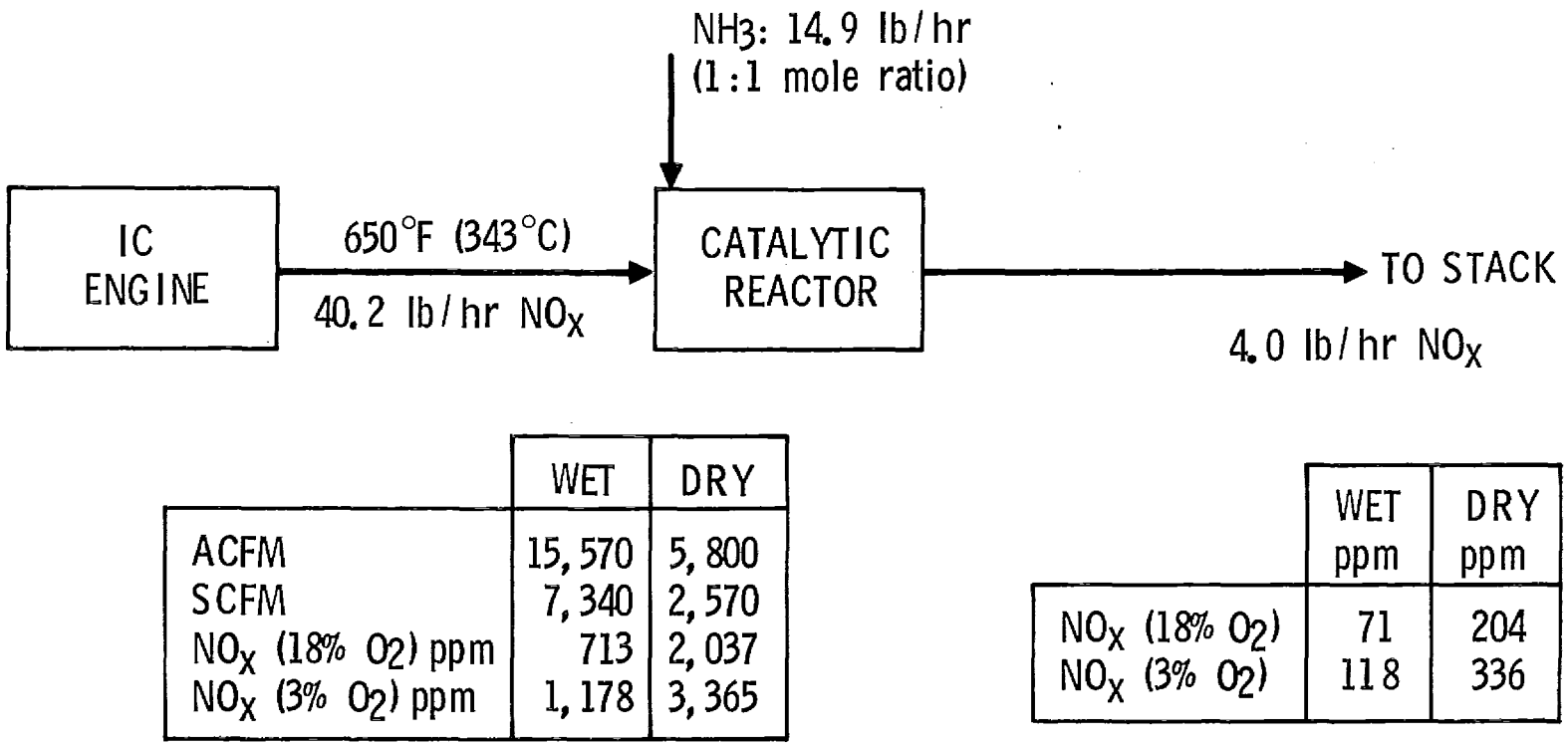


Figure 3-10. Effect of hydrogen injection on optimum NO reduction, Pittsburgh Coal (Ref. 3-26)



	WET	DRY
ACFM	15,570	5,800
SCFM	7,340	2,570
NO_X (18% O_2) ppm	713	2,037
NO_X (3% O_2) ppm	1,178	3,365

	WET ppm	DRY ppm
NO_X (18% O_2)	71	204
NO_X (3% O_2)	118	336

Figure 3-11. Selective catalytic reduction schematic for internal combustion engine

discussed in Section 2.4.4). At a one to one mole ratio of NH_3 to NO_x, 14.9 lb of ammonia per hour are required. Thus, the resulting emissions would be approximately 4.0 lb NO_x (as NO_2) per hour or 336 ppm (dry, 3% O_2). Because of the relatively clean nature of the exhaust gas, a 15,000 hr^{-1} space velocity is estimated as capable of removing 90%. The corresponding dimensions of the catalyst bed are 3.1 × 3.1 × 3.1, with a superficial velocity of 12.7 ft/sec (3.9 m/sec).

3.4 OIL FIELD STEAM GENERATOR

A 50 MM Btu/hr oil field steam generator typical of those operated in Western Kern County was selected for this study. Steam generators of this size are used extensively in oil fields in California for thermally enhanced oil recovery (TEOR). In this process, the steam which is generated is injected into wells to stimulate flow of heavy subsurface oil, thereby enhancing its recovery. Existing units are operated singly or in multiple units with stacks manifolded. In some cases, the flue gas is routed through an SO_2 scrubber.

3.4.1 Steam Generator Characteristics

The generators are single-burner horizontally fired units. They are water-tube package boilers, with their ratings based on the amount of heat absorbed. The 50 MM Btu/hr generator (62.5 MM Btu/hr input) is capable of producing 80% quality steam and requires infrequent maintenance or adjustment. Because the units are relatively unattended in their operation, standard practice is to adjust the generators to excess oxygen levels in the range of 4 to 5%. For standard burners, 1 to 2% excess oxygen is generally considered to be a practical minimum to avoid smoking and excess carbon monoxide emissions resulting from incomplete combustion. Additional operator surveillance would be required if low excess air operation were implemented as a NO_x control measure.

Recovered heavy crude oil (API gravity 11 to 14) is generally used as fuel although topped crude is sometimes used. Maximum crude oil consumption

is about 390 GPH* (6.5 GPM) or 223 bbl/day of crude. For relatively unattended operation, they are operated at 80% of maximum capacity (Ref. 3-27), consuming approximately 5.2 GPM (178 bbl/day).

A range of typical crude oil properties is given in Table 3-14. Nitrogen content of the crude oil may be as low as 0.4%, but is generally in the range of 0.6 to 0.8%. Sulfur content ranges from 1.0 to 1.5% and ash from 0.03 to 0.3%.

Many of the oil field steam generators in California are manufactured by Thermotics, Inc., Struthers Thermo-Flood, or CE-Natco (Combustion Engineering). All are generally equipped with North American Manufacturing Company conventional burner assemblies. Characteristics of a representative unit are shown in Figure 3-12.

Combustion air enters the burner assembly at ambient temperature. The flue gas temperature leaving the radiant section of the boiler is in excess of 1550°F (844°C). Subsequently, the flue gas enters the economizer where the feedwater is preheated. The stack exit temperature is typically about 450°F (232°C).

3.4.2 Steam Generator Emissions

NO_x emissions of 270 ppm (dry, 3% O_2), or 15.4 lb/hr (as NO_2), have been reported for crudes with approximately 0.7% fuel nitrogen and operating at 46 MM Btu/hr input (5.1 GPM) (Ref. 3-27). Based on the reported value of 15.4 lb/hr, it results in a computed concentration of 295 ppm (dry, 3% O_2). A baseline value of 310 ppm is reported by Standard Oil of California for one of their Struthers Wells units (Ref. 3-28). Under similar conditions with a low excess oxygen burner operating with 1% oxygen a NO_x concentration of approximately 415 ppm (dry, 3% O_2, as NO_2) was measured; heat input was not identified (Ref. 3-26). The effect of operating conditions on NO_x emissions are illustrated in Figures 3-13 and 3-14.

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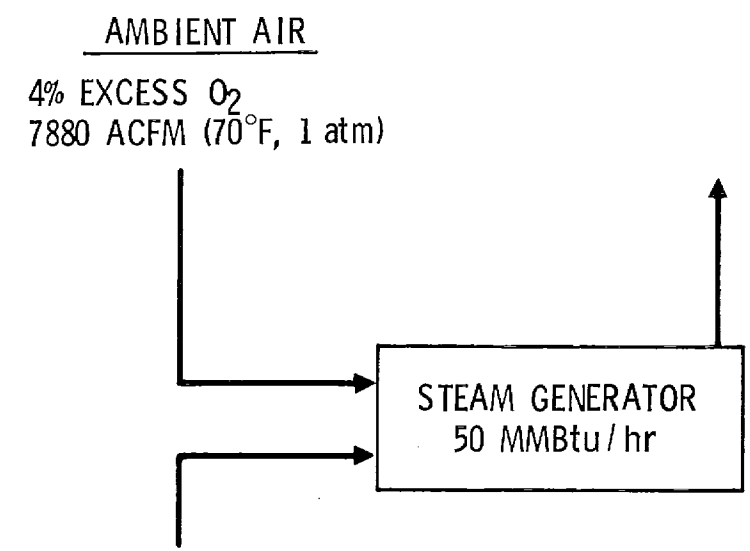
*Reference 3-26 indicates 425 gallons per hour (GPH) with 130,000 Btu/gal fuel oil. Crude oil ranges from 145,000 to 151,000 Btu/gal.

TABLE 3-14. TYPICAL CRUDE OIL ANALYSES (WEST KERN COUNTY)

CONSTITUENT	RANGE
Carbon	85.0 - 86.5%
Hydrogen	11.0%
Nitrogen	0.4 - 0.8%
Sulfur	1.0 - 1.5%
Ash	0.03 - 0.3%
API Gravity	11 - 12
Heat Content	17,700 - 18,400 Btu/lb



FLUE GAS: 450°F (232°C)

H_2O: 9.5% SO_2: 58 lb/hr = 704 ppm*
CO_2: 12.5% NO_x: 15.4 lb/hr (as NO_2)
CO: 50 ppm SO_3: 20 ppm
PARTICULATES: 0.064 GRAINS/SCF (147 mG/Nm^3)**

			NO_x, ppm		
FLOW RATE	WET	DRY	WET	DRY	DRY, 3% O_2
ACFM SCFM[a]	14,450 7,810	13,080 7,070	256[b]	283[b]	295[c]

[a] 32°F, 1 atm, standard burner
[b] 4% O_2
[c] Computed as 295 ppm based on 15.4 lb/hr (Ref. 3-26). Value is reported as 270 ppm in Ref. 3-26.

*ACFM, wet basis
**Wet basis

Figure 3-12. The 50 MM Btu/hr system generator characteristics (Refs. 3-26 and 3-27)

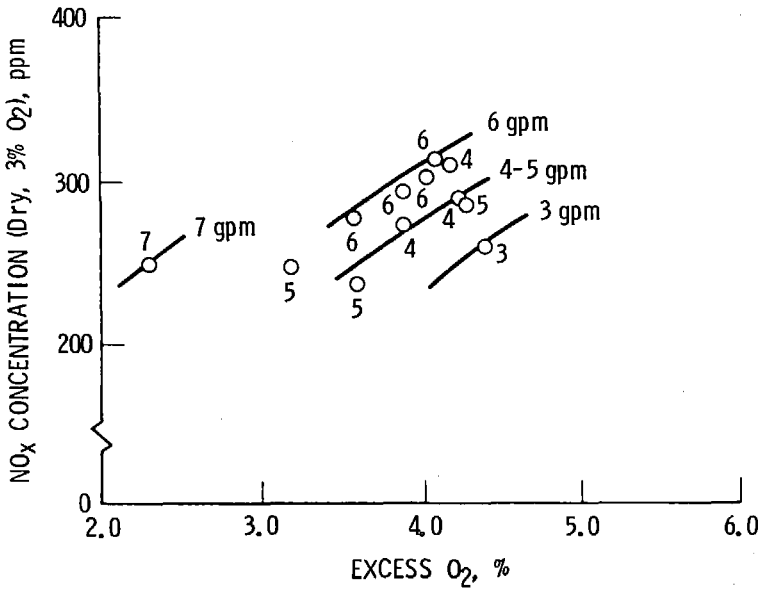


Figure 3-13. NO_x emissions with standard burner (Ref. 3-26)

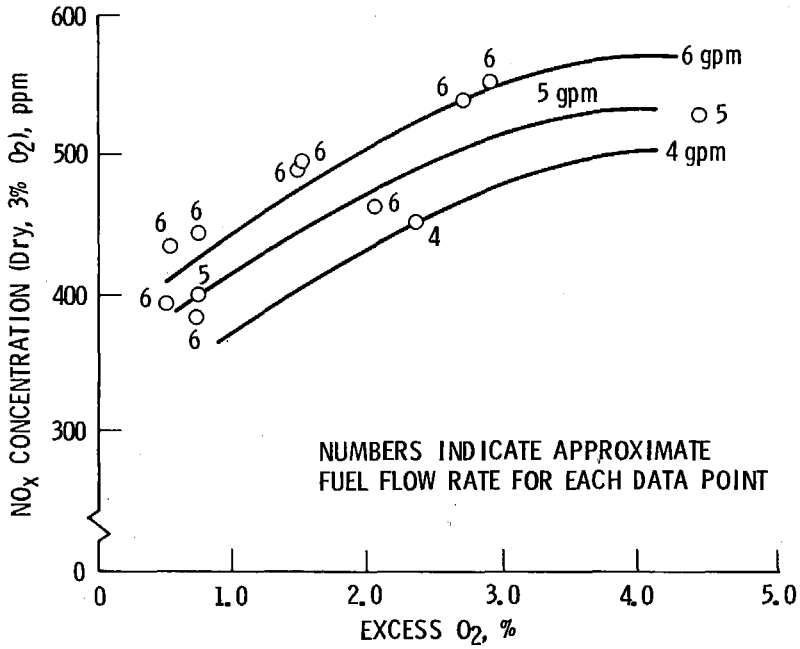


Figure 3-14. NO_x emissions for low oxygen burner (Ref. 3-26)

Typically, SO_2 emissions are about 700 ppm (1.1% in the fuel). SO_3 levels in the flue gas were reported as 20 ppm (Ref. 3-27). It is basically a H_2SO_4 mist with particles in the size range of 1 to 5 microns.

Particulate content is reported as 4.39 lb/hr (Ref. 3-27) (based on only the ash content of the crude, the particulates were computed as 1.76 lb/hr). Using the 4.39 lb/hr value, this is equivalent to 0.064 grains, or 147 mg/Nm^3. On the basis of the SO_2 and particulate concentrations of 700 ppm and 0.064 grains/SCF (147 mg/Nm^3), the flue gas is considered relatively dirty for catalytic reactor applications, and this will influence the catalyst and reactor configurations (Section 3.4.2.3).

The presence of moisture and SO_3 together with the unreacted NH_3 from the NO_x control measures form ammonium bisulfate (NH_4HSO_4) (Figure 3-15).

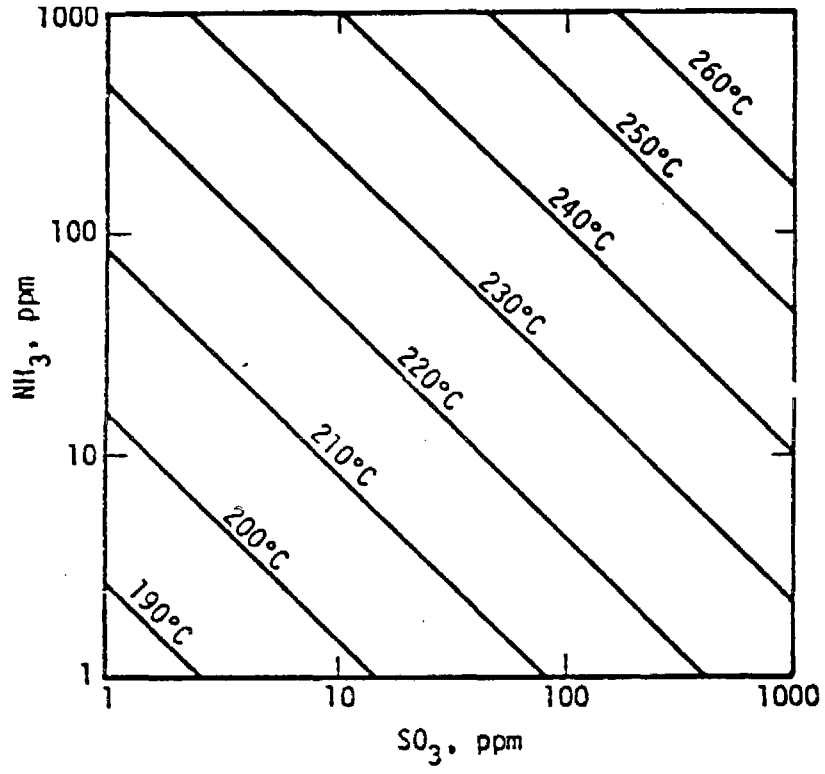


Figure 3-15. Formation temperatures of liquid ammonium bisulfate from gaseous reactants, NH_3, SO_3, and H_2O (Ref. 3-11)

TABLE 3-15. EMISSION CONTROL REQUIREMENTS

NO_x

Steam Generator Operating Level,[a] Fuel Flow Rate, Gpm	NO_x Emissions[b] (Uncontrolled)				Reduced Emissions[f] (Controlled)			
					New Units		Existing Units	
	lb/hr	lb/bbl	lb/10^6 Btu Heat Input	ppm[c]	Red., %	lb/hr[b]	Red., %	lb/hr[b]
5.1	15.4	2.1	0.35	295[d]	66.1	5.2	49.1	7.8
6.5	22.2	2.4	0.39	335[e]	70.1	6.6	55.2	9.9

SO_2

Steam Generator Operating Level, Gpm	SO_2 Emissions[g] (Uncontrolled)			Reduced Emissions (Controlled)			
	lb/hr	lb/10^6 Btu	ppm	lb/10^6 Btu	lb/hr	Red., %	ppm
5.1	63.0	1.36	885	0.12	5.6	91.2	79
6.5	80.3	1.36	885	0.12	7.1	91.2	79

[a] 4% excess O_2

[b] As NO_2

[c] Dry, 3% O_2

[d] Reported as 270 ppm in Ref. 3-27. Corrected to 295 ppm to be consistent with operating conditions and reported value of 15.4 lb/hr of NO_x

[e] Estimated as 335 ppm based on data in Ref. 3-27. Computed as 22.2 lb/hr to be consistent with 335 ppm and operating conditions.

[f] To achieve:

	NO_x, ppm
New	100
Existing	150

[g] 1.25% sulfur fuel

Under certain conditions, it may condense on the economizer tubes. SO_3 is also formed in the SO_2 scrubber and may generate additional complexities relative to post combustion NO_x control systems. The consequences of NH_4HSO_4 formation on ammonia-based NO_x control systems will be discussed in the thermal $DeNO_x$ and SCR paragraphs (Sections 3.4.3.2 and 3.4.3.3).

3.4.3 Emission Control Alternatives

Combustion modifications, thermal $DeNO_x$, and SCR to achieve NO_x control are discussed below. Crude oil burned in the steam generators contains appreciable amounts of sulfur which require control measures also. The use of sulfur control measures is expected to impact the application of thermal $DeNO_x$ and SCR in terms of affecting exhaust gas characteristics.

The NO_x control criteria to be evaluated included a limit of 100 ppm (dry, 3% O_2) for new units and 150 ppm for existing units. The percent reductions are approximately 70 and 55 percent, respectively, and are a function of the generator operating conditions (Table 3-15).

On September 26, 1979, the Air Resources Board adopted Rule 424 requiring that, commencing January 1, 1985, SO_2 emissions from existing generators be limited to 0.12 lb sulfur/MM Btu heat input. Operating at 6.5 GPM, SO_2 emissions for 1.25% sulfur oil are 80.3 lb/hr, or 1.36 lb/10^6 Btu (885 ppm, dry, 3% O_2). This requires a 91.2% SO_2 reduction (Table 3-15). Controlling the sulfur content of the fuel as a means of SO_2 reduction would require burning oil with approximately 0.11% sulfur. This is outside the realm of California crude oil sulfur levels. This necessitates post combustion SO_2 control, which can be achieved by wet scrubbing. No impacts are expected with the use of SO_2 scrubbers in combination with CM. However, some are anticipated with thermal $DeNO_x$ and SCR. These are addressed in Sections 3.4.3.2 and 3.4.3.3.

3.4.3.1 Combustion Modifications

Because of the relatively high nitrogen content of the crude oil, the fuel-bound nitrogen as well as thermal nitrogen contribute to the NO_x formed. For fuel with 0.7 to 0.8% fuel nitrogen, that source accounts for approximately three-fourths of the NO_x emitted with the remainder as a result of the thermal fixation of atmospheric nitrogen (Ref. 3-29). Therefore, achievement

of significant NO_x reduction in steam generators by combustion process modifications must incorporate methods that will reduce the conversion of fuel-bound nitrogen to NO.

Combustion modifications to reduce thermal NO_x are based on reducing (1) peak temperatures during combustion, (2) residence time of the combustion gases in high temperature zones, and (3) reducing the availability of oxygen in the primary, high-temperature combustion regions. Fuel NO_x can be reduced by decreasing the availability of oxygen in the primary combustion zones.

Low Excess Oxygen

Estimates have been made indicating a 15% reduction in NO_x emissions from the 300 ppm level by reducing excess oxygen from 4 to 2% (Figure 3-16) (Ref. 3-30). A reduction from 4 to 3% with a standard burner in a steam generator exhibited a drop in emissions from approximately 300 to 240 ppm (20%) (Figure 3-13).

With a low excess oxygen burner, at 6 GPM, a 2 percentage point reduction in excess oxygen (from 3 to 1%) resulted in a 20% reduction (from 550 to 440 ppm) (Figure 3-14).

Field test results published by Chevron Research Company (Ref. 3-31) are summarized in Table 3-16 and Figure 3-17. In these tests, baseline data were obtained from a single 50 MM Btu/hr Struthers Wells oil field steam generator equipped with a conventional North American Burner assembly (Model 6131-G) firing 30-D LACT crude (0.7% nitrogen) at 3.5% O_2. NO_x emissions and thermal efficiency were measured at the baseline uner operating conditions of varying O_2 levels and at approximately 93% load (see Table 3-16). Results are dependent on fuel characteristics, unit condition, and operating conditions; thus, data and conclusions reported by Chevron should be considered within the constraints of the test conditions since tests were performed on a single unit, at one load, and with one type of fuel.

The data presented in Table 3-16 indicate that an approximate 20% reduction in NO_x emissions can be realized with a conventional burner if operated at 1.5% O_2. This level is generally the minimum practical and safe operating point since smoking occurs at 0.7 to 0.8% O_2. North American Manufacturing Company (Ref. 3-32) has indicated that the only observed effect of

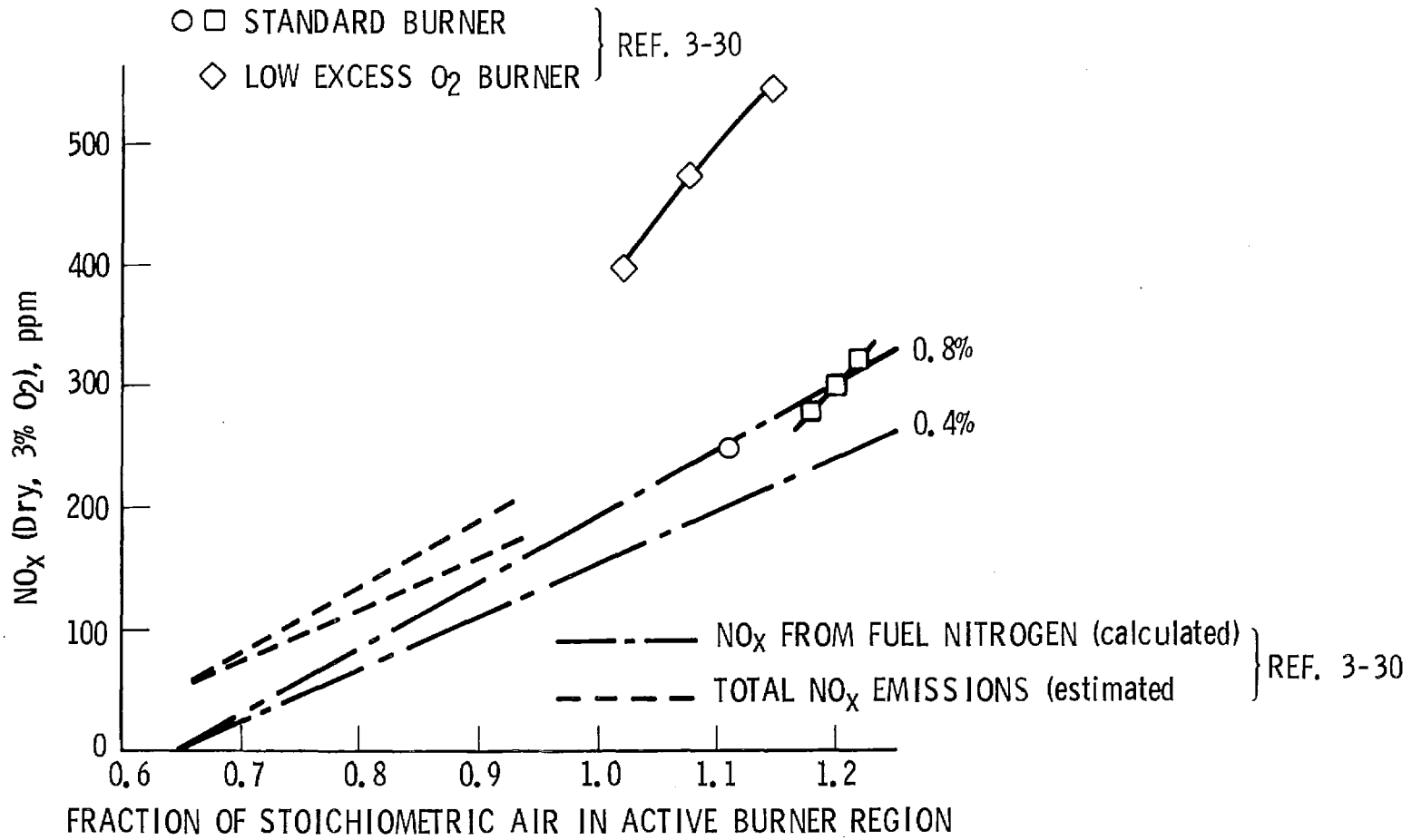


Figure 3-16. NO_x emissions from 50 MM Btu/hr thermally enhanced oil recovery steam generator

TABLE 3-16. NO_x EMISSION AND EFFICIENCY DATA (Ref. 3-31)

O_2 Level, %	TRW Low NO_x Burner	Original Burner[2]
	NO_x Emissions (ppm)	
3.5[3]	-	315
3.0	280	305
1.5[4]	230	250
	Thermal Efficiency(%)	
3.5	83	82
2.0-4.0	-	82-86[5]
1.5	86	83

[1]Operating at ~93% load, with 0.7% nitrogen crude oil. Efficiencies determined after water washing of convection section.

[2]Manufactured by North American Company.

[3]Baseline condition.

[4]Minimum O_2 level for safe operation.

[5]Range of field tests on existing steam generators.

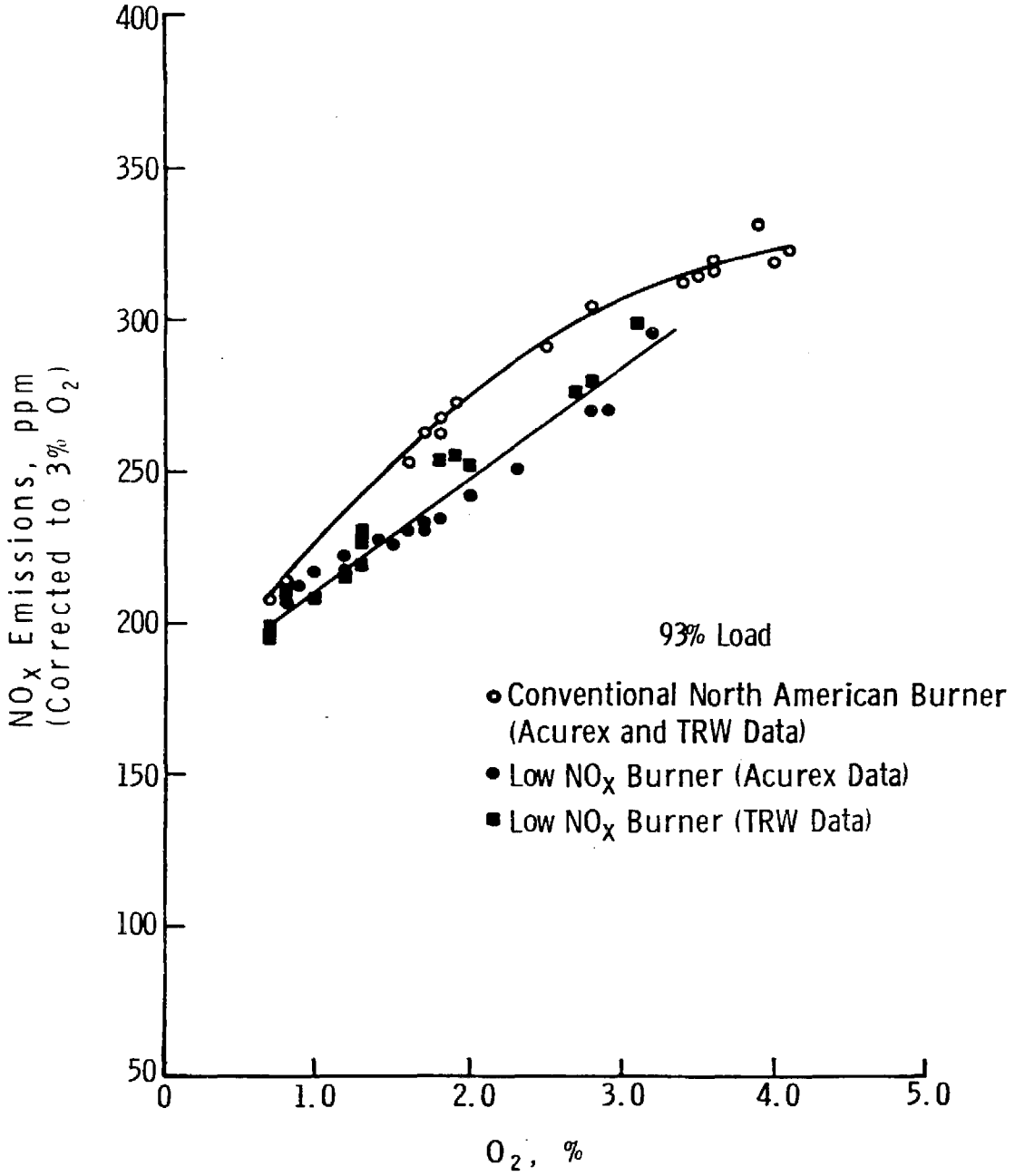


Figure 3-17. NO_x emissions: burner comparison (Ref. 3-31)

operating at reduced excess air above the smoke point is a slightly elongated and larger flame pattern.

Figure 3-17 shows NO_x emissions plotted against O_2 for the conventional North American burner and for the TRW LNB. Data were obtained from both TRW and Acurex participation in the Chevron study.

Another test was carried out where NO_x reductions were observed by decreasing fuel nitrogen levels, from 0.85 to 0.77% (~10% reduction), and by load reduction from 100 to 70% (~10% reduction).

Thus it appears that employing low excess oxygen combustion (reducing excess O_2 from 4% to approximately 2 to 3%) can achieve a reduction of 20% in NO_x emissions. However, its reliable implementation is expected to require instrumentation to determine oxygen concentration in the exhaust gas and a closed loop controller and regulators to maintain the emissions at the required levels to maximize fuel savings resulting from the reduced air volume and to ensure safe unattended operation. There is also a fuel saving that accrues as a result of using less excess air. Since less heat is required to heat the reduced air approximately 1%, fuel can be conserved for reducing excess oxygen from 4 to 2%.

Two-Stage Combustion

Although use of staged combustion has been analyzed and has a potential of yielding significant reduction (approximately 70%) in NO_x emissions (Figure 3-16) (Ref. 3-30), it has not been tested. It requires operating the first stage at 75% of stoichiometric air, with the remaining air introduced via overfire air (OFA) ports. This would require openings to be cut into the firebox and the installation of ducts from the blower to the new ports. A control system would also be required. It is expected that the flame size and shape would be affected with the potential for impingement and damage to the tubes. Other items such as combustion and flame stability, generation of particulates, and effect on boiler efficiency must be addressed prior to giving serious consideration to this approach for retrofits. However, incorporation into new steam generator designs may be a means for implementation of this method.

Mobil Oil has incorporated staged combustion into one of its 25 MM Btu/hr oil field steam generators located in Kern County. The unit burned 0.42% nitrogen fuel, and NO_x emissions at the 155 ppm level (2.5% O_2) have been achieved (Ref. 3-33).

Flue Gas Recirculation

Flue gas recirculation affects flame patterns creating potential impingement problems within the radiant section. More significantly, it reduces primarily only the thermal NO_x. Since thermal NO_x is approximately one fourth of the total emitted, NO_x reduction by FGR would be limited. The installation would require a blower, damper, ductwork, and controls and appears complex for its rather limited reduction potential.

Currently, FGR controls are utilized in two Mobil Oil 25 MM Btu/hr steam generators in Kern County. One generator which burned 0.42% nitrogen oil achieved NO_x emissions of 207 ppm at 2.3% O_2, with a recirculation ratio of 0.275 recycle/air (Ref. 3-33). The other generator had NO_x emissions at the 200 ppm level (1.5% O_2), using a recirculation ratio of 0.155 recycle/air (Ref. 3-33).

Low NO_x Burners

Reports of testing of LNB in steam generators or test furnaces simulating the units have been published (Refs. 3-28, 3-29, and 3-34). In addition, numerous burner manufacturers have commercialized or are in the process of developing LNB for other applications for use with oil and embodying various combustion modifying techniques to reduce NO_x emissions (Section 2.4.2).

As was noted earlier, the crude oil contains a high level of fuel-bound nitrogen which, when converted to NO in the combustion process, represents over three-fourths (>225 ppm) of the NO_x produced. Therefore, the LNB that would be more effective are those incorporating techniques that reduce fuel nitrogen conversion i.e., staged combustion types such as those using OSC and TSC. Since the major portion of liquid fuel burned in Japan

contains a low nitrogen content (approximately 0.35% or less), many of the burners are designed to reduce thermal NO_x and do not employ a staged combustion approach (Sections 2.4.2.1 through 2.4.2.3). Thus, even with a high thermal NO_x reduction, these types of burners would not be effective in reducing NO_x levels below 225 ppm (dry, 3% O_2) when operated with 4% excess oxygen. Therefore, these will not be discussed in this section. However, those that appear to have application in reducing fuel NO_x include those staged combustion types developed by Chugai Kogyo, MHI, Kawasaki, and others (Section 2.4.2.4) and combination types of Nippon Furnace Kogyo Kaisha, Ltd. (NFK), Ishikawajima-Harima Heavy Industries, Co., Ltd. (IHI) and Hitachi-Zosen (Section 2.4.2.5). However, all the data available are for gaseous fuels with low fuel-bound nitrogen contents. Any projected performance of these units would require considerable extrapolation.

In the United States, LNB tests reported by John Zink, CEA, Chevron, and small experimental burners tested by Energy and Environmental Research Incorporated (EER) use staged combustion techniques to control NO_x formation and have been discussed as potentially applicable to TEOR steam generators (Refs. 3-29, 3-31, 3-34, and 3-35).

In Reference 3-35, two LNB are identified. For fuel containing 0.7% nitrogen, a 12% reduction relative to its standard burner in overall NO_x emissions is reported for one of their LNB types at 4% excess air. In tests with their HIV model burning crude (0.7% N, 1.1% S), NO_x emissions of approximately 230 ppm (or 250 ppm adjusted to 0.8% N) were reported for operation at 4% excess oxygen (Ref. 3-28). Relative to burners currently installed in steam generators operating at 60 MM Btu/hr (approximately 6.5 GPM fuel consumption) and emitting 335 ppm (for 0.8% N), this represents a 30% reduction (Table 3-17).

The effect of reduced O_2 on NO_x emissions from an HIV burner is shown in Figure 3-18. A lowering of excess oxygen from 4 to 2% represents an additional 6% NO_x reduction (relative to 335 ppm), down to 230 ppm. However, it is noted that the potential emissions by extrapolating the operation of the standard burner at 2% oxygen are 205 ppm.

TABLE 3-17. ESTIMATE OF REDUCTION IN NO_x EMISSIONS WITH LOW NO_x BURNERS

Burner	% Reduction at 4% Excess O_2 [a]	Overall* Reduction at 2% Excess O_2
John Zink (HIV) (60×10^6) Btu/hr)	25% (250 ppm)	31% (230 ppm)
CEA (Axiflow) (10×10^6 Btu/hr)	Not Available (0%) [b]	Not Available (43%, 190ppm)[b]
EER (Test Burner) (40×10^6 Btu/hr)	Not Available	Not Available[c]
Standard Burner[a]	0% (335 ppm)	39% (205 ppm)[d]

[a]For 0.8 %N Fuel, 335 ppm Baseline, Dry, 3% O_2.

[b]Extrapolated from Information Provided in Figure 3-18.

[c]150 ppm at 0% O_2 (55%).

[d]Extrapolated from 3% O_2.

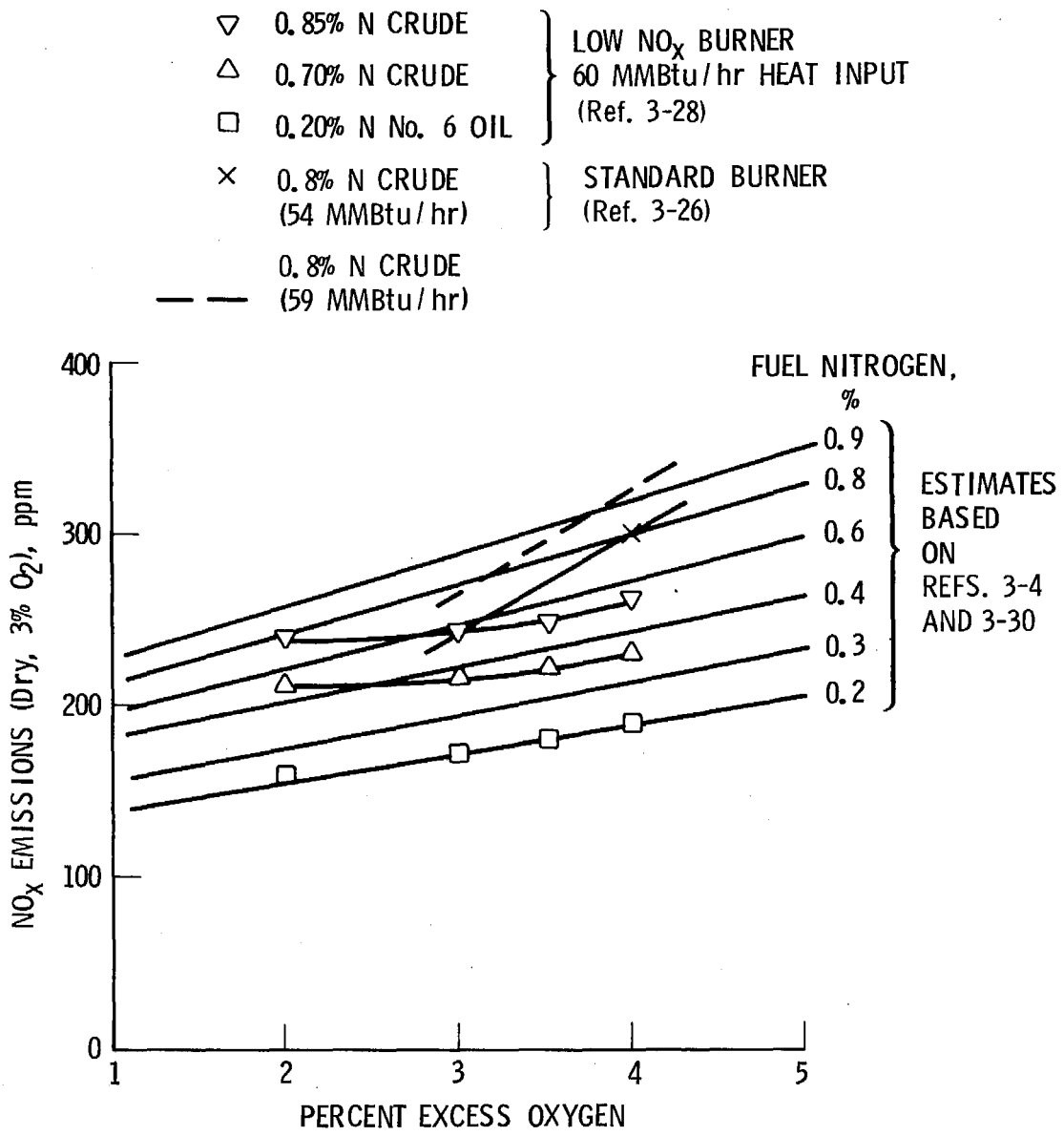


Figure 3-18. NO_x emissions from low NO_x burner

Therefore, if one is considering operating at 4% excess oxygen, the use of the HIV burner has the potential of reducing NO_x, approximately 30% relative to the standard burner. However, if operation with reduced excess oxygen, i.e., 2% is considered, then there does not appear to be a clear advantage (emissions wise) in using this LNB.

Another burner, with data reported by CEA (Ref. 3-34), displays the characteristics shown in Figure 3-19. The results are for a 10 MM Btu/hr unit using 0.32% nitrogen fuel, firing at 1 to 3% excess oxygen. However, data were not reported for operation at 4% with 0.32% nitrogen fuel nor with fuel containing 0.7 to 0.8% nitrogen at any condition. The effectiveness of this burner is assessable based on data currently available, because it appears to parallel the NO_x reduction effect of low excess oxygen exhibited by the standard burner. Thus 2% excess air with 0.8% nitrogen fuel emissions of 190 ppm, 43% reduction (relative to 335 ppm), were extrapolated (Table 3-18).

Energy Environmental Research reported bench scale test data showing that approximately 78% of the NO_x emitted from 0.77% nitrogen crude was from fuel nitrogen and the remainder from the atmosphere (Ref. 3-29) thereby placing emphasis on testing staged combustion units.

Preliminary data with limited numbers of tests indicated NO_x emissions of 250 ppm for a burner with a heat input of 8 × 106 Btu/hr and 150 ppm for a 40×10^6 Btu/hr (Table 3-17). It was operated at 70% of theoretical air in the primary combustion zone, with an overall stoichiometric ratio of 1.0 (0% excess oxygen). No information on flame shape, smoke, and other emissions was available.

Using an unstaged experimental burner operating at 40 to 50 MM Btu/hr in a large water tube simulator at 150,000 Btu/hr-ft^3 (3.8 kcal/hr-m^3), NO_x emissions were reported (Ref. 3-9). This results in a reduction of NO_x from 354 to 148 ppm (0% O_2) when excess oxygen is reduced from 4 to 2% (Figure 3-20). No smoke was observed at the reduced oxygen level. Also a decrease in NO_x emissions of 10 and 48% (at 4 and 2% excess oxygen) was noted when burner size was increased from 8 to 40 MM Btu/hr.

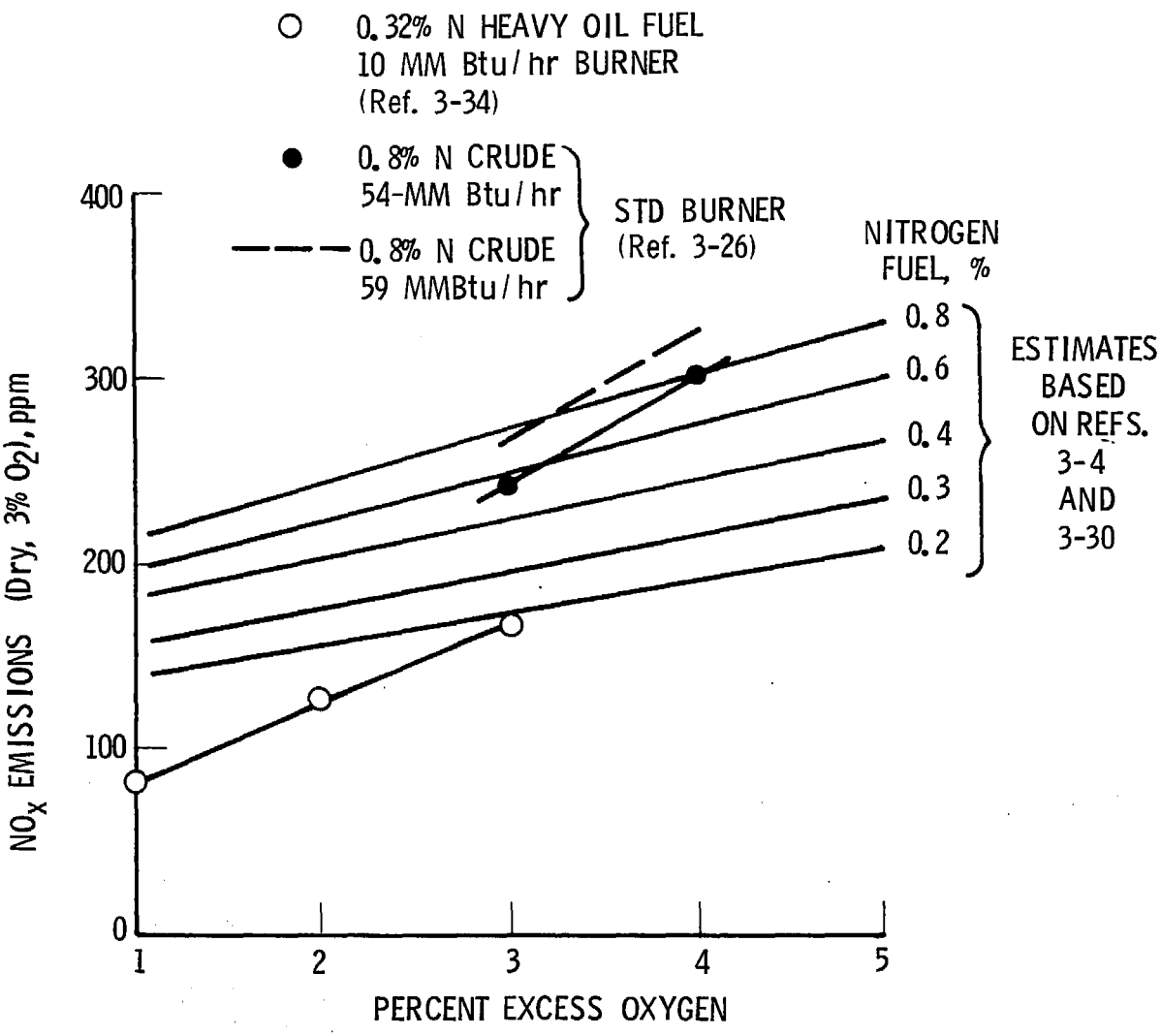


Figure 3-19. NO_x emission test results

TABLE 3-18. SELECTIVE NONCATALYTIC REDUCTION UNITS IN JAPAN (Ref. 3-37)

Toho Gas Co. Plant Site	Gas Source	Fuel	Gas Treated, Nm^3/hr	NO_x Removal, %	Startup Date
Sorami	Boiler	Kerosene	3,490	62	Oct 76
	SNS[a]	Naphtha	2,570	56	Oct 76
	NE[b]	Naphtha	3,540	67	Oct 76
	Boiler	Kerosene	3,840	54	Oct 77
	SNS	Naphtha	4,362	44	Oct 77
	NE	Naphtha	6,216	58	Oct 77
	Boiler	Kerosene	3,870	54	Nov 77
	SNS	Naphtha	4,762	44	Nov 77
	NE	Naphtha	6,216	58	Nov 77
	SNS	Naphtha	3,168	44	Dec 77
	NE	Naphtha	4,166	58	Dec 77
	Boiler	Naphtha	3,169	44	Jun 78
	NE	Naphtha	4,166	58	Jun 78
	Boiler	Low-S Oil	10,700	60	Jul 78
	SNS	Low-S Oil	3,800	62	Jul 78
	NE	Low-S Oil	5,200	62	Jul 78
Chita	NH[c]	Kerosene	6,000	50	Oct 77
	SNS	Kerosene	4,300	50	Oct 77
	NH	Kerosene	3,200	50	Oct 77

[a]Steam naphtha superheater

[b]Naphtha evaporator

[c]Naphtha heater

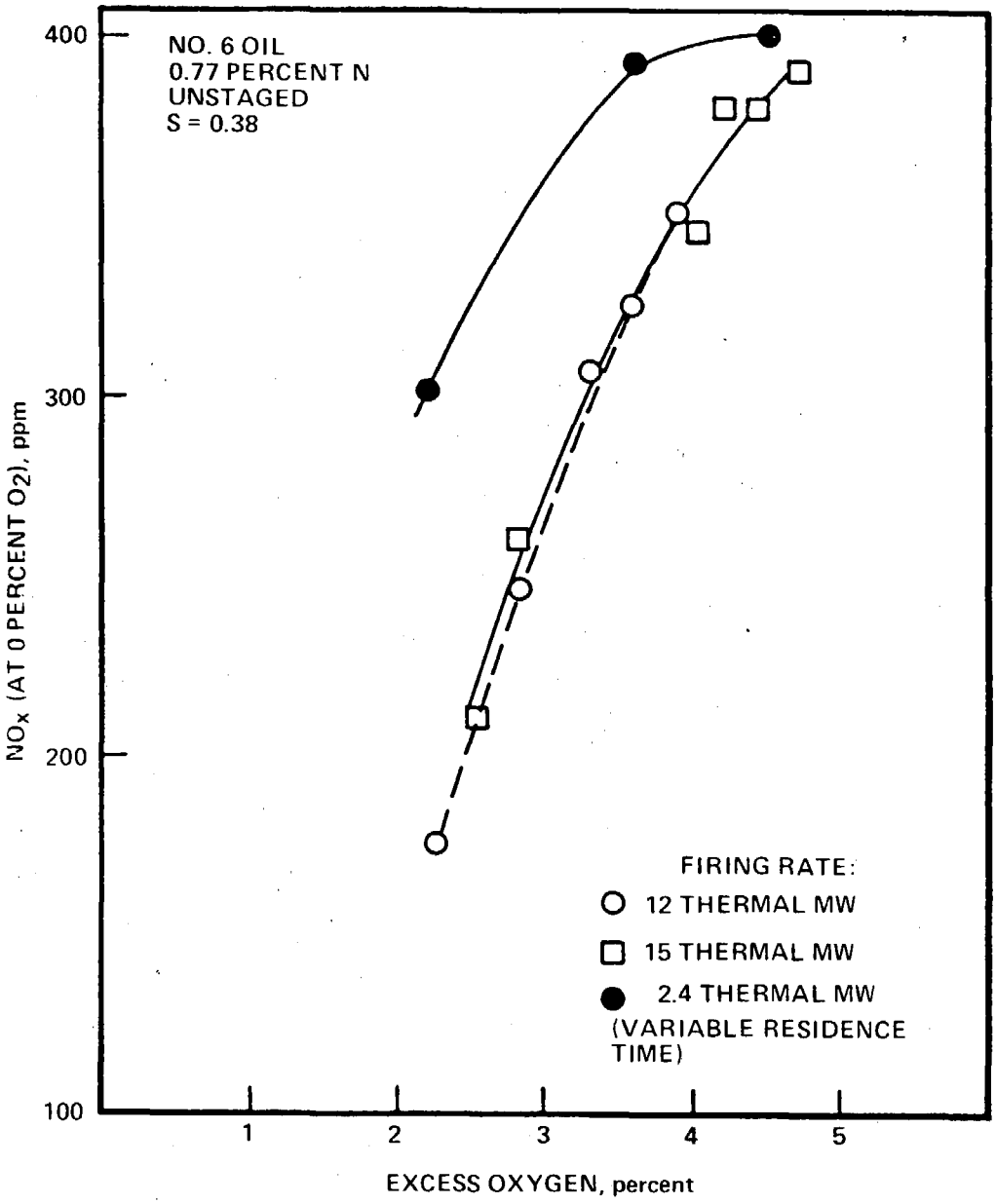


Figure 3-20. Effect of operating conditions and burner size on NO_x emissions (Ref. 3-9)

A TRW LNB with automatic O_2 control was tested on the same Chevron steam generator described in this section under "Low Excess Oxygen" under similar conditions. The TRW burner plus an automatic O_2 control system has the potential to reduce NO_x to levels approximately 8% below those realized with the conventional burner plus O_2 control. NO_x emissions were reduced to levels of 280 and 230 ppm from 305 and 250 ppm, respectively (Ref. 3-31).

Although the tests reported by Chevron were conducted on a single unit, thermal efficiency with the LNB was maintained or slightly improved compared to the thermal efficiency achieved with the conventional burner. This result is shown in Table 3-16.

The main disadvantage reported with the TRW LNB was accelerated tube hanger deterioration due primarily to the rapid change of oxidizing and reducing atmospheres over metal surfaces resulting from the LNB flame. TRW has stated (Ref. 3-36) that the problem occurred abruptly during a phase of testing that involved intense burning over a period of 2 days and that, while operating for the past 4 months under normal conditions since hanger replacement, no abnormal hanger deterioration has been observed. Although this may indicate a transient condition associated with extreme conditions, no definite conclusion can be attempted at this time; thus, Chevron recommends further tests to determine the lowest oxygen level that should be maintained in order to preclude smoke formation, minimize NO_x emissions and hanger deterioration, and maintain thermal efficiency.

3.4.3.2 Thermal Denitrification

A number of commercial applications of selective noncatalytic reduction (SNR) thermal $DeNO_x$ installations comparable in size to the steam generators are installed in Japan (Table 3-18). The volume of gas treated ranges from 3,200 to 10,700 Nm^3/hr. The 50 and 20 MM Btu/hr steam generators produce approximately 13,000 and 5,000 Nm^3/hr, respectively. The Japanese facilities burn clean, low-sulfur fuels.

A demonstration test was conducted in 1977 where thermal $DeNO_x$ was applied to a Chanslor-Western 50 MM Btu/hr Struthers Thermo-Flood steam generator (Ref. 3-26). The results are shown in Figures 3-21 and 3-22 for testing use of a standard and low excess oxygen burner. With a standard

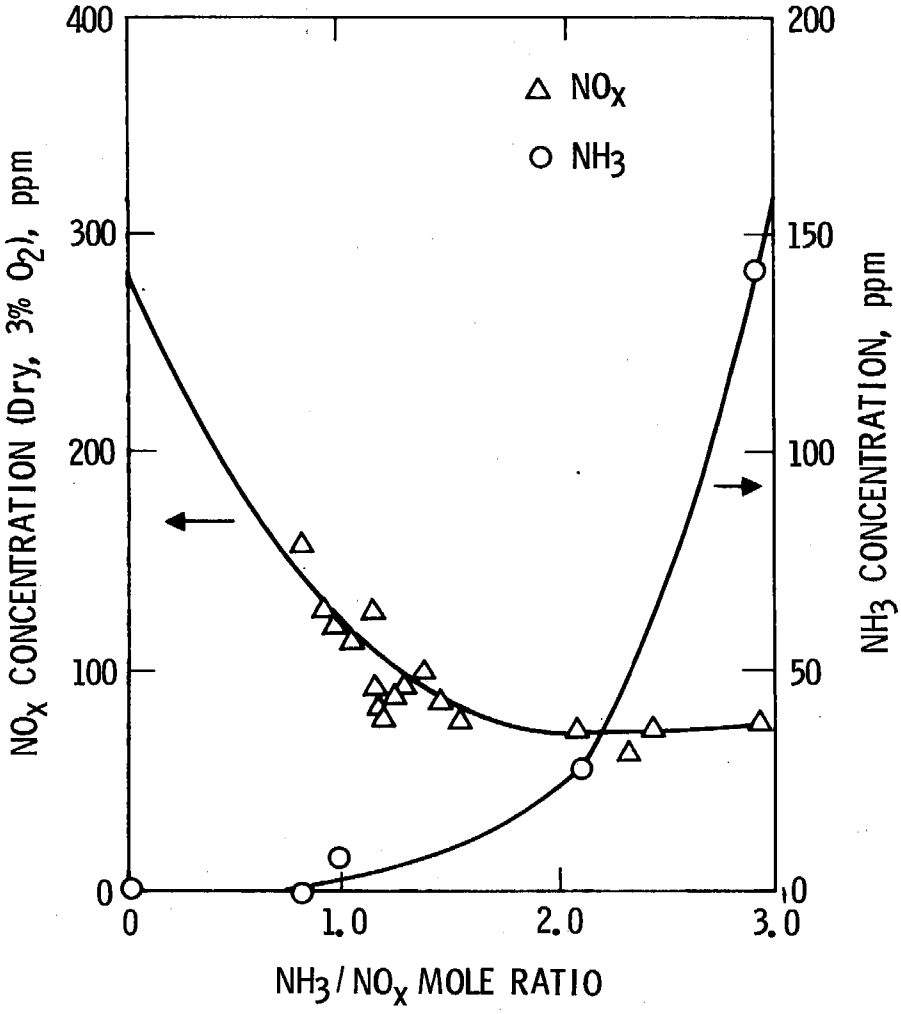


Figure 3-21. NO_x and NH_3 concentration with ammonia injection: standard burner at 5 to 7 GPM fuel flow (Ref. 3-26)

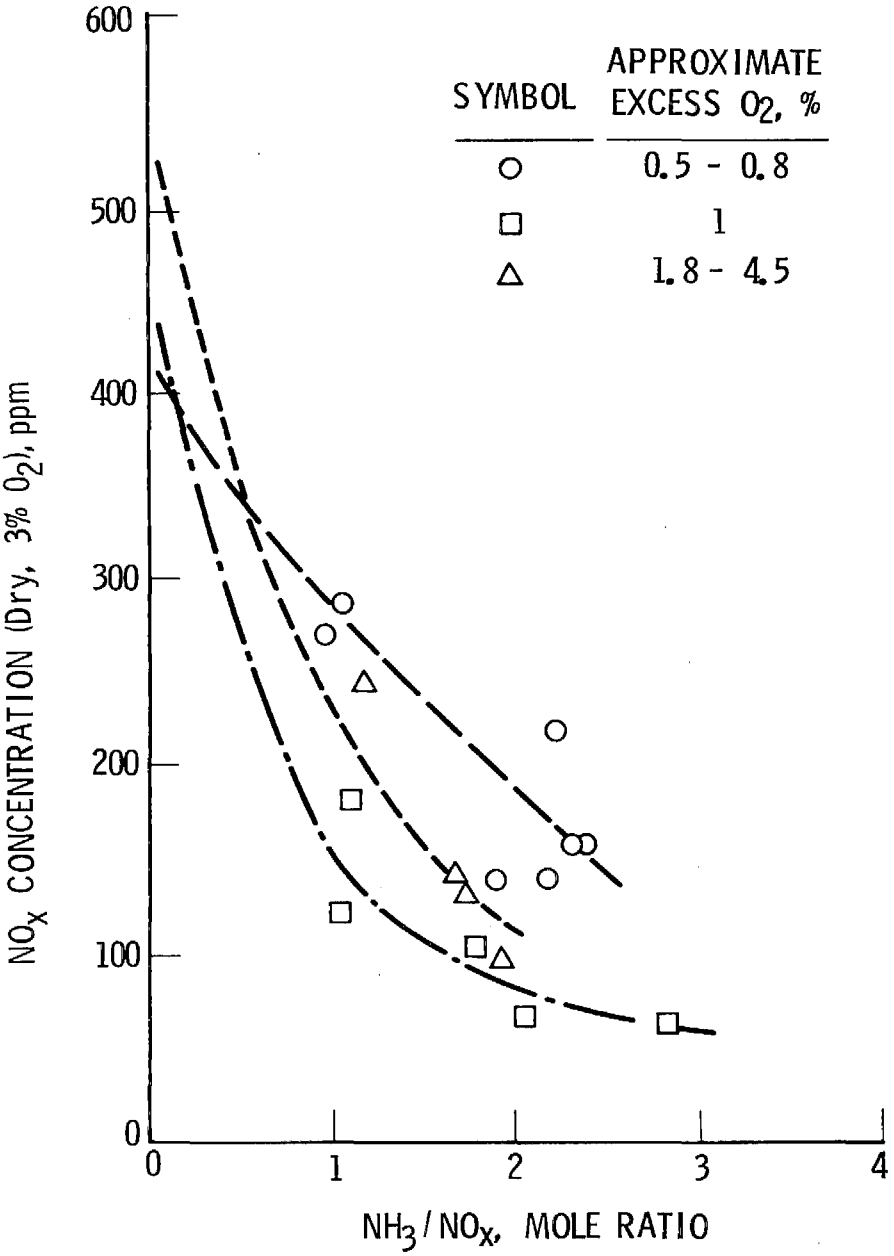


Figure 3-22. NO_x concentration with ammonia injection: low O_2 burner at 5 to 7 GPM fuel flow (Ref. 3-26)

burner, NO_x emissions were reduced from 270 to 120 ppm, or 55%, with a NH_3/NO_x mole ratio of 1.0 and with a fuel rate of 5 to 6 GPM.

Gas temperatures measured immediately downstream of the radiant section ranged from 1615 to 1775°F (880 to 969°C) for 6.1 GPM operation and 1525 to 1700°F (830 to 927°C) at 5.1 GPM for the standard burner configuration.

With a North American low excess oxygen burner (not an LNB), uncontrolled NO_x emissions were 445 ppm at 1% excess O_2 and 530 ppm with 1.8 to 4.5% excess O_2. Thus, with both burners operating at normal, i.e., 4% excess O_2 levels, reductions of 57 and 70% were achieved with NH_3/NO_x mole ratios of 1.0 and 1.5, respectively.

The results were from a single steam generator operating over a limited period of time, and normal variations in equipment and operation could be expected to affect the performance somewhat. At present, this variability is unknown, and additional testing with other generators is needed to establish it. Based on the limited data reported on the performance of the SNR (thermal $DeNO_x$) process (Table 3-19) the 55 and 70% NO_x reduction criteria (Table 3-15) for existing and new steam generators with uncontrolled emissions of about 335 ppm appear feasible, but marginally attainable with NH_3/NO mole ratios of 1.0 and 1.5, respectively. In order to improve margins and allow for equipment and operating variability, the use of thermal $DeNO_x$ in combination with a combustion process modification is suggested.

Ammonia in the gases will react with the SO_3 formed in the combustion process to form ammonium bisulfate. With SO_3 concentration of about 10 ppm and NH_3 concentrations of 10 to 20 ppm, the formation temperature of NH_4HSO_4 is 390°F (200°C) (Figure 3-15).

With the gas temperatures in the convective section exceeding this temperature, its formation is expected, and it may tend to condense on the cooler water tube walls in the convective section. Intermittent water washing has been suggested (Ref. 3-26).

The possible formation of a "blue plume" was suggested based on the interaction of ammonia-based control processes with a post combustion wet SO_2 scrubber (Ref. 3-38). This was postulated on the basis of the presence in the scrubber of NH_3, SO_3, and submicron particles of their reaction products such

as ammonium sulfate, bisulfate and ammonium sulfite, and bisulfate which are not removed in the scrubbing process. However, no data are available on the actual formation or specifics of the suggested phenomenon.

3.4.3.3 Selective Catalytic Reduction

The alternatives available for locating the catalytic reactor are upstream or downstream of the scrubber (Figures 3-23 and 3-24, respectively). Conditions depicted are those expected when operating the generator at a fuel flow of 6.5 GPM.

With the catalytic reactor located upstream of the scrubber (Figure 3-23), an additional 5% (2.65 lb/hr fuel) is consumed to the heat flue gas to 325°C (615°F), which is generally the minimum temperature required in the catalytic reactor. Because of the additional NO_x generated in the reheat process, 72 and 57% removal is required rather than 70 and 55% for new and existing units to achieve 100 and 150 ppm. For SCR application, the gas is relatively dirty, with 885 ppm SO_2 and 0.064 grains/SCF (147 mg Nm_3) of particulates. Therefore, a catalyst configuration that will tolerate these conditions will be required.

Locating the reactor downstream of the scrubber results in a number of operating limitations. The flue gas enters the scrubber at 450°F and is cooled to its adiabatic temperature of approximately 150°F (66°C) (Figure 3-24). The flue gas will then require a significantly greater amount of reheating than the upstream configuration to reach the minimum reaction temperature oi 615°F (325°C). Reheating consumes 8.4 lb/min of fuel or an additional 16% relative to the quantity burned in the steam generator. It will also produce a corresponding amount of additional NO_x and SO_2, approximately 4.2 lb/min and 12.6 lb/hr, respectively. Therefore, the SO_2 removal in the scrubber must be 92.4% rather than 91.1% in order to limit the emissions to 7.1 lb/hr, which would have occurred without the reheat. Also, the SCR must remove 74.5 and 61.4% instead of 70.1 and 55.2% to achieve the equivalent to 100 and 150 ppm (7.6 and 11.5 lb/hr) without the reheat. Also the reactor must be 22% larger (than the alternative configuration) to accomodate the increased volume due to the water evaporated by the flue gas in the scrubber and the larger amount of reheat.

TABLE 3-19. THERMAL DENITRIFICATION CONDITIONS FOR 50 MM Btu/HR STEAM GENERATORS[a]

STANDARD BURNER[b]

Mole Ratio, NH_3/NO_x	Excess O_2, %[c]	NO_x, ppm[d]	% Reduction	NH_3, ppm
0	4	270	--	0
1.0	4	120	57.1	10
1.5	4	85	69.6	20

LOW EXCESS O_2 BURNER[e]

Mole Ratio, NH_3/NO_x	Excess O_2, %	NO_x, ppm[d]	% Reduction	NH_3, ppm	Excess O_2, %	NO_x, ppm[d]	% Reduction	NH_3, ppm
0	1.0	455	--	0	1.8-4.5	530	--	0
1.0	1.0	150	67.0	N/A[f]	1.8-4.5	225	57.5	N/A
1.5	1.0	105	76.9	N/A	1.8-4.5	155	70.7	N/A

[a]Ref. 3-26.

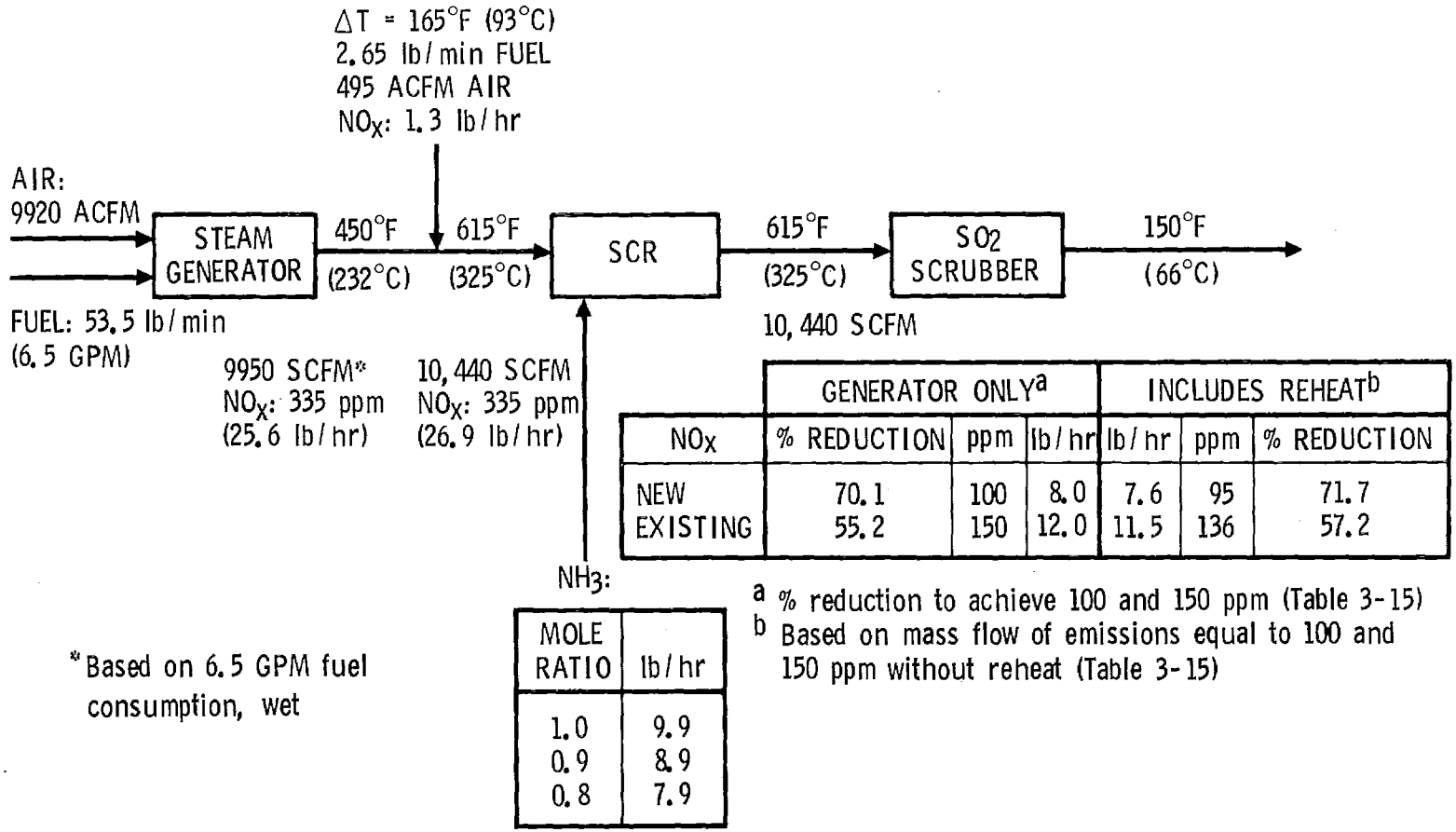
[b]Figure 3-20.

[c]Nominal (as fired).

[d]Dry, 3% O_2.

[e]Figure 3-22.

[f]Not Available.



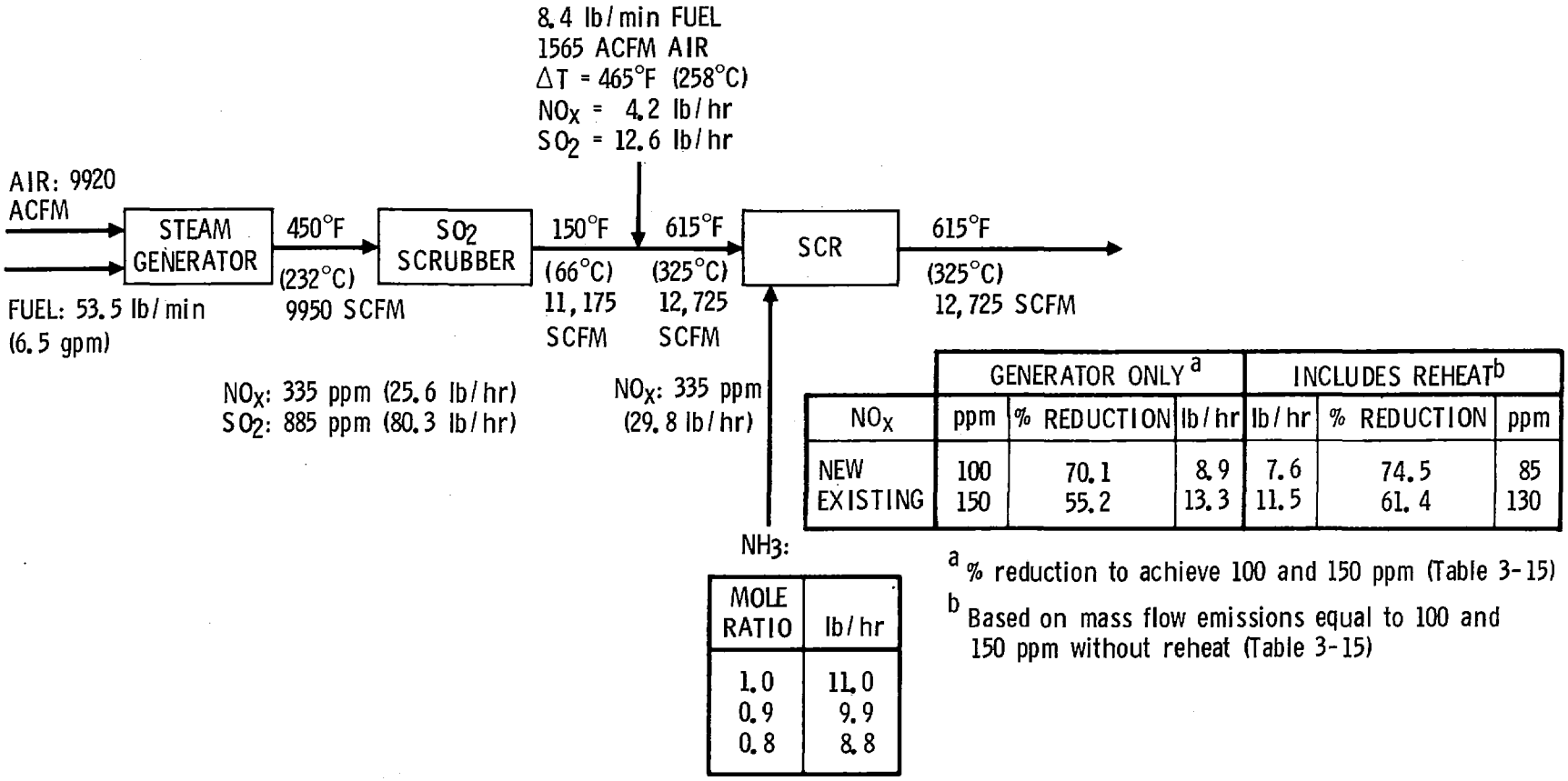
	GENERATOR ONLY[a]			INCLUDES REHEAT[b]		
NOx	% REDUCTION	ppm	lb/hr	lb/hr	ppm	% REDUCTION
NEW	70.1	100	8.0	7.6	95	71.7
EXISTING	55.2	150	12.0	11.5	136	57.2

[a] % reduction to achieve 100 and 150 ppm (Table 3-15)
[b] Based on mass flow of emissions equal to 100 and 150 ppm without reheat (Table 3-15)

MOLE RATIO	lb/hr
1.0	9.9
0.9	8.9
0.8	7.9

*Based on 6.5 GPM fuel consumption, wet

Figure 3-23. Selective catalytic reduction located upstream of SO_2 scrubber



NOx	GENERATOR ONLY [a]			INCLUDES REHEAT [b]		
	ppm	% REDUCTION	lb/hr	lb/hr	% REDUCTION	ppm
NEW	100	70.1	8.9	7.6	74.5	85
EXISTING	150	55.2	13.3	11.5	61.4	130

MOLE RATIO	lb/hr
1.0	11.0
0.9	9.9
0.8	8.8

[a] % reduction to achieve 100 and 150 ppm (Table 3-15)

[b] Based on mass flow emissions equal to 100 and 150 ppm without reheat (Table 3-15)

Figure 3-24. Selective catalytic reduction located downstream of SO_2 scrubber

Guaranteed catalyst lifetimes of 1 to 2 years are being reported (Refs. 3-11, 3-14, 3-17, 3-39, and 3-40). Dust plugging is considered one of the major problems limiting the catalyst lifetimes. To counteract this problem, parallel plate and honeycomb catalysts for fixed bed operation have been developed (Section 2.4.4.1), as well as granular catalysts for moving bed use (Table 3-20). Therefore, in considering SCR, the configuration with the reactor located upstream is not only a feasible approach but also more practical than the alternative because of the lesser amount of reheat required.

In order to achieve high (90%) NO_x removal using honeycomb catalysts with dirty gases, superficial gas velocities of 2 to 6 m/sec (6 to 18 ft/sec), bed depths of 1 to 2 meters (3 to 6 ft), and space velocities of 5000 to 8000 hr^{-1} are required. This results in a pressure drop across the reactor of 30 to 80 mm H_2O (1.2 to 3.1 in. H_2O) (Table 3-21).

Because less stringent (than 90% NO_x removel requirements are required (Figure 3-23), a relaxation of some of the conditions described in the previous paragraphs especially space velocities are possible. As an example, using the information in Figures 3-25 and 3-26 and 10,500 SCFM (17,800 Nm^3/hr), space velocities and superficial velocities of 13,500 hr^{-1} and 8 m/sec can achieve the required 72% removal and 20,300 and 8 m/sec, 57% removal. This results in catalyst bed dimension of 2.6 ft (0.8 m) × 2.6 ft (0.8 m) × 7 ft (2.1 m) for new steam generators and 2.6 ft × 2.6 ft × 4.6 ft (1.4 m) for existing units.

3.5 REFINERY PROCESS HEATER

A representative refinery heater examined for this study is one operated by Chevron, U.S.A., Inc., at its refinery in El Segundo, California. Its rated heat input or total duty is 70 MM Btu/hr, and operating heat input or fire duty is between 60 to 65 MM Btu/hr. It is used to heat a gasoline mixture prior to treatment in a catalytic reformer.

Denitrification technology for reduction of NO_x emissions of 90% relative to current levels was assessed. These included CM and thermal and SCR.

TABLE 3-20. REACTORS FOR DUSTY GAS TREATMENT (Ref. 3-11)

Type	Process Developer	User	Plant Site	Gas Source	Fuel	Capacity (Nm^3/hr)	Completion
Moving Bed	Hitachi Ltd.	Kawatetsu Chemical	Chiba	Coke Oven	COG/RFG[a]	500,000	Nov 1976
	Hitachi Ltd.	Chiyoda Kenzai	Kaizuka	Boiler	HO(HS)[b]	15,000	Oct 1977
	Hitachi Ltd.	Nippon Oils & Fats	Amagasaki	Boiler	HO	20,000	Apr 1978
	Mitsubishi H.I.	Sumitomo Chemical	Sodegaura	Boiler	HO(LS)[c]	382,000	Sept 1976
	Kobe Steel	Kansai N.K.	Amagasaki	Coke Oven	COG	104,000	Aug 1977
	Kurabo	Kurabo	Hirakata	Boiler	HO(HS)	30,000	Aug 1975
	Asahi Glass	Asahi Glass		Furnace		70,000	
Honeycomb	Mitsubishi H.I.	Fuji Oil	Sodegaura	Boiler	HO(LS)	200,000	Jan 1978
	Mitsubishi H.I.	Chubu Electric	Chita	Boiler	HO(LS)	1,920,000	Feb 1980
	Ishikawajima H.I.	Chubu Electric	Taketoyo	Boiler	Crude Oil	20,000	Apr 1977
	Ishikawajima H.I.	Ajinomoto	Kawasaki	Boiler	HO(LS)	180,000	Jan 1978
	Ishikawajima H.I.	Company E	--	Boiler	HO(LS)	960,000[d]	Apr 1978
	Ishikawajima H.I.	Company F	--	Boiler	HO(LS)	480,000[d]	June 1978
	Ishikawajima H.I.	Chugoku Electric	Kudamatsu	Boiler	Crude Oil	1,000,000	Apr 1979
	Ishikawajima H.I.	Chugoku Electric	Kudamatsu	Boiler	HO(LS)	1,900,000	July 1979
	Ishikawajima H.I.	Tohoku Electric	Niigata	Boiler	HO(LS)	1,660,000	Aug 1981
Parallel Passage	JGC	Kashima Oil	Kashima	Furnace		50,000	Nov 1975
	JGC	Fuji Oil	Chiba	Boiler	CO	70,000	July 1976
	JGC	Nippon Steel	Kimitsu	Coke Oven	COG	150,000	Mar 1977
Parallel Plate	Mitsubishi H.I.	Tokyo Electric	Yokosuka	Boiler	HO(LS)	40,000	Mar 1977
	Mitsubishi H.I.	Company C	--	Boiler	HO(LS)	1,010,000	Feb 1978
	Mitsubishi H.I.	Company D	--	Boiler	HO(LS)	490,000	July 1978
	Hitachi, Ltd.	Company A	--	Boiler	HO(LS)	490,000	June 1978
	Hitachi, Ltd.	Company B	--	Boiler	HO(LS)	550,000[d]	June 1978
Tube	Mitsui Engineering	Ukishima Pet. Chem.	Chiba	Boiler	HO(LS)	220,000	Apr 1978

[a] Coke oven gas/refinery off gas.
[b] Heavy oil (high sulfur).
[c] Heavy oil (low sulfur).
[d] Combination with selective noncatalytic reduction.

TABLE 3-21. COMPARISON OF CATALYSTS AND REACTORS (Ref. 3-11)

	Moving Bed	Honeycomb (Metallic)	Honeycomb, Tube (Ceramic)	Parallel Passage	Parallel Plate (Ceramic)	Parallel Plate (Metallic)
Catalyst Size (mm)						
Diameter	4-8[b]					
Thickness		0.5-1	1.5-3	7-10[e]	8-10	1
Opening		4-8	6-20	7-10	8-14	5-10
Gas Velocity (m/sec)[c]	0.5-1.5	2-6	5-10	5-10	5-10	4-8
Bed Depth (m)	0.2-0.6	1-2	1.5-5	3-5	4-6	2-5
SV (1000 hr^{-1})[d]	5-10	5-8	4-8	3-5	1.5-3	2-5
Pressure Drop (mmH_2O)[f]	40-80	30-80	40-140	80-140	80-160	60-120

[a] 90% NO_x removal at an NH_3/NO_x ratio of 1.1 and a temperature of 350-400°C for dirty gas.
[b] Ring type catalysts with larger diameters are also used.
[c] Velocity at 350-400°C in open column (superficial velocity).
[d] Gas volume (Nm^3/hr)/catalyst bed volume (m^3).
[e] Thickness of envelope.

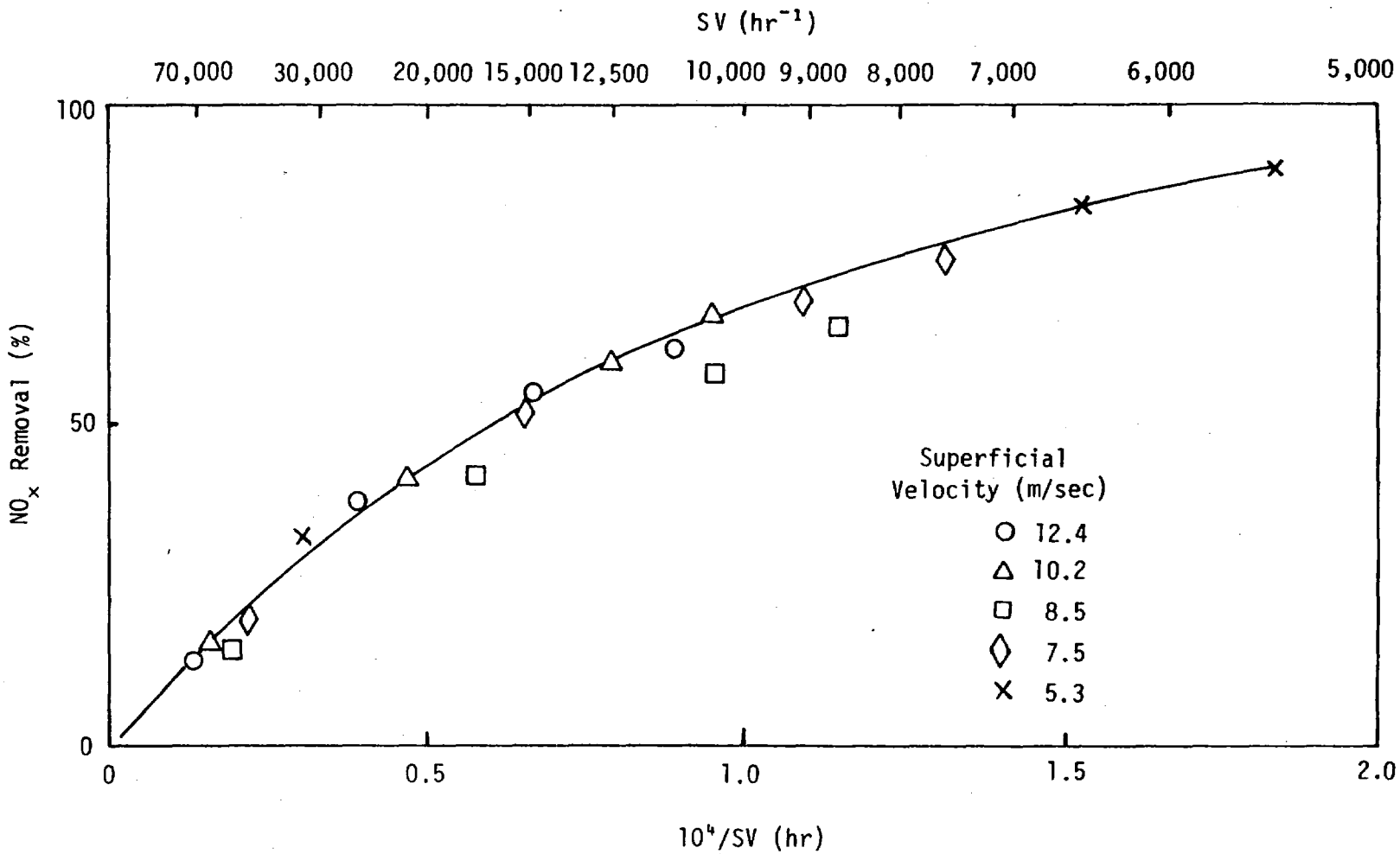


Figure 3-25. NO_x removal at 300°C (NO = 163 to 178 ppm, NH_3/NO = 1) (Ref. 3-11)

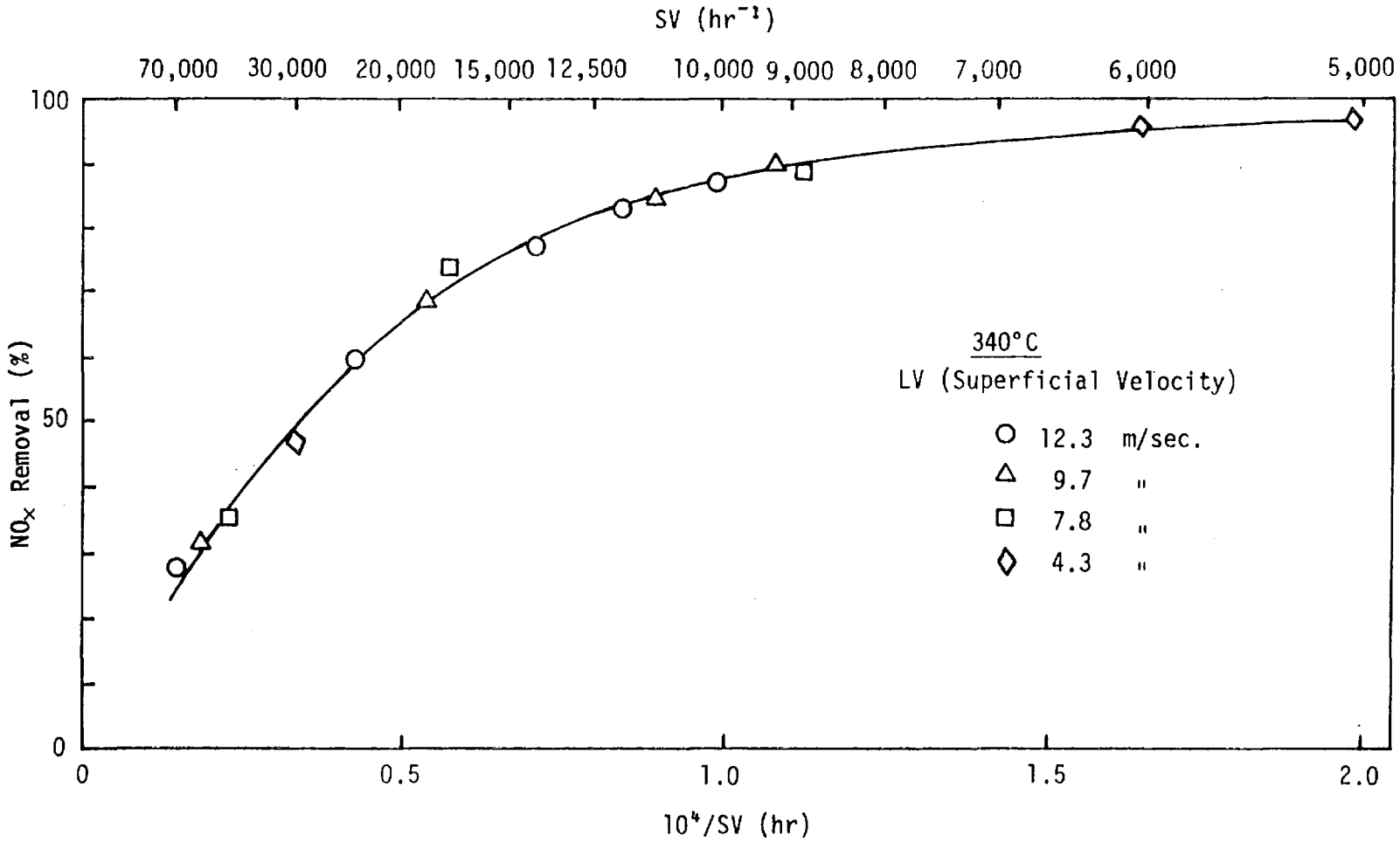


Figure 3-26. NO_x removal at 340°C (NO = 176 to 184 ppm, NH_3/NO = 1) (Ref. 3-11)

3.5.1 Heater Characteristics

Off-gases from various processes are collected and mixed in a common storage system and supplied to various heaters as required. As a result, the composition of the gases may change during steady operation of a given heater. A typical gaseous fuel composition is shown in Table 3-22. The fuel is burned with more than normal excess air (about 20% excess air or about 4% O_2 in the flue gas) at constant flow rate to assure the availability of excess air despite variations in the fuel composition. Figure 3-27 illustrates the fuel and air flow rates to the unit.

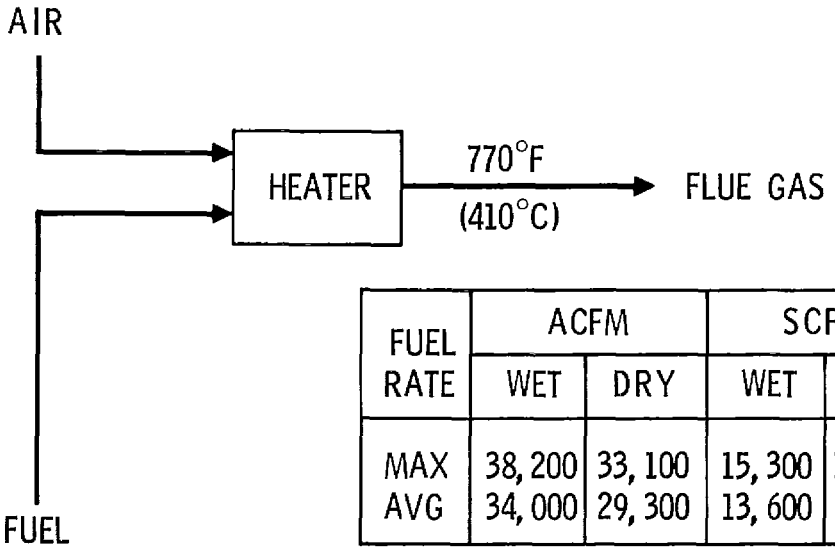
TABLE 3-22. TYPICAL REFINERY HEATER GASEOUS FUEL

Constituent	Volume %
H_2	12.6
N_2	0.2
CO	0.6
CH_4	32.8
C_2	17.7
C_3	27.2
C_4	7.5
C_5	1.4
	100.0
Btu/SCF	1476

The fuel is burned in 24 natural draft gas burners, which are arranged linearly along the floor near a refractory wall. Burner capacity is 2.9×10^6 Btu/hr* (73×104 kcal/hr). The heaters utilize combustion air at ambient temperature. Tubes carrying process fluid are located on the opposite wall in the radiant section and in the convective section along the roof.

*70×10^6 Btu/hr x 1/24 burners = 2.9×10^6 Btu/hr-burner.

FUEL RATE	AIR, SCFM	EXCESS O_2, %
MAX	14,300	5.0
AVG	12,300	3.8



FUEL RATE	ACFM		SCFM		NO_x	
	WET	DRY	WET	DRY	ppm[a]	lb/hr[b]
MAX	38,200	33,100	15,300	13,200	85	7.9
AVG	34,000	29,300	13,600	11,700	77	6.7

[a] 3% O_2, dry
[b] As NO_2

FUEL RATE	FUEL, SCFM	HEAT INPUT, MMBtu/hr
MAX	1050	65
AVG	950	58

Figure 3-27. Refinery heater characteristics

Part of the combustion air is premixed with the fuel, with the rest entering close to the burner as secondary air. Both fuel and combustion air are introduced at ambient temperature. The combustion gases are directed against and along the refractory wall.

The wall is heated (glowing in some spots) and provides radiant heating of the tubes carrying the gasoline mixture. The gases, which have cooled considerably, then pass through a bundle of tubes located in the roof of the heater and a steam generating coil in a final convective pass, before entering the stack. Temperature at the stack is about 770°F (410°C). The combustion system is relatively simple, and the heat transfer arrangement assures relatively uniform heating of the process fluid despite any localized hot spots which might exist in the gases or on the refractory wall.

Furnace operation is essentially continuous at approximately 60 MM Btu/hr heat input with scheduled shutdowns approximately every 4 to 6 months (for 2 weeks) for catalyst regeneration and minor repairs, and every 2 years for about a month during catalyst dumping, screening, reloading, and major maintenance. The furnaces are about 2 years old with an unknown life expectancy. No specific air pollution controls are used on these units at this time other than operation at minimum excess air.

3.5.2 NO_x Emissions

Current NO_x emissions as reported are 70 to 85 ppm (adjusted to 3% O_2) at maximum duty. Oxygen concentration in the flue gas averages 3.8% and is 5.0% maximum. The quantity of NO_x emitted is summarized in Figure 3-27. NO_x emission rates (expressed as NO_2) are 6.7 to 7.9 lb/hr for heat input of 58 and 65 MM Btu/hr, respectively. Emissions of 7.8 and 8.8 lb/hr were computed for heaters using refinery gas fuel and based on factors reported in Reference 3-18 are in reasonable agreement with those computed for this unit.

3.5.3 NO_x Control Alternatives

The control alternatives reviewed for retrofitting natural draft refinery heaters for reducing NO_x emissions 50% include CM (principally LNB), and thermal and selective catalytic $DeNO_x$ with the use of ammonia.

The gaseous fuel contains no organically bound nitrogen (although it may contain small quantities of molecular nitrogen). Therefore, all the NO_x is formed as a result of thermal fixation of the molecular nitrogen in the combustion air. As will be discussed in succeeding sections, the potential reduction for LNB is 30 to 55%, for thermal $DeNO_x$ is 40 to 60%, and for SCR is 90% or greater.

3.5.3.1 Combustion Modifications

There are various CM techniques in addition to LNB that may be applied to reduce NO_x emissions on natural draft heaters. These include low excess air firing, BOOS, OFA, FGR, and reduced firing rate. Although these techniques may be used on existing heaters, modifications to the units may be necessary, and in all cases the effect of flame shape, size, characteristics, and stability on the operation of the unit must be considered.

Modification Techniques

Since formation of virtually all of the NO_x is due to the combustion temperatures involved, two methods of staging the combustion to control the NO_x are possible:

a. A conventional approach, involving initial fuel-rich combustion, with the remaining excess air introduced higher up in the furnace after the initial combustion products have sufficiently cooled

b. The reverse, involving initial air-rich combustion followed by introduction of the remaining fuel

The latter case is only possible because the fuel contains no organically bound (only molecular) nitrogen.

The choice between these two approaches is based largely on avoiding air-fuel ratios near the stoichiometric value. The final air-fuel ratio must be greater than stoichiometric, to assure complete combustion. If the first stage is very fuel-rich, then great care must be taken to assure relatively low-peak temperatures when the second stage air is added and the local, mixed air-fuel ratio passes through stoichiometric in reaching the final air-rich mixtures. On the other hand, if the initial, premixed reactants are slightly air-rich and are immediately quenched to very air-rich, low-temperature (less

than about 2960°F) mixtures, then the remaining fuel can be added, in a second stage after adequate cooling, with minimum time spent at stoichiometric mixtures. The small amount of fuel remaining to be added, in the second stage, can be more easily and rapidly mixed with the products of air-rich combustion from the first stage. In addition, this technique would avoid not only a highly fuel-rich atmosphere in the first stage and the corresponding potential problem of sooting on the refractory wall and the product tubes, but also a more difficult condition in controlling combustion.

A retrofit modification of this type would require that the amount of fuel to the burners be reduced and that new fuel injection and mixing ports (probably one per burner) be installed in the refractory wall. This would require some information on temperature profiles in the furnace. Since this furnace often operates at skin temperature limits, any increase in flame length could be critical to its normal operation and require furnace derating. Also, because this method has not been demonstrated, a preliminary development and demonstration program would be required. NO_x reduction in excess of 50% appears feasible.

For the fuel used in this heater, combustion of 86% of the fuel with all of the combustion air (approximately 71% excess air premixed with the fuel and the rest rapidly mixed with these products) would yield an initial first stage combustion product temperature of about 2960°F (1900 K). After rejecting sufficient heat to the process fluid to cool these gases to 2945°F (1726 K), the remaining fuel could be added to bring the final mixture to that represented by 4% O_2 and to return the mixture temperature to 2960°F (1900 K) for further heat rejection. The fuel-air staging with all reactants introduced at ambient temperatures would be approximately as shown in Table 3-23.

Subscale Tests

Results of subscale tests conducted with natural draft process heater burners are reported in Reference 3-41. These tests were conducted to develop CM requiring minor hardware modifications for use by operators or manufacturers to control emissions. Subscale testing was considered necessary to ensure acceptable performance prior to developing full-scale configurations. While the results may not be indicative of full-scale performance, it is

TABLE 3-23. STAGED COMBUSTION CHARACTERISTICS

	Fraction of Total by Volume		Temperature	
	Fuel	Air	°F	°K
Pre-mixed	0.86	0.71	3830	2383
First Stage Quench	-	0.29	2960	1900
After Cooling	-	-	2645	1726
Total First Stage	0.86	1.00	-	-
Second Stage (Initial)	0.14	0.00	2960	1900
Into Stack	1.00	1.00	770	680

believed that they represent an opportunity for assessing the relative potential of various NO_x control alternatives.

The various modifications that were tested included:

a. Lowered excess air (LEA)

b. Steam injection (SI)

c. Staged combustion (two methods)

d. Flue gas recirculation (FGR)

e. Modified fuel injection

f. Commercial LNB (two designs)

Test conditions included the use of natural gase burned at a nominal firing rate of 5 MM Btu/hr (1.5 MW_e) in a natural draft uncooled furnace with dimensions of 8 ft (34 m) wide × 6 ft (1.8 m) deep × 32 ft (9.8 m) high.

The largest percent reduction in NO_x emissions (30 to 67%) occurred with staged combustion air (SCA) and with FGR modifications (59 to 63%) (Table 3-24). With SCA, the reductions were found to be a relatively strong function of the amount of excess air. Reductions of 45 to 67% were observed for low excess oxygen which had been reduced from 3% to approximately 1%. With FGR, the effect of reducing excess air was relatively slight, only a few percent. The simplest modifications studied other than LEA were the central staging cylinder and altered fuel injection geometry (AIG). AIG as implemented in that study produced local fuel-rich zones in the flame. The OSC resulted in reduced flame zone temperatures and lower overall NO_x production. However, the burner produced a shorter than normal segmented flame which may be undesirable for practical application. The central cylinder technique for staged combustion whereby secondary air was introduced into the flame zone after the primary combustion zone produced large reductions in NO_x increased furnace efficiency and was considered one of the simplest modifications to implement.

Full-Scale Modifications

United States. A series of operational modifications to refinery heaters was reported in Reference 3-18. These included the effect of fuel compositions, reduced excess air, and TSC. In general, the design of natural draft units was found to be very limited in the degree of adjustments. As a result, emissions and efficiency were fairly unresponsive to CM capable of being incorporated on existing units.

The effect of fuel gas higher heating value on NO_x emissions was minor. There was approximately a 6% increase in NO_x when the heating value was increased 10% to approximately 1170 Btu/SCF by addition of propane.

The effect of lowering excess air on NO_x emissions was reported as difficult to evaluate because of process limitations on the amount of air adjustment. On some heaters, air register adjustments resulted in NO_x reductions of 8 to 27% (Table 3-25), while in one case NO_x increased relative to baseline levels.

TABLE 3-24. NO_x REDUCTION TESTS ON A SUBSCALE NATURAL DRAFT PROCESS HEATER (Ref. 3-41)[a]

Modifications	Baseline NO_x[b]			Burner[c] Desig.	Max NO_x Reduction, %	Remarks
	O_2	ppm, Dry 3% O_2	ng/J			
Lowered Excess Air	3.0	113	58	MA16	10	
Steam Injection	3.0	107	54.6	MA16	33	
Staged Combustion[d]	3.0	120	61.2	MA16	46	Normal Excess O_2
Staged Combustion[d]	3.0	120	61.2	MA16	67	Low Excess O_2
Staged Combustion[e]	3.0	131	66.8	DBA16	31	Normal Excess O_2
Staged Combustion[e]	3.0	131	66.8	DBA16	59	Low Excess O_2
Flue Gas Recirculation	3.0	116	59.2	MA16	59	Normal Excess O_2
Flue Gas Recirculation	3.0	116	59.2	MA16	63	Low Excess O_2
Altered Fuel Injection	3.0	131	66.8	DBA16	31	Normal Excess O_2
Altered Fuel Injection	3.0	131	66.8	DBA16	44	Low Excess O_2
Low NO_x Burner (Tertiary Air Burner)	3.2	107	54.6	--	30	Lowest NO_x Configuration
Low NO_x - Recirculation Tile Burner	3.1	107	54.6	--	3	Lowest NO_x Configuration

[a] Fuel: Natural Gas

[b] NO in NO_x: 95-100%

[c] Heat Input Rate: MA-16 and DBA-16, 5.2 MMBtu/hr (1.53 MW_e); Tert Air Burner, 5.1 (1.49 MW_e); Recirc. Tile Burner, 5.1 (1.49)

[d] Floor Lances

[e] Central Cylinder

TABLE 3-25. SUMMARY OF NO_x REDUCTION FIELD TESTS ON NATURAL DRAFT REFINERY HEATERS (Ref. 3-18)

Unit	Input, MMBtu/Hr, (MW_t)	Fuel	No. Burners	Baseline NO_x[a]			Maximum NO_x Reduction, %	Modification
				O_2 %	3% O_2 Dry			
					ppm	lb/hr		
1	86.6 (25.0)	Ref. Gas + Nat. Gas	32	5.4	116	11.9	18	4 of 32 BOOS[b]
2	65.5 (18.9)	Ref. Gas	16	4.6	97	7.5	22	Adjust air register
3	34.6 (10.1)	Ref. Gas	10	3.7	76	3.1	21	High load, low O_2
4	52.8 (15.1)	Nat. Gas	16	7.0	103	6.4	8	2 of 16 BOOS
5	50.9 (14.9)	Nat. Gas	12	8.1	98	5.8	27	Adjust air register

[a] NO in NO_x: 95-100%

[b] BOOS: Burners Out Of Service

Staging the combustion by BOOS resulted in reduction of 18% on unit 1, 8% on unit 4, and no reduction was observed on several tests with two BOOS combinations attempted on unit 5.

Japan. In Japan a number of CM techniques are employed on oil heating furnaces (Ref. 3-11). The modifications are primarily LEA combustion and steam injection, with a few controlling NO_x by a change in fuel or reducing the heat load (Table 3-26). Low NO_x burners are used extensively for both new and existing furnaces especially in the size range of the heater considered in this study. Generally, TSC and FGR are not used because most of the furnaces use natural draft and incorporating these modifications would require extensive modifications. Also, the LNB in use are generally a combination of self-recirculation and TSC. The level of NO_x reduction specifically from the LEA combustion and TSC was not available; however, data from LNB that incorporate these features indicate that NO_x reductions in the range of 35 to 60% are being attained. NO_x levels in the range of 57 to 68 ppm (corrected to 3% O_2) are emitted in oil heating furnaces with self-recirculating type burners burning gaseous fuels. Existing unmodified units emit in the range of 124 to 170 ppm, while new units are in the range of 79 to 113 ppm.

Nippon Furnace Kogyo Kaisha, Ltd. (NFK) has developed a self-recirculation gasification (SRG) burner as part of its "Mini NO_x" series for natural draft application (Ref. 3-42). Application of this LNB includes gas, oil, and combination firing in process heaters. Over 4000 burners of this type are in operation on various installations in Japan (Ref. 3-43).

NO_x levels can be reduced by 30 to 50% relative to their existing burners (Figure 3-28) with levels less than 50 ppm (corrected to 3% O_2) for operation with 10 to 20 % excess air (Figure 3-29).

The burner employs the effects of both FGR and TSC. The burner recirculates air and part of the combustion products in the burner tile by the action of the injected fuel and atomizing steam in the case of oil fuel. The burner also introduces secondary air (Figure 3-30).

TABLE 3-26. COMBUSTION MODIFICATION TECHNIQUES FOR OIL HEATING FURNACES[a]

Capacity ($1,000 Nm^3/Hr$)	LEA[c]		SI[d]		LNB[e]		CF[f]		LEA+SI		FGR		FGR+ LNB		OTHER COMB.		TOTALS	
	N[g]	E[h]	N	E	N	E	N	E	N	E	N	E	N	E	N	E	N	E
> 100	-	1	-	-	1	2	-	2	-	-	-	-	-	-	1	-	2	5
40 - 100	-	3	-	2	11	8	1	1	-	-	-	3	3	-	-	1	15	18
10 - 40[b]	-	5	-	-	17	22	-	3	-	4	-	-	3	-	-	2	20	36
> 10	-	2	-	1	7	5	-	-	-	-	-	-	-	-	-	-	7	8
Totals	-	11	-	3	36	37	1	6	-	4	-	3	6	-	1	3	44	67
																	111	

[a] Ref. 3-11

[b] 60 MMBtu/hr Refinery Heater Equivalent to 20,500 Nm^3/hr

[c] Low Excess Air

[d] Steam Injection

[e] Low NO_x Burner

[f] Change of Fuel

[g] New

[h] Existing

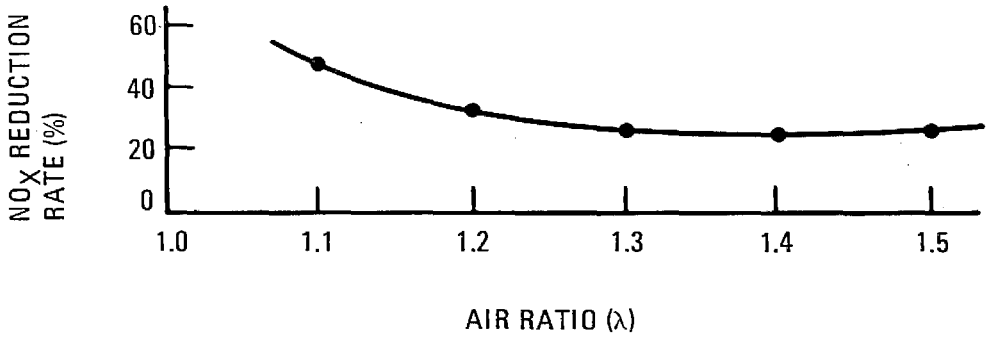


Figure 3-28. NO_x reduction rate: self-recirculation gasification versus conventional burner (Ref. 3-42)

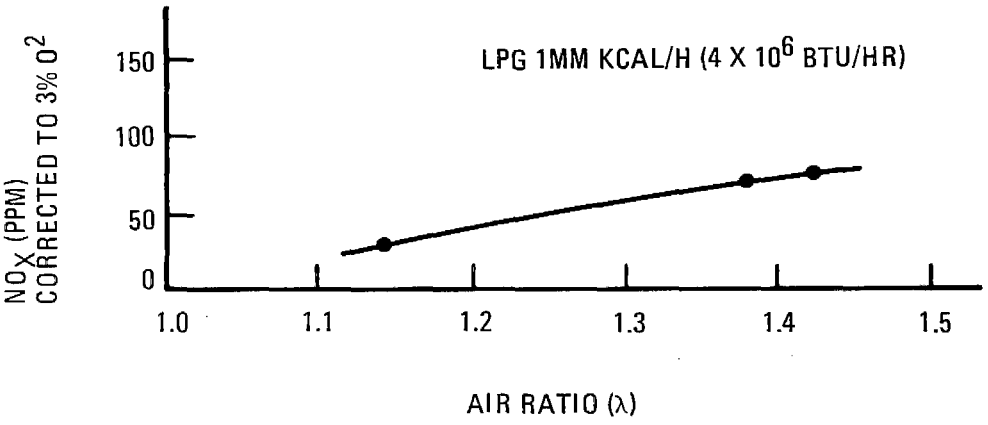


Figure 3-29. NO_x versus air ratio: self-recirculation gasification burner (Ref. 3-42)

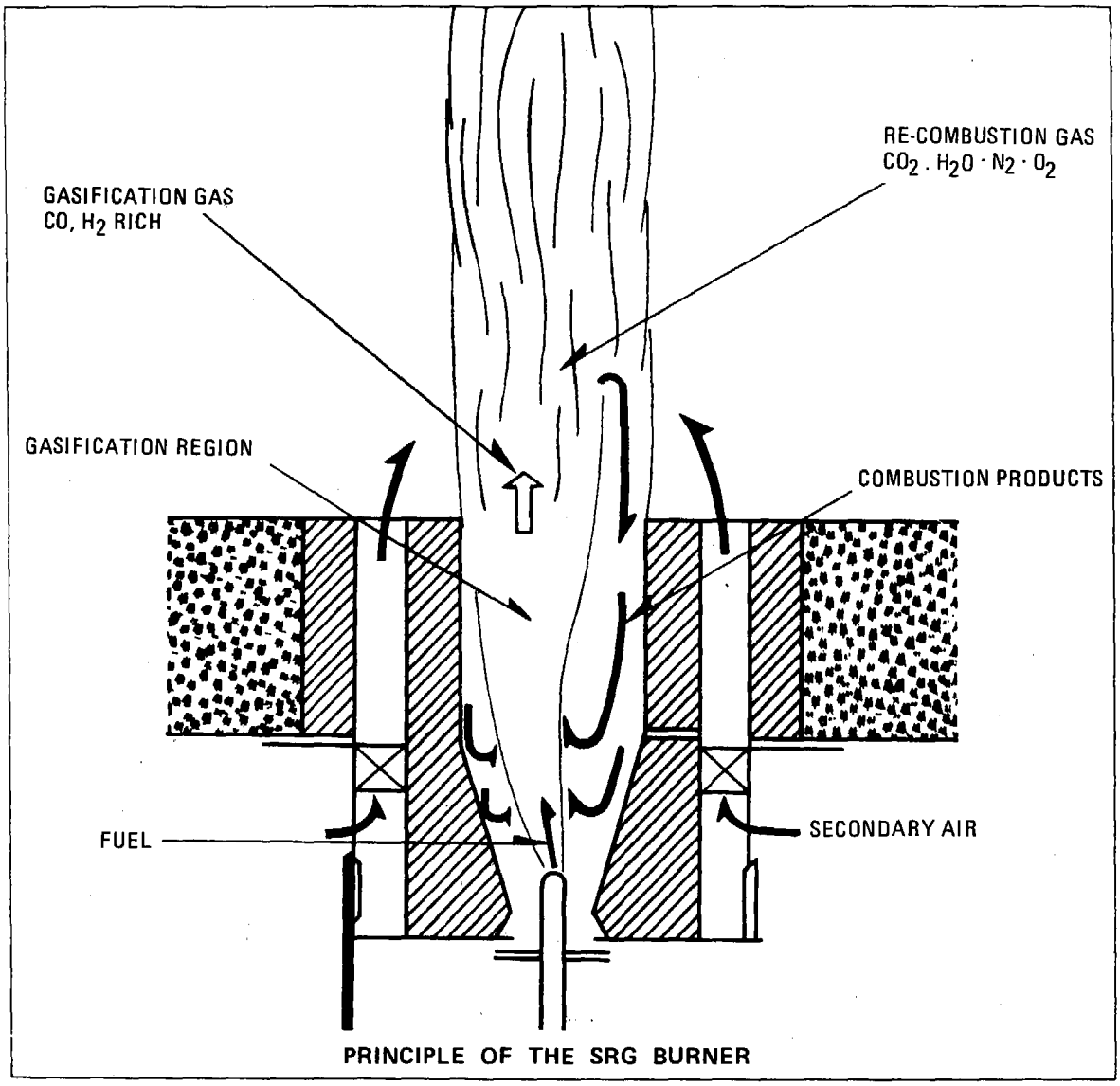


Figure 3-30. Principle of the self-recirculation gasification burner (Ref. 3-47)

Mitshubishi Heavy Industries has developed an LNB for natural draft furnaces (Ref. 3-15) for various fuels and capable of being installed on existing units (Table 3-27). Emission characteristics of burners determined in a test furnace are shown in Figures 3-31 and 3-32 as a function of oxygen content in the flue gas. The effect of the heat release in the furnace capacities installation is shown in Figure 3-33. For burner capacities in the 1.0 to 3.0 $\times 10^6$ kcal/hr (4 to 10 $\times 10^6$ Btu/hr) range, operation with propane gas results in NO_x emissions of 30 ppm (6% O_2) for operation at 4% O_2 in the flue gas. This equates to 35 ppm (3% O_2). The refinery furnace generates approximately 80 ppm (3% O_2) at 4% O_2 in the flue gas (Figure 3-27) for burners operating at 3 $\times 10^6$ Btu/hr. Therefore, for this application, use of the MHI EM-type burner has the potential of reducing NO_x 55% relative to existing burners.

Kobe Steel (Refs. 3-44 and 3-45) has developed an LNB for its high-temperature heating furnaces. These furnaces are forced draft units and have achieved reductions in NO_x emissions of 50 to 70% for coke oven gas (COG) and butane, relative to their conventional burners. The LNB is a self-recirculation type burner wherein a soft flame is formed with a uniform temperature distribution. Because of a higher flame emissivity and burner radiant heat flux relative to conventional burners, a reduction of fuel consumption of 5 to 10% is reported for heating furnaces and soaking pits. Effects of varying parameters such as excess air and combustion air temperature for burners with 40 $\times 10^4$ kcal/hr (1.6 $\times 10^6$ Btu/hr) tested in a refractory wall furnace 1000 mm in diameter and 4000 mm in length is shown in Figures 3-34 and 3-35.

3.5.3.2 Thermal Denitrification

Thermal DeNOx has been applied in Japan on several crude oil heaters and atmospheric/vacuum pipestills in addition to industrial and utility boilers (Ref. 3-46). Characteristics of the installations are summarized in Table 3-28. NO_x reduction is in the range of 51 to 63% at approximately full load. An example effect of load variation on $DeNO_x$ performance in a pipestill is shown in Figure 3-36. Location and configuration of NH_3 injection sites were not available. As discussed in Section 2.3.2, temperatures in the range of 900 to 950°C (1650 to 1740°F) are required for optimal NO_x conversion, with

TABLE 3-27. MHI STANDARD TYPE LOW NO_x BURNER FOR NATURAL DRAFT APPLICATIONS (Ref. 3-15)

Pressure Inside Furnace	-5 to -30 mmH_2O (0.2 - 1.2 in H_2O)
Capacity Per Burner	50 - 450 x 10^4 Kcal/hr (2 - 18 x 10^6 Btu/hr)
Fuel	Single Fuel or Mixed Fuels (Gases and/or Oils)
Turndown	1/6 Gas 1/3 Heavy Oil 1/4 Light Oil (e.g. naphtha)
*Capacity Range	
Type 100	50 - 150 x 10^4 Kcal/hr
200	150 - 250
300	250 - 350
400	350 - 450

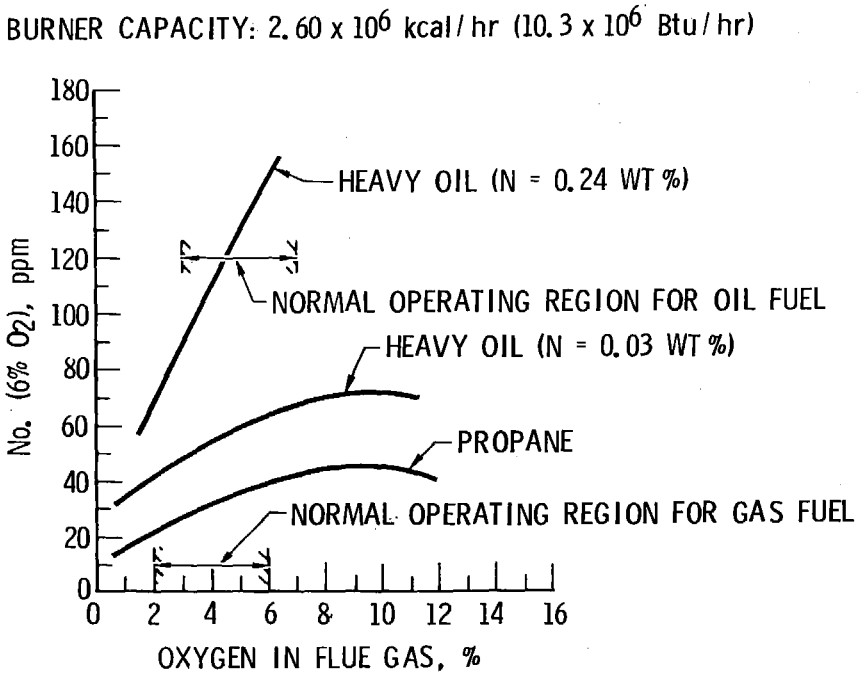


Figure 3-31. NO_x emissions from a rectangular type burner (EM-R) with deflector in a test furnace (Ref. 3-15)

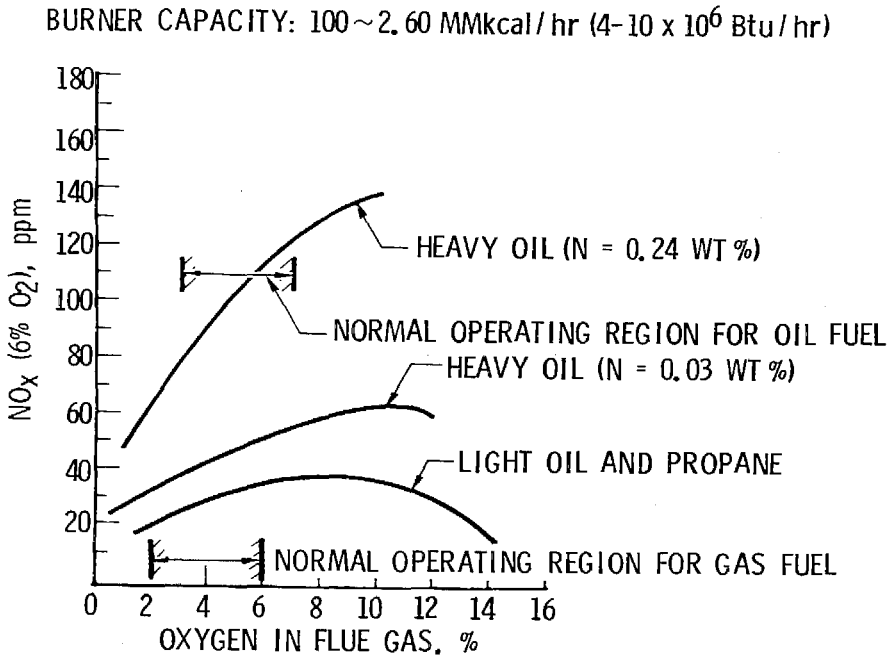


Figure 3-32. NO_x emissions from a rectangular type burner (EM-R) with deflector in a test furnace (Ref. 3-15)

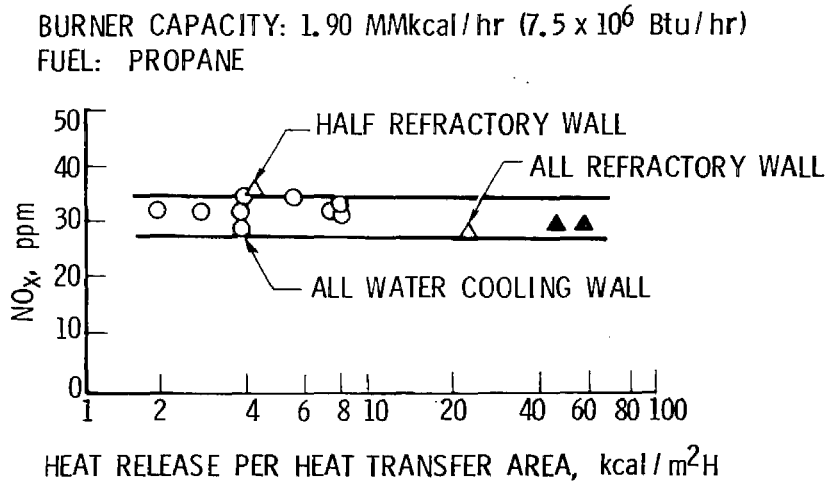


Figure 3-33. Relationship between NO_x emission and heat relief (Ref. 3-15)

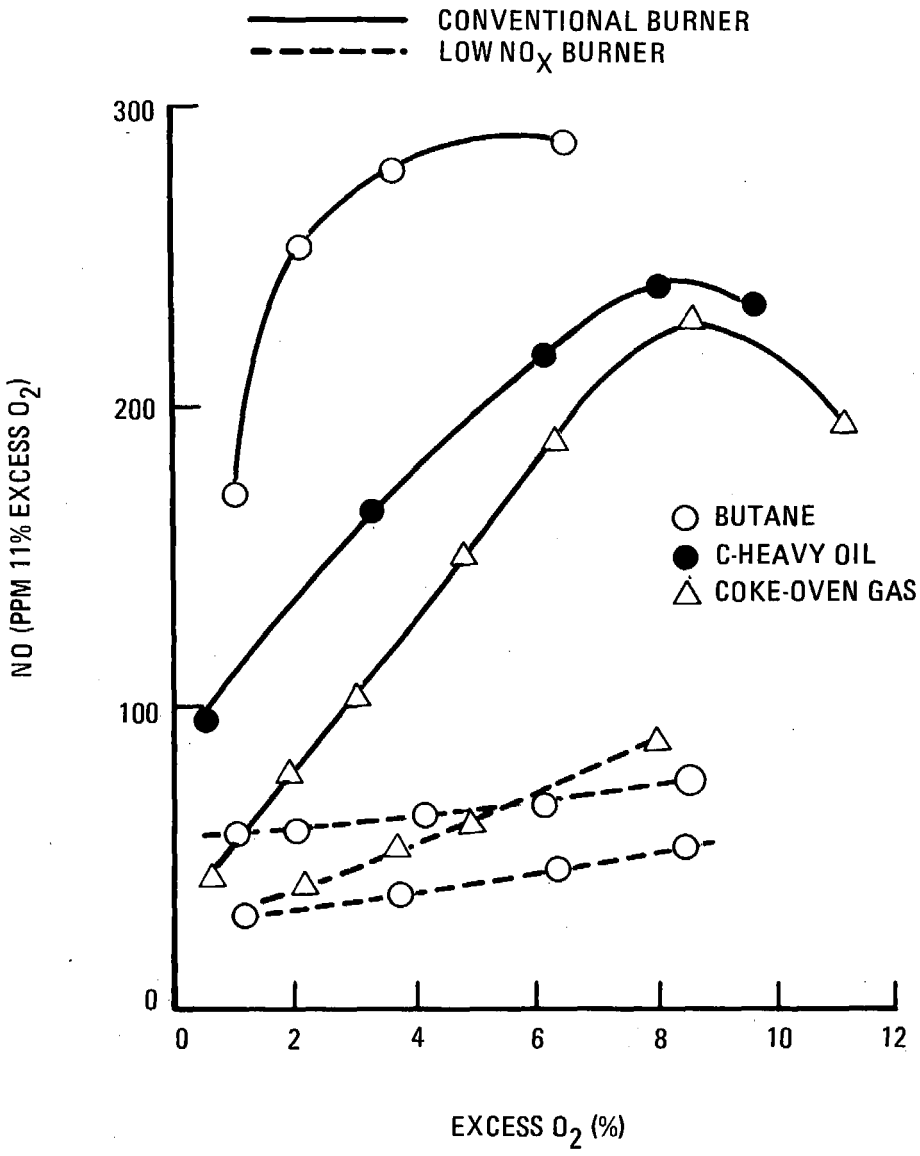


Figure 3-34. Effect of excess air on NO emission (Ref. 3-45)

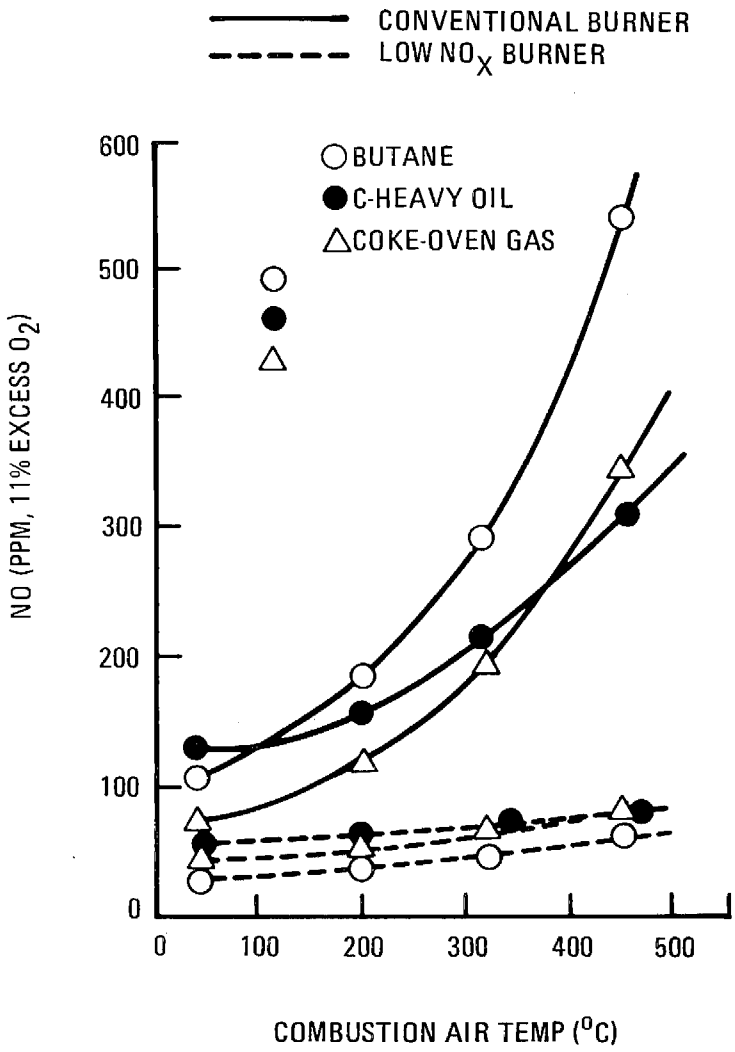


Figure 3-35. Effect of combustion air temperature on NO emission (Ref. 3-45)

TABLE 3-28. THERMAL $DeNO_x$ INSTALLATION SUMMARY[a]

Unit Ref. No.	Emission Source[b]	Design Heat Release 10^6BTU/hr	Load Point % of Design	Initial NO_x Conc., ppm	NH_3/NO_x Mole Ratio	H_2/NH_3 Mole Ratio	% NO_x Reduction[c]
5, 6	Atmos/Vac Pipestills	515/190	95	130	2.0	0	63
8	Crude Heater	250	86	79	2.1	0.23	51
9	Crude Heater	250	85	85	1.9	0.26	53

[a] Ref. 3-46.

[b] Installations in Japan

[c] % O_2 and NH_3 breakthrough: Not reported.

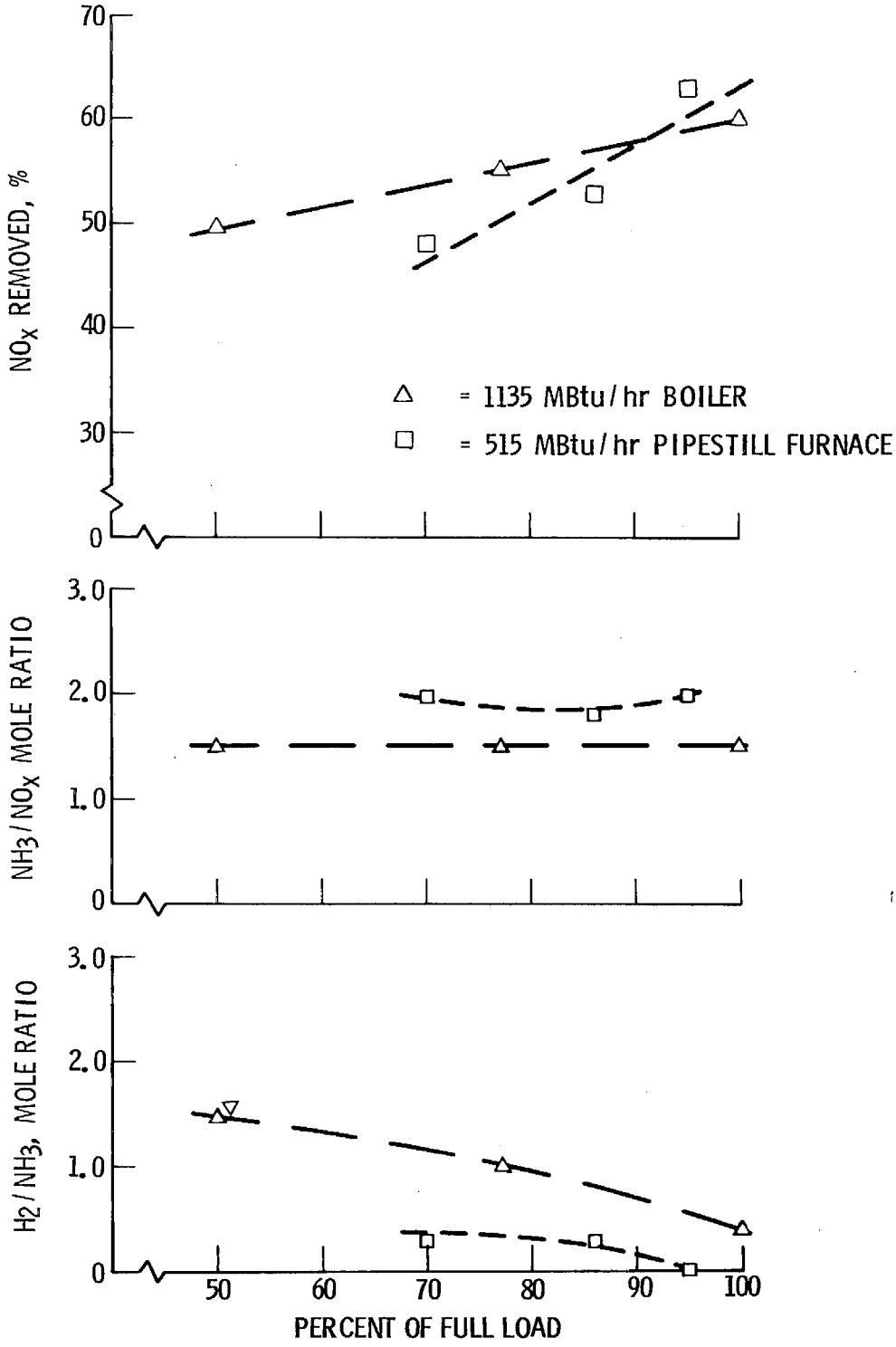


Figure 3-36. $DeNO_x$ performance versus unit load (Ref. 3-46)

residence times required for the reaction being in the range of 0.1 to 0.3 sec (Ref. 3-47).

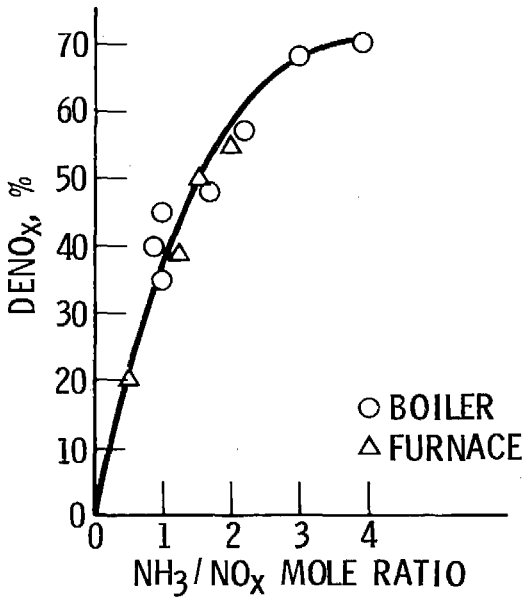
Use of this technique appears applicable to refinery heaters but would require determination of temperature profiles within the heater to locate appropriate temperatures and to ascertain the availability of adequate residence time. Based on limited information on refinery furnaces (Ref. 3-48), up to 70% reduction in NO_x was achieved with a NH_3-to-NO_x ratio in the range of 3 to 4 at a demonstration on a 500 MMBtu/hr gas-fired furnace at the Kawasaki Refinery of Tonen Petrochemical of Japan. However, the initial NO_x concentration and the amount of NH_3 emitted ("breakthrough") was not available. The effect of the NH_3/NO_x mole ratio is shown in Figure 3-37. It is likely that it is in excess of 50 ppm as evidenced by information. presented in Reference 3-15 and summarized in Figure 3-38.

While it appears that a reduction of 50% is feasible, its potential in reducing NO_x from any specific emission source is dependent on the source characteristics, e.g., gas temperature and residence time available, and also the allowable NH_3 emissions. Because of the relatively low temperatures defined for the process, the use of hydrogen may be required to attain the level of NO_x reduction desired.

3.5.3.3 Selective Catalytic Reduction

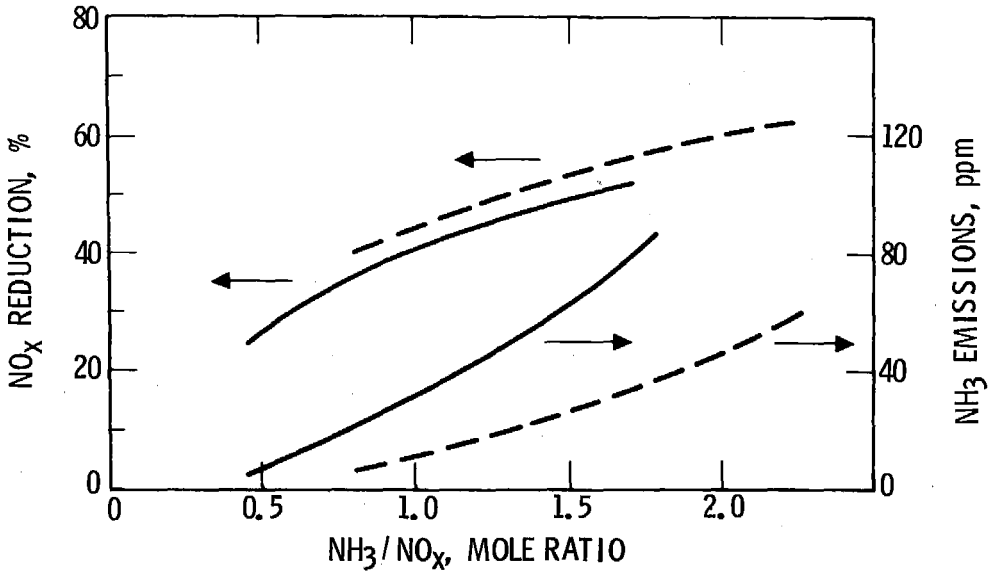
Because the heater burns refinery gas as fuel, the resultant flue gas is virtually free of particulates and contains no SO_2. The exhaust temperature is 410°C, which is slightly above the 360°C for maximum (>90%) reduction. These conditions are conducive to application of SCR. Because the furnaces employ natural draft, a fan to overcome the resistance of the SCR reactor and ducting is required.

Several alternatives are available to achieve 50% overall reduction. All the gas can be treated to obtain 50% removal, or 55.5% can be treated to remove 90%, with the remainder bypassed. These are depicted in Table 3-29. Because a larger space velocity can be used for 50% reduction (with all the flue gas treated), the catalyst bed is somewhat smaller, 32.6 ft^3 vs 45.3 ft^3.



- TESTS PERFORMED AT KAWASAKI REFINERY OF EXXON's JAPANESE AFFILIATE, TONEN PETROCHEMICAL K.K.
- RETROFIT ON UNITS AS FOUND BY THE REGULAR OPERATING STAFF WITHOUT INTERRUPTION OF NORMAL OPERATING SCHEDULE
- DEMONSTRATION BOILER: OIL FIRED, 140×10^6 Btu/hr, INTERMITTENT USE
- DEMONSTRATION FURNACE: GAS FIRED, 500×10^6 Btu/hr, CONTINUOUS USE

Figure 3-37. Thermal $DeNO_x$ commercial test data (Ref. 3-48)



	LOAD, %	GRID	TEMP, °C	EXCESS O_2, %	INITIAL NO_X (4% O_2), ppm
———	100	HIGH	970	2.5	135
– – – –	50	LOW	940	5.0	160

Figure 3-38. Example of effects of load and NH_3 injection location on NO_x reduction for 156 MW boiler (Ref. 3-48)

TABLE 3-29. CATALYST BED SIZE FOR 50 PERCENT NO_x REDUCTION

Case	% Removal in Reactor	Fraction Through Reactor[a]	$SV\ hr^{-1}$[b]	Cat. Vol. Ft^3	Catalyst Bed Dimensions, Ft
1	90	0.555	10,000	45.3	2.75 x 2.75 x 5.9
2	50	1.000	25,000	32.6	3.7 x 3.7 x 2.4

[a] Superficial velocity = 5 m/sec = 16.5 ft/sec.

[b] See Figure 3-26.

3.6 CEMENT KILN

A cement kiln using the dry process, operated by the California Portland Cement Company (CPCC), Colton, California, was designated for this study. The entire plant, consisting of two rotary kilns, is capable of producing 1200 tons/day of product (Refs. 3-49 and 3-50), but is regularly operated at about 90% of full capacity (1100 tons/day).

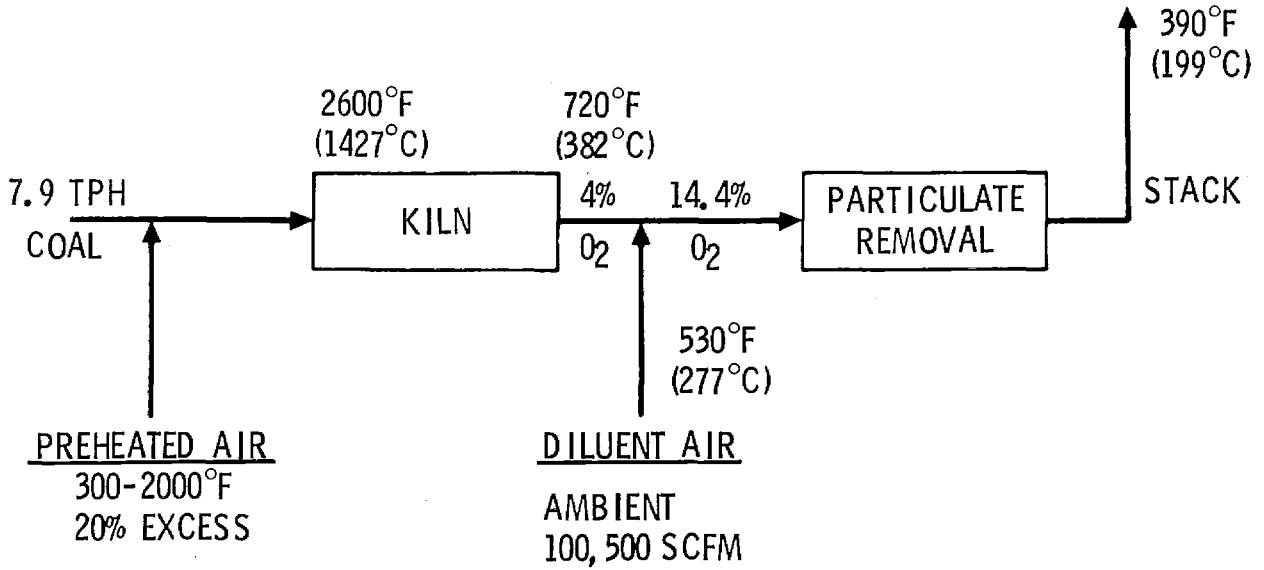
3.6.1 Cement Plant Characteristics

Firing is accomplished by a single horizontally configured burner operating at 20% excess air at the end of each 13 × 15 × 490 ft rotary kiln. Kiln feedstock, which is ground, sized, and homogenized, is converted to clinker in each unit at approximately 2600°F. The clinker is then discharged into air-quench coolers, where the hot cooling air is recycled back to the combustion zone for use as secondary (combustion) air. The kiln feedstock runs counter-current to the combustion exhaust gases, which are diluted with ambient air outside the kiln and filtered through a particulate removal system (baghouse) prior to discharge to the atmosphere through a pair of stacks. Flue gas temperatures average approximately 300°F (199°C), and temperatures throughout most of the kiln generally do not exceed 2600°F (1427°C). Gas temperatures exiting the kiln are about 720°F (382°C) and are cooled to 530°F (277°C) by diluting with ambient air prior to entering the baghouse.

Coal is the principal fuel used at the facility although oil may be used on a limited basis. For this study, only coal was considered. Facility data (Appendix A) indicate that the coal is a Utah low-sulfur type, with 0.35% sulfur, 1.5% fuel-bound nitrogen, and 12.0% ash, with a gross heating value of 11,400 Btu/lb. The coal firing rate is 7.9 tons/hr (TPH), or 180 MM Btu/hr; however, at full capacity (1200 TPH of product), approximately 10.2 TPH of coal is consumed (Ref. 3-49).

3.6.2 Emissions

With coal, NOx emissions per kiln are approximately 274 ppm (dry, 3% O_2) or 146 lb/hr (as NO_2), at a firing rate of 7.9 TPH. A schematic is presented in Figure 3-39, which was derived from CPCC facility data, Appendix A. Total exhaust gas flow rate is 249,000 ACFM at 390°F (199°C).



FROM KILN

43,500 SCFM
146 lb/hr NO_x (as NO_2)
11 lb/hr SO_2

TO STACK

144,000 SCFM EXHAUST GAS
146 lb/hr NO_x (as NO_2)
274 ppm NO_x (3% O_2, dry)
11 lb/hr SO_2 (7 ppm)
0.01 GRAINS/SCF (22.7 mg/NM^3) PARTICULATES

Figure 3-39. Cement kiln emissions characteristics

Glass fabric filters (baghouse) and multicones are used to control particulate emissions. Facility data indicate that the particulate emissions are approximately 0.01 grains/SCF (22.7 mg/Nm_3). Alkalinity of the particulates is extremely high. Calcium content in excess of 80% has been reported (Ref. 3-20).

Sulfur dioxide emissions are very low, 7 ppm (11.1 lb/hr), because the calcining step of the process acts in a manner analogous to a dry SO_2 scrubber, removing a substantial portion, approximately 90% of the SO_2 that would have been emitted in the flue gas.

Industry-wide emissions for cement kilns range from 250 to 800 ppm for NO_x and from 0 to 800 ppm for SO_2.

3.6.3 NO_x Emission Control Alternatives

Discussions with the Nippon Cement Company in Japan indicated that in that country the schedule for compliance with emission regulation is such that it is pratical to phase out conventional dry and wet kilns and replace them with a new suspension preheater (SP) kiln system (Ref. 3-51). This will be done for a number of reasons: (1) it will meet NO_x emission regulations; and (2) it significantly improves thermal efficiencies over the conventional kilns. It is estimated that the new kiln will reduce emissions approximately 50% relative to standard dry kilns. In correspondence with a U.S. licensee of a Japanese new reinforced suspension preheater (RSP) kiln (Refs. 3-52 and 3-53), it was indicated that the conversion of existing units to an RSP is very costly. Therefore, it is appropriate that the alternative control methods such as those being evaluated in this study be examined.

In this study, the use of CM, selective $DeNO_x$, and SCR are assessed for their potential to reduce NO_x emissions by 90% from the cement kiln.

3.6.3.1 Combustion Modifications

Results of tests conducted in Japan (Ref. 3-54), indicate that 50% reduction was achieved by employing LNB with a dual flow register in a 265 MW pulverized coal-fired boiler. NO_x was reduced from 500 to 250 ppm using the LNB. There is the potential of further significant reductions with the use of an optimum configuration LNB plus FGR. It is estimated that the burners have

a heat release capability of 100 to 150 MM Btu/hr, which is comparable to the kiln requirement. However, caution in extending such boiler experience to the cement industry is advised. In discussions with Nippon Cement (Ref. 3-51), they indicated that in staged combustion tests conducted for purposes of improved thermal efficiency, they found that cement quality was affected and a problem coating was formed on the kiln wall.

Kawasaki Heavy Industries conducted a number of LNB tests, firing pulverized coal at a rate of 52.4 lb/hr (23.8 kg/hr). One configuration provided a 61% reduction, 225 ppm (6% O_2), relative to a conventional burner emitting 575 ppm (Ref. 3-55).

It appears that the potential exists for a 50% reduction in NO_x emissions with the adaption of such an LNB. Further reductions may be possible with the use of FGR and staged combustion, if adaptable to the kiln. An LNB by itself or in combination with other CM will not attain the desired 90% reduction. However, the LNB (50%) in combination with partial (80%) SCR may be feasible technically). Also, combinations of thermal $DeNO_x$ (50% reduction) and a highly efficient LNB with CM may hold some promise.

Data on CM for a dry process using coal were not available. An indication of some effects of reducing combustion air temperature and excess oxygen content was given in Reference 3-20. In those tests, on a kiln employing the wet process and using natural gas, combustion air temperature was varied in an effort to evaluate the influence on NO_x emissions. However, combustion air temperature, kiln temperature, and exit gas O_2 concentration cannot be varied independently and maintain product integrity. As combustion air temperature increased, NO emissions increased for nearly constant O_2 concentration. The effect of kiln temperature is shown in Figure 3-40, which shows increasing NO emissions with increasing kiln temperature.

The data indicate that a given kiln operating temperature, for instance at 1755 K (2700°F), a reduction in NO emissions of 317 ppm or 14% was realized by reducing excess O_2 and increasing combustion air temperature to maintain kiln temperature.

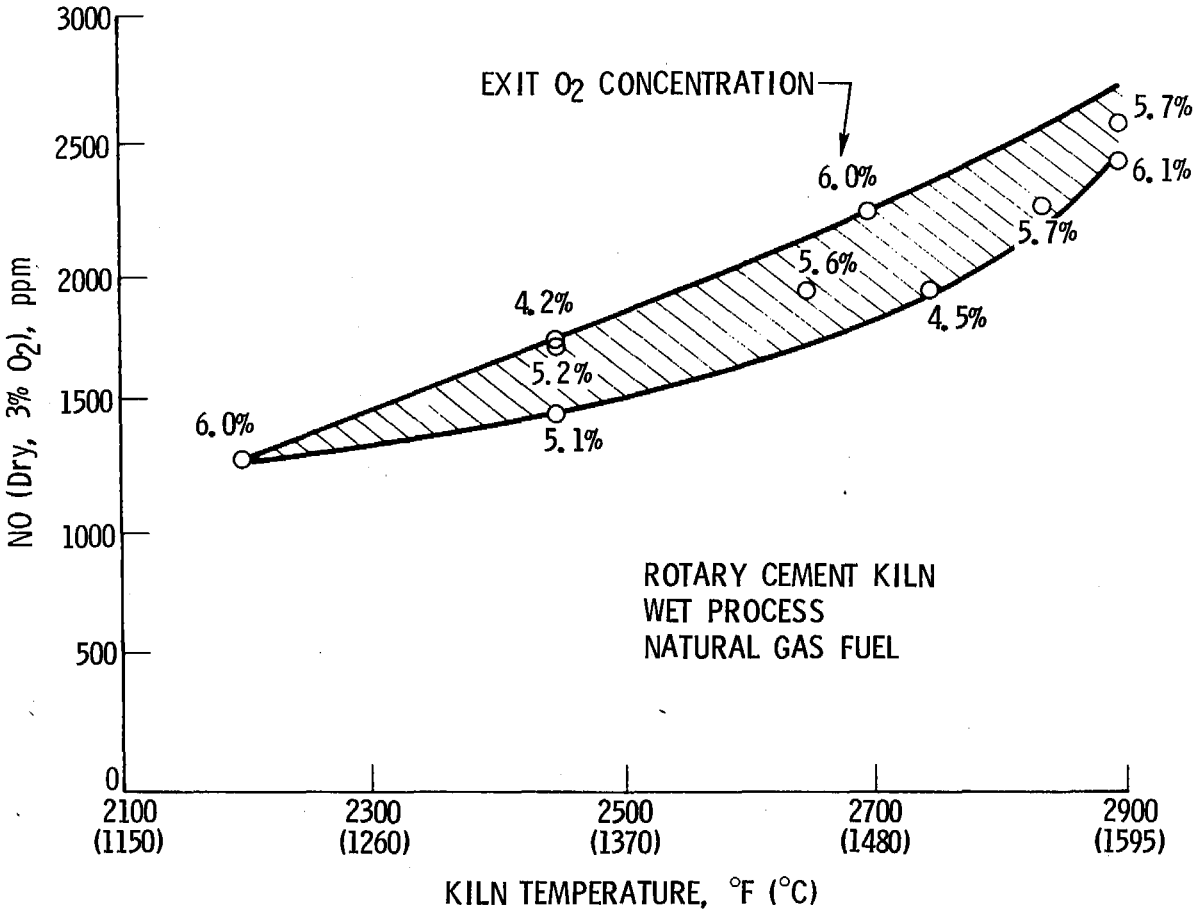


Figure 3-40. Effect of cement kiln temperatures on NO emissions (Ref. 3-18)

3.6.3.2 Thermal Denitrification

Applying thermal $DeNO_x$ between the kiln entrance and baghouse (Figure 3-41) may have the potential of a 50% reduction after reheating the exhaust gases to the required temperature at 1652°F (900°C). This reheating process would require approximately 3820 lb/hr of fuel oil and would generate an estimated additional 14 lb/hr of NO_x (as NO_2). The thermal input for the reheat represents about 40% of that used in the cement-making process.

The significant heat input resulting from reheat requirement increases the amount of air necessary to cool the exhaust gases to the 277°C temperature imposed by the baghouse to 650,000 SCFM. Without NO_x controls, the flue gas volume is 144,000 SCFM. This four-fold increase in gas volume would necessitate additional particulate removal equipment to handle the larger volumetric loads involved. In addition, duct diameters would increase significantly, mixing chambers would be required to achieve uniform thermal $DeNO_x$ temperatures and provisions made to assure the availability of the requisite residence time for the NO_x conversion reactions to take place. Since thermal $DeNO_x$ alone does not meet the 90% removal criterion and since significant modification to the particulate removal system and ductwork leading to it are required, this alternative was not considered any further.

3.6.3.3 Selective Catalytic Reduction

Based on catalyst information currently available, an SCR reactor can only be installed downstream of the particulate control equipment in order to attain a 90% overall NO_x reduction (Figure 3-42). Installation upstream of the baghouse would be unsatisfactory due to extremely high particulate loading, up to 1000 times the particulate outlet concentration (10 grains/SCF or 23,000 mg/Nm_3), and consequent clogging of the catalyst bed. Additionally, the liklihood is that the alkalinity of the flue gas caused by a high product concentration in the particulates would adversely affect catalyst performance.

Application of SCR downstream of the baghouse would necessitate reheating the gas 150°F (83°C) to reach 680°F (360°C), requiring 1415 lb/hr of fuel oil and generating approximately 5 lb/hr of NO_x. The total flue gas volume requiring treatment by the SCR unit including the gas from the reheat would be approximately 145,000 SCFM, an increase of only 0.7% relative to the

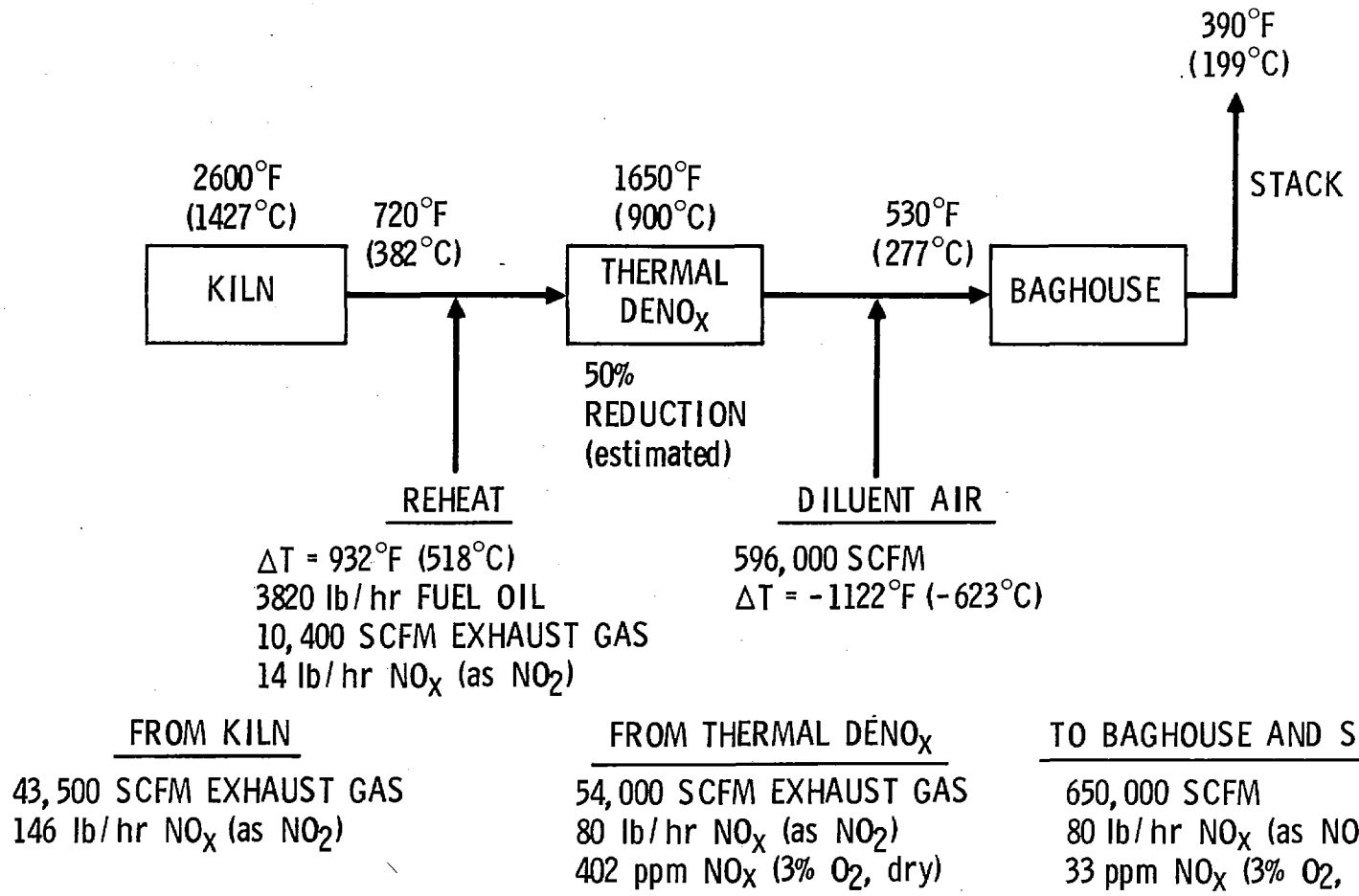


Figure 3-41. Application of thermal $DeNO_x$ to cement kiln

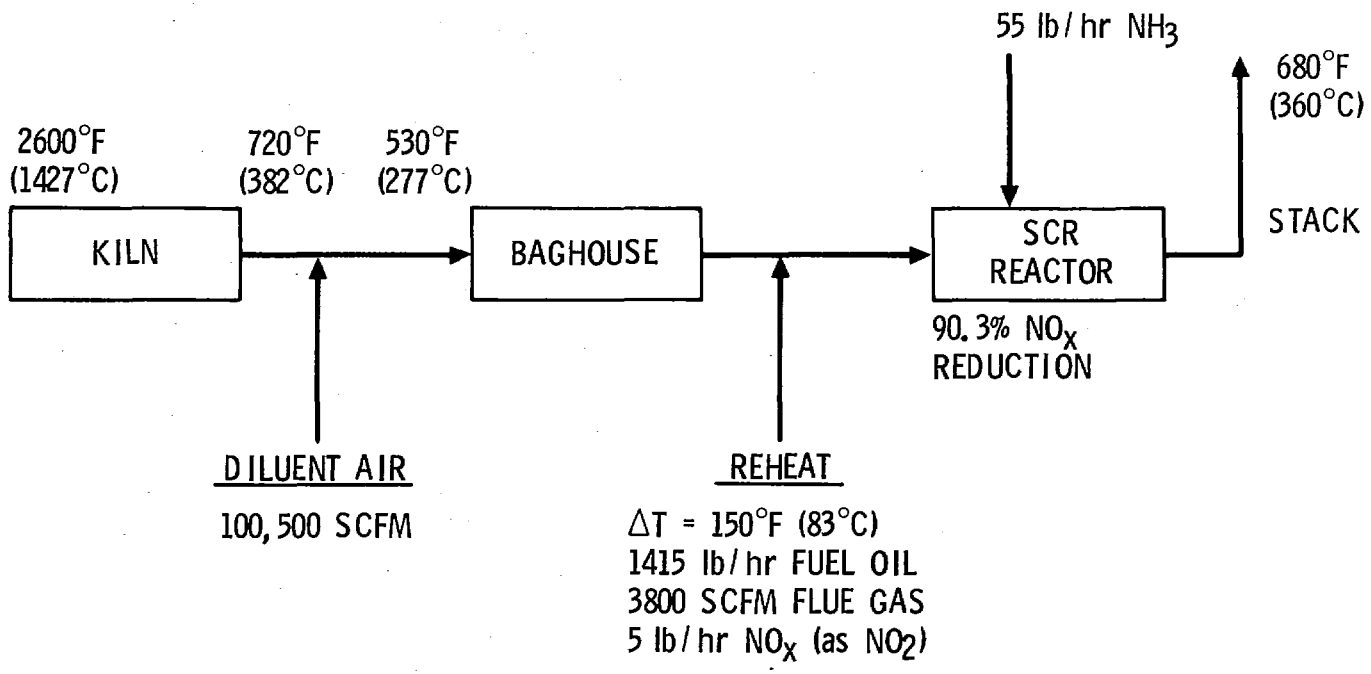
existing configuration. With 90.3% removal the resulting NO_x emissions existing, the stack would be approximately 27 ppm (dry, 3% O_2) or 14.6 lb/hr (as NO_2) (Figure 3-42).

If a suitable catalyst were available to handle the high particulate loads and alkalinity encountered in the coal-fired cement kiln, then an SCR unit could be placed downstream of the kiln (and upstream of the baghouse), with no reheat necessary because the temperature of 720°F (320°C) is appropriate for SCR. Estimating that the baghouse as currently installed at the CPCC facility operates at 99.9% removal efficiency, then the particulate loading in the flue gas prior to the baghouse would be on the order of 10 grains/SCF (22,700 mg/Nm3),* which is classified as an extremely dirty gas and generally cannot be considered for use with SCR without upstream particulate removal.

Recent Hitachi-Zosen pilot plant experience has shown the feasibility of applying SCR to two Japanese heavy oil-fired cement kilns (Ref. 3-56). Oil-fired kiln particulate loading in the flue gas are expected to be similar to those for coal-fired kilns since the major portion of the particulates are process-generated rather than of combustion origin (ash), 95% calcium at the kiln exit and 83% in the stream exiting the particulate control device (electrostatic precipitator) (Ref. 3-20).

In response to an inquiry about the application of SCR (90% NO_x removal) to a cement kiln with volumetric and emission characteristics similar to the CPCC facility (Refs. 3-57 and 3-58), a reactor 6.5 m (21.3 ft) x 7 m (23 ft) x 11 m (36 ft) in height with a honeycomb catalyst was defined for installation downstream of the particulate control device. Estimates indicate that the catalyst bed space velocity for this configuration is approximately 2500 hr^{-1} and a superficial velocity of 5.1 ft/sec (1.6 m/s) for a bed about 21 × 23 × 7.3 ft. This is consistent with 90% NO_x removal from dirty gases (Table 3-21). Figure 3-43 illustrates a typical reactor schematic which has been recommended by Hitachi Zosen to treat dust-laden gases as in coal-fired

*For a wet process, particulate concentrations of 22,000 and 19,000 mg/Nm^3 were reported (Ref. 3-20).



FROM KILN

43,500 SCFM EXHAUST GAS
146 lb/hr NO_x (as NO_2)

FROM BAGHOUSE

144,000 SCFM EXHAUST GAS
146 lb/hr NO_x (as NO_2)
274 ppm NO_x (3% O_2, dry)

FROM SCR TO STACK

147,800 SCFM EXHAUST GAS
14.6 lb/hr NO_x (as NO_2)
26 ppm NO_x (3% O_2, dry)
0.01 GRAINS/SCF (22.7 mg/Nm^3)
PARTICULATES

Figure 3-42. Application of selective catalytic reduction to cement kiln

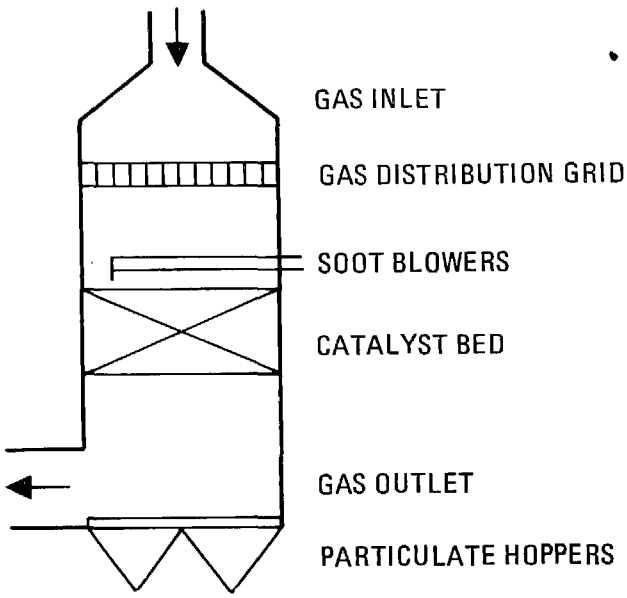


Figure 3-43. Potential Selective catalytic reduction reactor configuration for cement kiln application (Ref. 3-56)

boiler applications (Ref. 3-56). Periodic soot blowing is required and water washing of the catalyst (1 to 2 times annually) was recommended.

3.7 GLASS MELTING FURNACE

The glass furnace reviewed for applicability of NO_x controls was the PPG Industries unit, Fresno, California. In the flat glass process a batch of premixed materials that includes sand, soda ash, limestone, and dolomite are fed into the melting furnace tank. The charge also includes cullet, which is cleaned and crushed glass recovered from previous glass-making operations. The charge is melted in the furnace at temperatures in excess of 2800°F (1540°C). Natural gas is burned to provide the appropriate thermal conditions in the furnace.

At the exit of the furnace, there is a bath consisting of a molten tin layer that forms a flat surface onto which the molten glass flows. In the float bath, the glass ribbon exiting the melting furnace is maintained in a chemically controlled atmosphere to allow irregularities to be smoothed out and the surfaces to become flat and parallel. The glass ribbon then enters an annealing oven where the temperature is lowered gradually and leaves it at

approximately room temperature. The glass is then inspected, cut tempered as required, and prepared for shipment.

3.7.1 Furnace Characteristics

A schematic of the melting furnace is shown in Figure 3-44. This furnace has two pairs of primary and secondary regenerators (checkers) that are used to recover heat from the flue gas prior to entering the stack. The regenerators are filled with a refractory brick work and operated on an alternating basis. While one pair of regenerators is being heated by the flue gases from the combustion, the other is preheating the combustion air. The flow of air and flue gas is reversed every 10 to 20 min.

Characteristics of the furnace are summarized in Figure 3-45. Operation of the process is continuous, with planned maintenance shutdown occurring every several years.

Approximate exhaust gas temperatures are as shown in Figure 3-44. The temperature of the flue gas exiting the primary regenerator is about 1700°F (930°C), and at the stack entry it is approximately 1000°F (540°C).

After mixing with about 2 parts of ambient air from the ejector, the flue gas exits the stack at about 450°F (230°C). A total of 95,000 SCFM exit the stack at approximately 450°F (230°C). Approximately two thirds of the total is contributed by the ejector air entering the stack, which also cools the 31,500 SCFM (1000°F) combustion gases. Based on burner operation average, oxygen concentration in the flue gas from the second regenerator is 1%, with a maximum concentration of 2%. Other characteristics such as fuel consumption, burner characteristics, and glass production rates are considered proprietary and were not available for this study. However, lack of these data does not affect an evaluation of NO_x control alternatives. Miscellaneous other information is provided in Appendix A.

Natural gas is used as fuel. Therefore, thermal NO_x is formed exclusively. The reported emissions are 220 lb/hr or 834 ppm (dry, 3% O_2) at the stack exit.

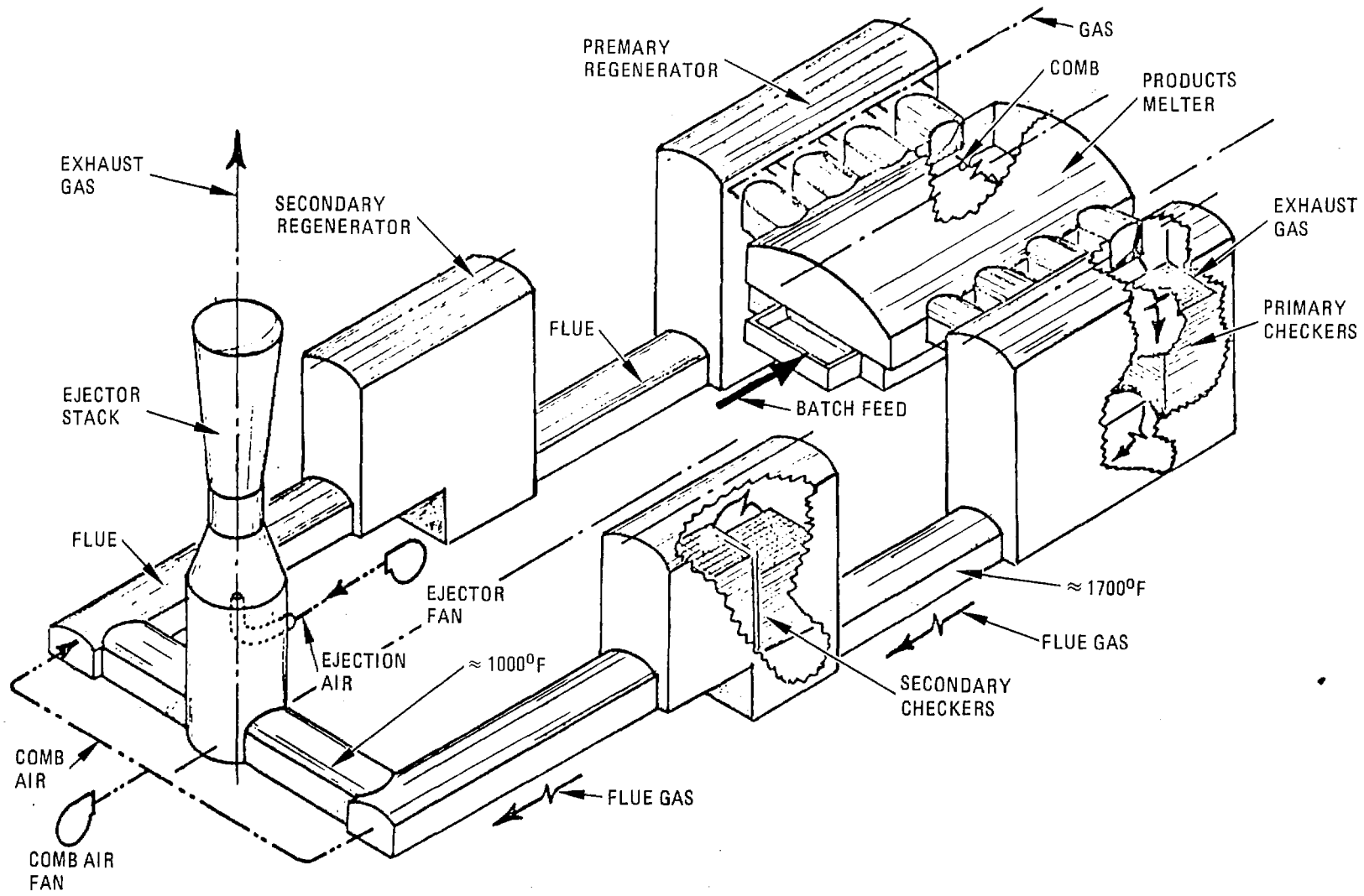


Figure 3-44. Float glass melting furnace (Ref. 3-60)

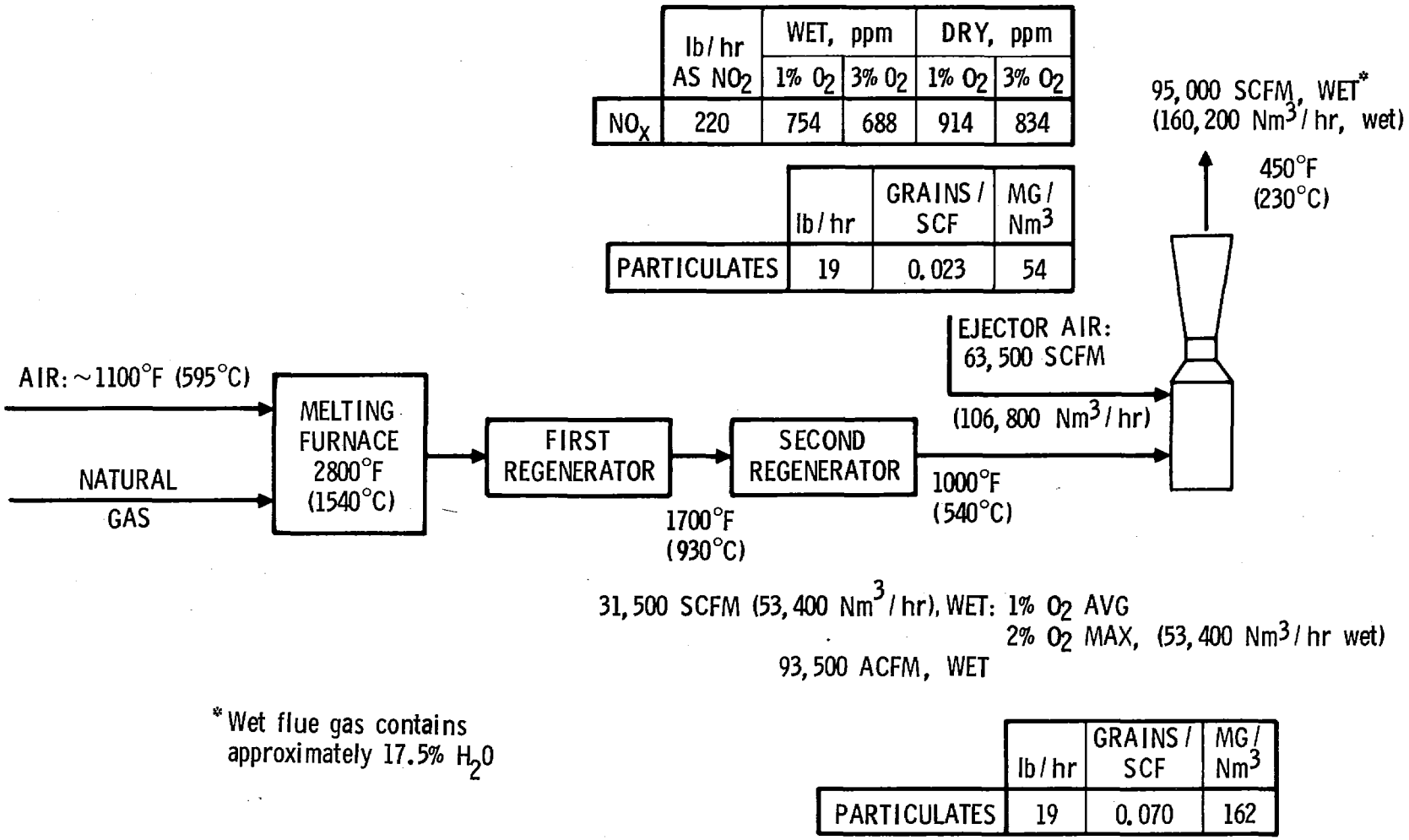


Figure 3-45. Glass melting furnace characteristics (Ref. 3-11)

3.7.2 NO_x Control Alternatives

The fuel burned is natural gas and is used exclusively in the furnace. Number 2 fuel oil is considered as an alternative for use on a standby basis.

The feasibility for reducing NO_x 90% based on current technology exists (Table 3-30). However, no full-scale application of control measures in this country or Japan have been reported. In Japan, some pilot-scale work on NO_x control measures using SCR has been conducted for this industry (Ref. 3-11), and will be discussed later. Also, NO_x reduction has been accomplished by supplementary heating of the glass in the furnace electrically (electric boosting), which lowers the peak flame temperatures and the fuel consumed, and thereby the amount of thermal NO_x formed (Ref. 3-59).

3.7.2.1 Combustion Modifications

It was reported (Ref. 3-60) that the quality of the glass is very sensitive to the characteristics and intensity of the flame and therefore could be affected by CM. Thus in addition to not having 90% reduction potential, CM could possibly have serious effects on the product. However, there appears to be some latitude in burner configuration and flame characteristics because number 2 fuel oil can be substituted for natural gas. Any potential to reduce NO_x emissions by reducing excess air must be considered low because the fuel in this furnace is burned on the average with only 5% excess air (1% excess O_2). Operation in the direction of a reducing atmosphere would affect product quality (Ref. 3-60). This low excess air condition is considered unique for this type of furnace; generally, 20% excess air is typical.

3.7.2.2 Thermal Denitrification

Thermal $DeNO_x$ has not been applied to glass furnaces in this country or Japan.

Examination of the temperature conditions (Figure 3-44), reveals that the flue gas temperature between the first and second regenerator is very nearly optimum 1700°F (930°C) for thermal $DeNO_x$. However, in addition to the potential complexities relative to the installation of an ammonia distribution and control system, a number of questions arise relative to the influence and

TABLE 3-30. NO_x CONTROL ASSESSMENT: GLASS FURNACE

NO_x Control Method	Control Technique Affected By	Estimated NO_x Reduction Potential	Remarks
Combustion modifications	Product quality sensitive to flame characteristics and intensity	25%	Operating with low excess air (5%). Burner modifications should not produce reducing atmosphere. Some flexibility may be possible as No. 2 oil can be used on standby basis.
Thermal $DeNO_x$	See Remarks	40 - 50%	Uncertainties related to moderate availability of adequate residence time for denitrification reaction to occur
Selective catalytic reduction	Deposition of sodium sulfate on catalyst affects its activity	90%	No full-scale installations are operational in this country or in Japan. Limited pilot testing has been conducted in Japan. A particulate/Na_2SO_4 removal system is required upstream of the SCR reactor to reduce loss of catalyst activity

effect on the second regenerator and the product. These include the question of adequate residence time for the reaction to occur, the effect of ammonia on the hot refractory material, and the effect on the glass of any residual NH_3 that may be swept back into the furnace with the combustion air on the reverse cycle.

In discussions with refractory applications personnel, it was determined that it was unlikely that ammonia in the concentrations involved, 10 to 20 ppm, for 1.0 to 1.5 NH_3/NO_x mole ratio would affect the brickwork in the secondary regenerator (Refs. 3-61 and 3-62). It is also unlikely that trace quantities of ammonia, if any, entrained in the combustion air would affect the glass quality.

On the basis of a gas temperature of 1700°F upstream of the secondary regenerator and a 3-sec stay time for the gas in the secondary regenerator with the flue gas exiting at 1000°F, there appears to be reasonable expectation that the residence time of less than 1 sec required for the thermal $DeNO_x$ reaction is available. However, testing would be required to determine if the appropriate temperature and mixing conditions are present. On the basis of these unknowns, the potential of achieving NO_x reductions in the range of 40 to 50% is considered moderately feasible. Also, it is understood that other glass furnace installations may not have a secondary regenerator. Therefore, the foregoing discussion may be limited in its potential applicability.

3.7.2.3 Selective Catalytic Reduction

The temperature of the flue gas prior to entering the stack is 1000°F (540°C), which is greater than the 350 to 400°C normally considered optimum for SCR. The temperature exiting the stack is 450°F (230°C). In neither case is the temperature compatible with the SCR process, necessitating cooling of the gases in the one case and reheating in the other.

The flue gas contains a relatively high solids loading of particules, 19 lb/hr or 0.070 grains/SCF (162 mg/Nm^3) prior to dilution by the ejector air. Its composition is estimated to contain 70 to 90% sodium sulfate. Deposition of the sodium sulfate on the catalyst lowers its activity and requires removal from the flue gas prior to its entry into the reactor, or the catalyst requires frequent washing to remove the sulfate (Ref. 3-11).

Asahi Glass Company, in 1975, began testing a prototype SCR installation with a capacity of treating 70,000 Nm^3/hr of flue gas from a glass melting furnace burning grade C heavy oil and containing 500 to 700 ppm NO and 300 to 500 ppm SO_2 (Ref. 3-11). A granular catalyst and intermittent moving bed developed by Asahi Glass was used.

Since the gas contained not only solid dust but also a considerable amount of sodium sulfate vapor, which tended to condense on and poison the catalyst, a catalyst carrier of sintered MgO (periclase) was used which is fairly resistant to poisoning. The carrier exhibited high strength and virtually no attrition in the moving bed. However, the granular catalyst has a relatively small surface area, resulting in a low space velocity value. By treating the gas with about 1.1 mole of NH_3 for each mole of NO_x, 90 to 95% of the NO was removed at 350 to 400°C at a space valocity of 1,500 to 2,000 hr^{-1}.

The catalyst was water-washed once every 1 to 2 months to remove sulfate and dust and was reused after the addition of the base metal catalyst component lost in the wash cycle. The life of the carrier was reported as about 1 year.

Although over 90% of NO_x was removed, the process was considered troublesome and costly due to the interference by the sodium vapor, which may not be avoided even with a parallel passage reactor of honeycomb type catalyst. There are no reported plans by Asahi to install a commercial plant for NO_x removal from flue gas from a glass melting furnace. No other details were available on the installation (Ref. 3-63).

In discussions with MHI (Ref. 3-15), they indicated that although they have developed and are applying SCR to various NO_x emission sources they were presently not involved in application of SCR for glass furnaces because of some question on its suitability and because of complexities arising from the nature of the exhaust gases.

There did not appear to be any activity or interest in glass industry applications other than that shown by Hitachi Zosen. Hitachi Zosen via Allis Chalmers (which is its U.S. affiliate for application of its SCR catalysts to pyrolysis processes) has responded to an inquiry regarding the application of SCR to control emissions from the glass furnace (Refs. 3-57 and 3-58).

Testing since 1975 was reported for a SCR honeycomb catalyst pilot installation on a glass melting furnace (Ref. 3-17). Heavy oil was burned and the SCR installation sized to handle 200 Nm^3/hr (120 SCFM). Other details were not available; however, it is believed that the space velocity was approximately 5,000 hr^{-1} and the catalyst required periodic "soot" blowing to remove the Na_2SO_4 particulates.

Hitachi-Zosen estimates the catalyst useful life as 1 year for the glass furnace application and expects an electrostatic precipitator to be installed upstream of the SCR system. Although the use of natural gas in the furnace being evaluated in this study produces virtually no sulfur oxides, a desulfurization facility would be required if SO_2 concentration were greater than approximately 200 ppm which could be expected if fuel oil were burned.

Considering a catalyst bed with a space velocity of 1500 to 2000 hr^{-1}, the dimensions would be 11.6 × 11.6 × 11.6 ft with a superficial velocity through the bed of 5.6 ft/sec (1.7 m/sec). It is estimated that a reactor with a catalyst bed of this size can be accommodated at PPG, Inc. However, the space available at this location may not be typical of other glass furnace installations. It is expected that the periodic water washing, or "soot blowing" of the catalyst would be accomplished during normal furnace operation with the flue gases temporarily bypassing the catalytic reactor. Annual catalyst replacement could be completed during a 1 to 2 week period when the furnace is idling at low capacity, thereby producing reduced emission levels and during which time the flue gases could bypass the reactor.

Kobe Steel has adapted its moving bed filter to be used with a catalyst for NO_x removal where the dust concentration is relatively high as in the case of a pelletizing plant in the steel industry. Data for a 1000 Nm^3/hr (590 SCFM) pilot plant is shown in Figures 3-46 and 3-47 (Ref. 3-64). The dust concentration is in the range of 200 to 400 mg/Nm^3. NO_x removal efficiencies of 90% at 350°C were reported for periods of about 2000 hr. A pressure drop of 50 mm H_2O (2 in. H_2O) was maintained. The effect of space velocity on bed pressure drop for two pellet catalyst sizes, i.e., 3 to 7-mm diam and 8 to 10 mm are shown in Figures 3-48 and 3-49.

An alternative to an electrostatic precipitator suggests itself in consideration of a moving bed filter upstream of the SCR reactor for the glass furnace application. Kobe Steel (Ref. 3-44) indicated their use of a moving bed filter to remove dust from cement kiln gases. Stone pellets, 1.625-mm diam, reduced dust, approximately 20 microns in diameter from 600 to 20 mg/Nm^3. Feasibility and development tests to determine the amount of total particulate and sodium sulfate removal would be required. If limestone were applied as successfully as the filter medium, it may be possible to use the spent material in the glass making process.

A schematic of an SCR instaallation is shown in Figure 3-50.

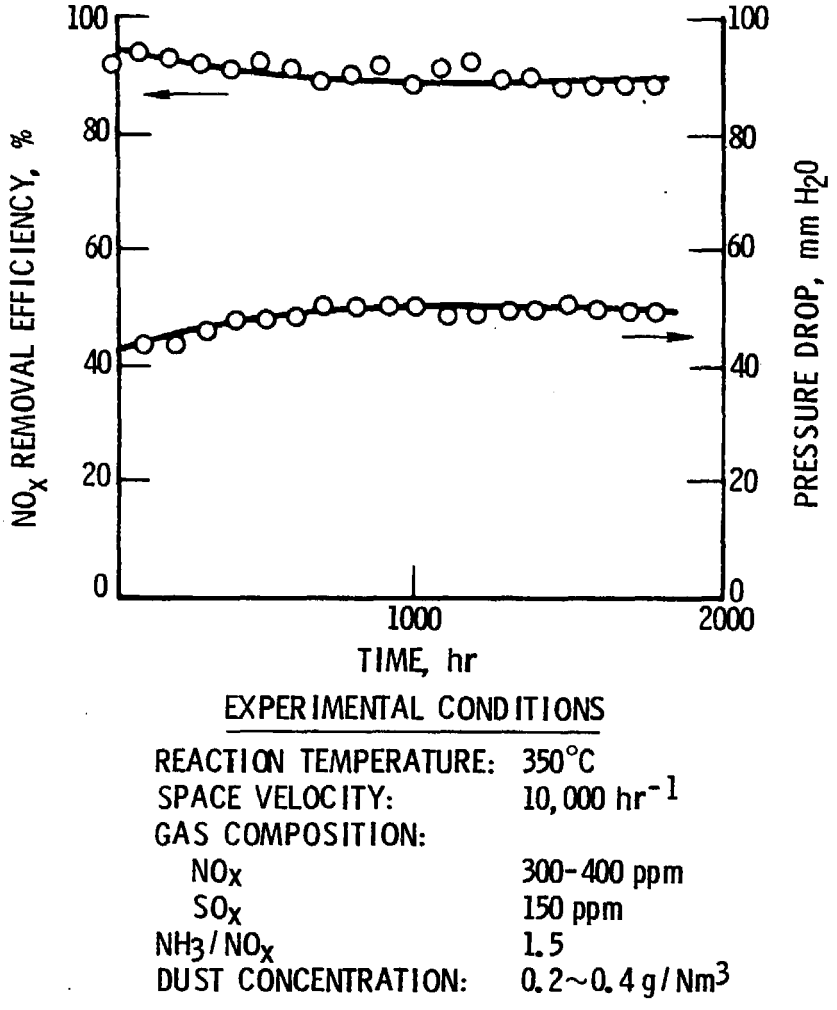
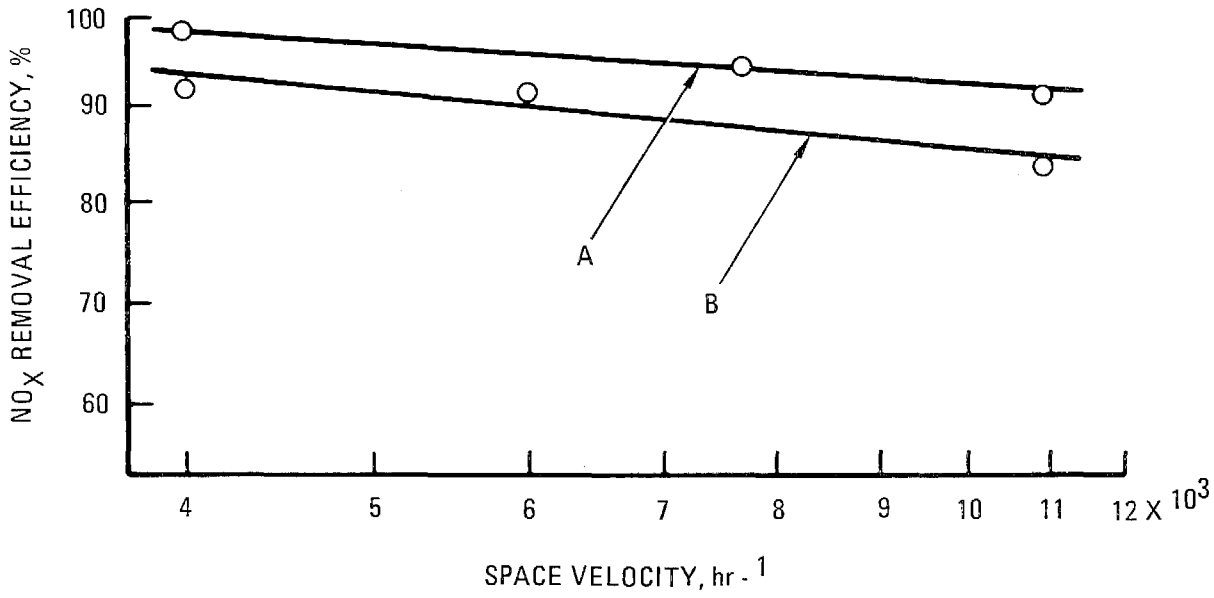


Figure 3-46. Effect of continuous operation on NO_x removal efficiency (Ref. 3-11)



EXPERIMENTAL CONDITIONS

REACTION TEMPERATURE: 350 °C

GAS COMPOSITION:

NO_X, 300 ~ 400 PPM

SO_X, 150 PPM

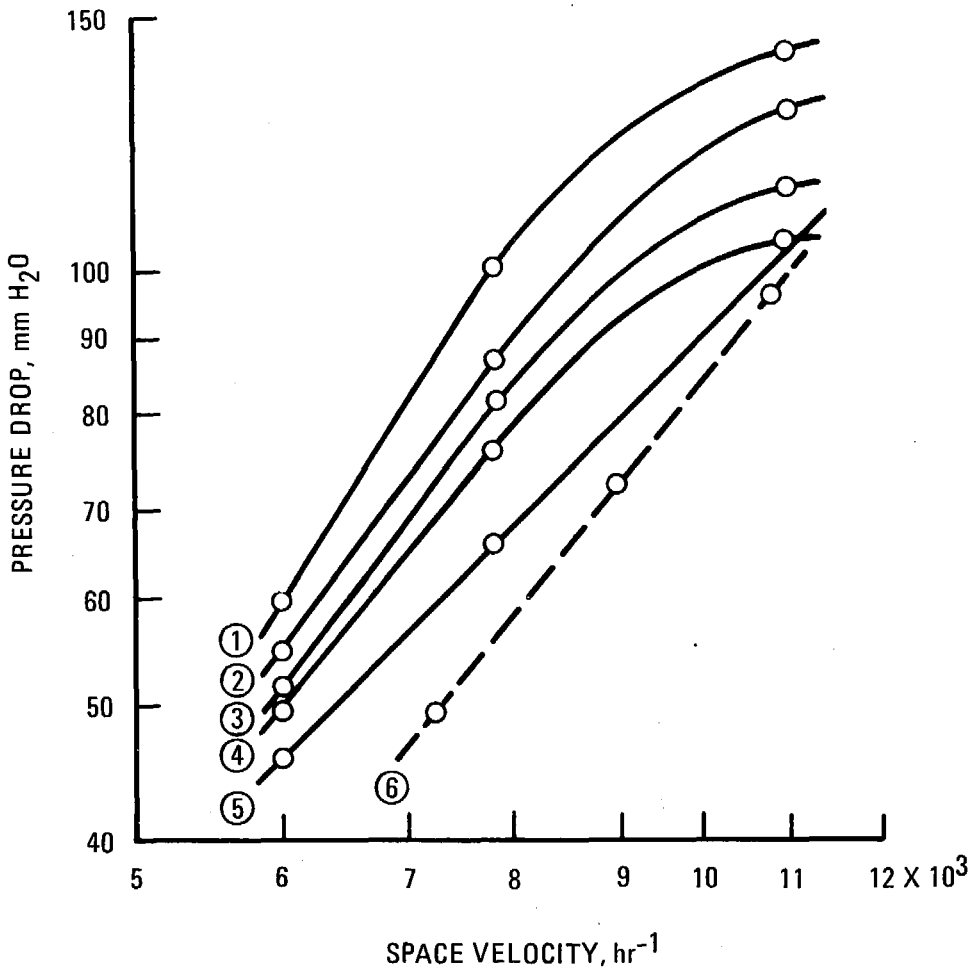
NH_3/NO_X, 1.5

CATALYST SIZE:

A, 3 ~ 7 MM

B, 8 ~ 10 MM

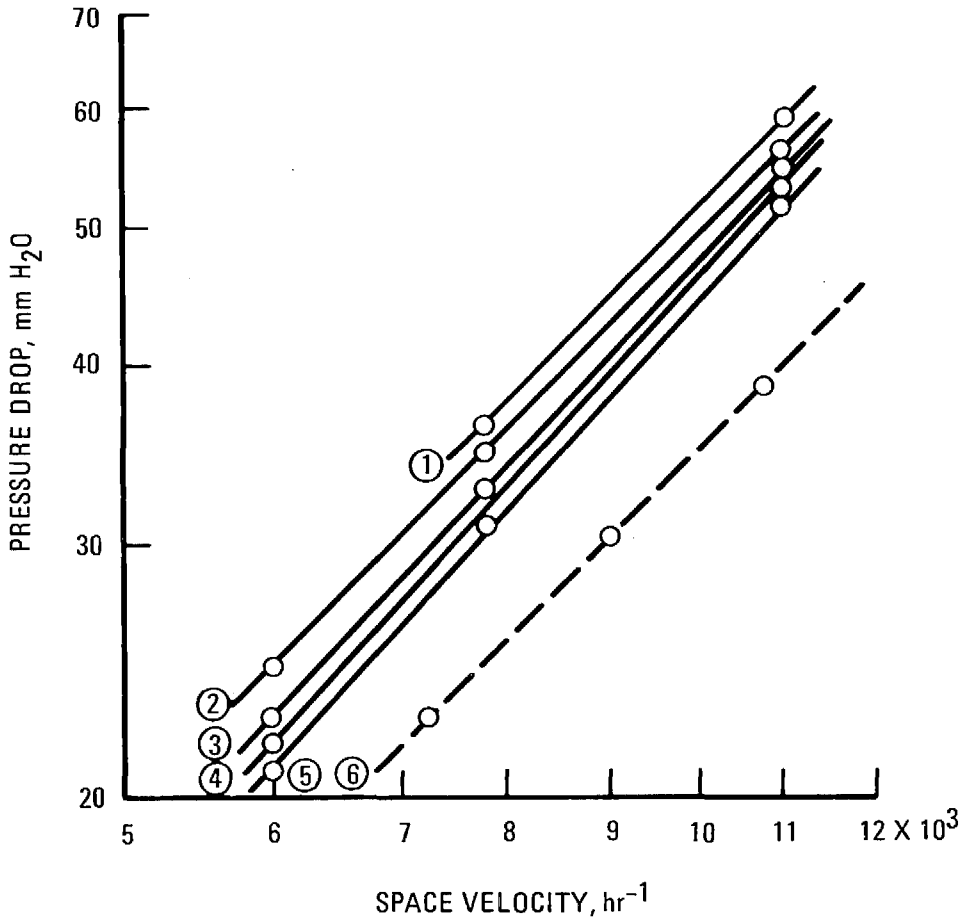
Figure 3-47. Relationship between the Kobe steel reactor NO_x removal efficiency and space velocity (Ref. 3-44)



EXPERIMENTAL CONDITIONS
REACTION TEMPERATURE: 350°C
DUST CONCENTRATION: 0.2 ~ 0.4 g / Nm^3
CATALYST SIZE: 3 ~ 7 mm
CATALYST MOVING VELOCITY:

① 12.0 ℓ/hr
② 13.0 ℓ/hr
③ 17.0 ℓ/hr
④ 26.0 ℓ/hr
⑤ 38.0 ℓ/hr
⑥ THEORETICAL VALUE (CLEAN GAS)

Figure 3-48. Relationship between space velocity and Kobe steel reactor pressure drop: small catalyst pellets (Ref. 3-44)



EXPERIMENTAL CONDITIONS
REACTION TEMPERATURE: 350°C
DUST CONCENTRATION: 0.2 ~ 0.4 g / Nm^3
CATALYST SIZE: 8 ~ 10 mm
CATALYST MOVING VELOCITY:

① 9.6 ℓ/hr	④ 25.6 ℓ/hr
② 11.2 ℓ/hr	⑤ 36.8 ℓ/hr
③ 16.0 ℓ/hr	⑥ THEORETICAL VALUE (CLEAN GAS)

Figure 3-49. Relationship between space velocity and Kobe steel reactor pressure drop: large catalyst pellets (Ref. 3-44)

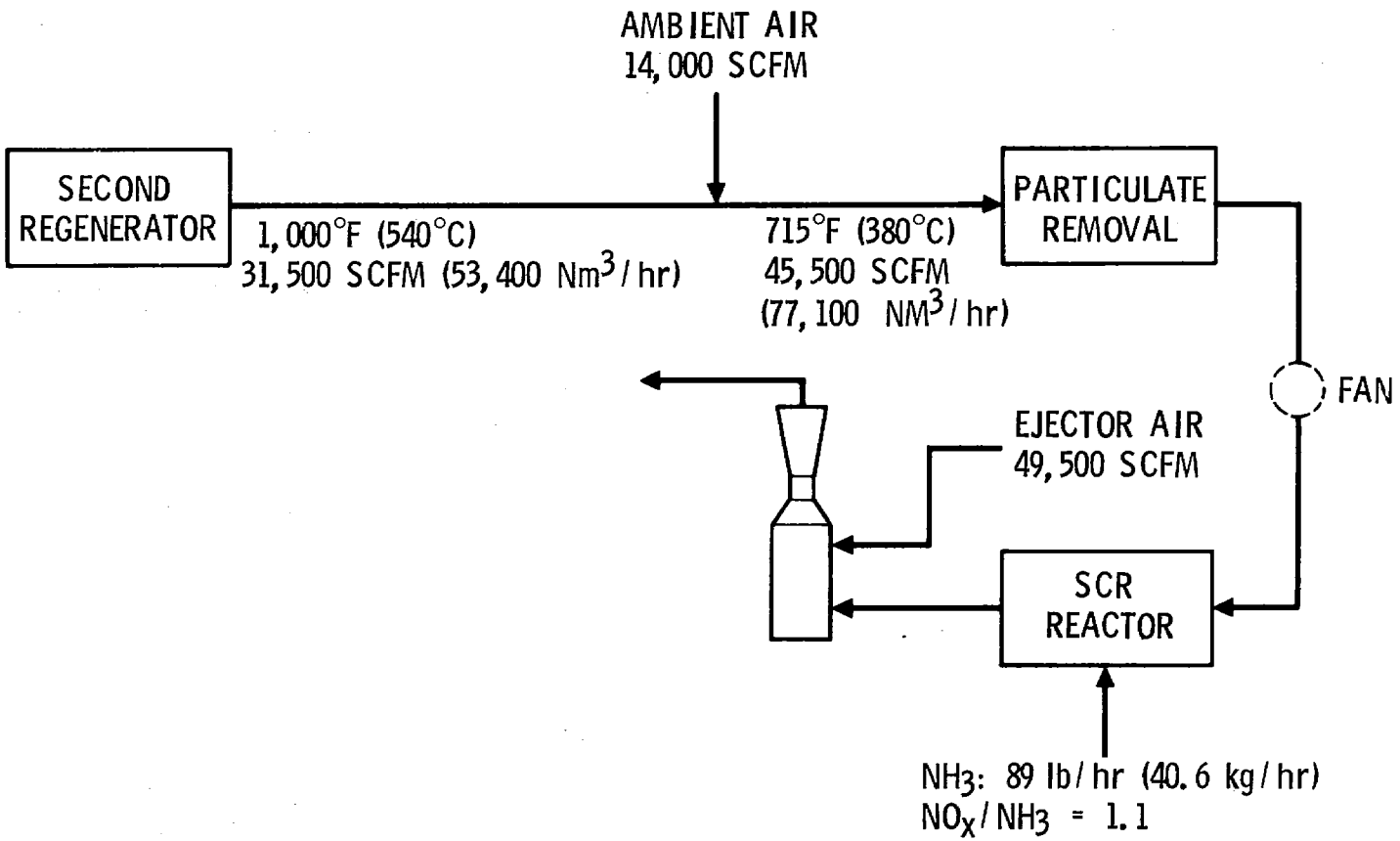


Figure 3-50. Flow diagram of selective catalytic reduction process adapted to glass melting furnace

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4. CONTROL SYSTEM COSTS

Capital and annual costs applicable to early 1979 were developed. Capital costs of SCR systems were based on generalized and budgetary costs provided by Japanese NO_x control system vendors and U.S. supplier costs (derived from Japanese sources). Thermal denitrification ($DeNO_x$) costs were based on information published by Exxon. Low NO_x burner (LNB) and combustion modification (CM) capital costs were based on U.S. vendor data and generalized cost data from Japanese sources. All these were adjusted to a common base (as described in subsequent paragraphs) to reflect plant facility investment and total capital investment, as well as retrofit factors. Annual costs include both operation and maintenance costs and annual charges on the total investment.

4.1 CAPITAL COSTS

Capital costs in this study consider not only the various pieces of equipment, foundations, ducts, controls, and piping, generally considered as battery limit costs, but also other direct facility investment costs. These encompass engineering design, supervision and fees, construction facilities, service facilities, initial charges, and startup and performance tests to define a direct plant facility investment (PFI) cost. Moreover, additional items such as licensing fees, sales tax, interest during construction, and working capital during construction are included to arrive at a total capital investment for an individual control system. Further, to reflect the effect of retrofitting, additional factors, generally 10% of PFI, were applied.

Plant facility and total investment costs are reported in the following paragraphs for installing LNB, thermal $DeNO_x$, and selective catalytic reduction (SCR) on the specific emission sources deemed to be feasible of meeting the NO_x reduction criteria established for this study (Table 4-1).

4.1.1 Low NO_x Burners

Capital costs of LNB for utility boilers have been reported by several sources (Refs. 4-1 and 4-2). These are summarized in Table 4-2, which

TABLE 4-1. CONTROL SYSTEM COST ESTIMATES

Stationary Source[a]	MW_e[b]	Hrs/ Yr[c]	Control Method[d]	NO_x Reduction Criteria, %	Reheat, bbl/hr[e]
Utility Boiler	175	0.45 (cap. fac.)	LNB only	27 to 55	0
			LNB + exist CM	7[f]	0
Stationary Turbine	121	100	SCR	90	0
Combined Cycle Turbine	236	1300	SCR	90	41.4
Internal Combustion Engine	1.6	1800	SCR	90	0
Oil Field Steam Generator	6.6	7000	Thermal	70 N[g]	0
				55 E[g]	0
			SCR	70 N	0.5
				55 E	0.5
Refinery Heater	6.8	7900	Thermal	50	0[h]
			SCR	50	0
Cement Kiln	18.9	7900	SCR	90	4.6
Glass Furnace	15.4	8760	SCR	90	0

[a] For those sources and control methods meeting reduction criteria defined for this study (Table 2-1)

[b] Megawatt equivalent to an electric power plant (based on thermal input to generate 1 MW of electricity).

[c] Equivalent hours at rated capacity.

[d] LNB = low NO_x burner, CM = combustion modification, SCR = selective catalytic reduction.

[e] Fuel oil.

[f] Incremental increase.

[g] N = new installation, E = existing installation.

[h] Assuming conditions exist for thermal $DENO_x$ applicability. Assumption based on crude heater and pipe still information (Ref. 4-4).

TABLE 4-2. REPORTED COMBUSTION MODIFICATION CAPITAL COSTS[a] -- RETROFIT, OIL-FIRED

Combustion Modification	U.S.[b]		Japan[c]	
	$/kW	Ratio	$/kW	Ratio
OSC	0.30	1.0	0.34	1.0
LNB	2.03[d]	6.8	1.70	5.0
TSC	-	-	3.81	11.2
LNB + TSC	-	-	6.29	18.5
TSC + FGR	5.71	19.0	6.22[e]	18.3
LNB + FGR	-	-	6.43	18.9
LNB + TSC + FGR	-	-	8.60	25.3

[a] 1977 dollars.

[b] Ref. 4-2 (single, wall-fired utility boiler).

[c] Ref. 4-1 (exchange rate 240 yen/$, 3,140 Nm^3 /hr/MW).

[d] Coal-fired burner.

[e] $3.88 for Chubu Electric, Chita Station 1-4 (Ref. 4-6).

also includes the cost of other CM, either singly or in combination with LNB. Generally more combinations were available from Japanese sources than U.S. In order to assess the applicability of the Japanese data to U.S. installations, ratios of CM cost (including LNB) to off-stoichiometric combustion (OSC) were calculated for both U.S. and Japanese data. The OSC costs for the U.S. and Japan are $0.30 and $0.34/kW, respectively. For the two instances where direct comparison was provided, viz., LNB and two-stage firing (TSC) plus flue gas recirculation (FGR), the ratios were relatively close to each other. This provided a measure of confidence when applying Japanese costs to the installation of LNB and other CM.

A comparison of LNB capital costs for various types of burner retrofits is shown in Table 4-3. For large forced draft burners, (40 to 100 MM Btu/hr), the cost estimates ranged from $26,000 to $60,000 per burner, depending on size and application.

For retrofitting a utility boiler, the hardware and installation labor cost was determined as $2.37/kW, or $26,000 per burner (early 1979 dollars) (Table 4-3). Considering retrofit design, contingencies, and other installation expense as an estimated 25%, the direct facility investment is $2.96/kW (Table 4-4). Additional items such as sales tax, interest during installation, and working capital result in a total capital investment of $3.18/kW, or $34,800 per burner for a 175 MW boiler with 16 burners.

4.1.2 Thermal Denitrification

Equipment costs for equipment to retrofit a 175 MW boiler were reported as $6.37/MW by Exxon (Ref. 4-4). It included those items shown in Table 4-5. This estimate is not representative of the total PFI and does not include a variety of other items such as engineering design and supervision, engineering fee, construction contingency, initial charges, and startup and performance costs (Ref. 4-5). The overall factor applied to the hardware and installation cost to obtain total facility cost was 1.74, which places the total plant facility investment at $11.11/kW (Table 4-6). Licensing fee for an oil field steam generator (62.5 MM Btu/hr thermal input) was reported as $20,000 (Ref. 4-4). Therefore, for equipment in this thermal capacity (also

TABLE 4-3. REPORTED LOW NO_x BURNER CAPITAL COSTS -- RETROFIT (EARLY 1979 DOLLARS)

	MW[e]	$/kW	No. of Burners	Input, MMBtu/hr	Burner Type	$/ Burner	$/MMBtu/ hr
Utility Boiler	175	2.37[a]	16	1650	FD[b]	26,000	251
Oil Field Steam Generator	6.6	5.15	1	6.25	FD	34,000[c,d]	544
Metal Heating Furnaces	3-30	4.27[e]	-	-	FD	-	-
Oil Heating Furnace	~26.5	11.74[f]	48	250	ND[b]	6,500	1,244
Oil Heating Furnace	~16	11.74[i]	4	150	FD	47,000	1,252

[a]Ref. 4-2. Costs esclated 7 and 9% for 1977 and 1978, respectively.

[b]FD = forced draft; ND = natural draft

[c]Ref. 4-3; 1st unit $50,000.

[d]Range of TRW & John Zink estimates $40,000 to $60,000 each (Ref. 4-7).

[e]Ref. 4-1.

[f]Ref. 4-1.

[g]Gaseous fuel; all others, oil.

[i]Self-recirculation type burner (Ref. 4-1).

TABLE 4-4. TOTAL CAPITAL INVESTMENT FOR LOW NO_x BURNER INSTALLATION FOR RETROFITTING 175 MW UTILITY BOILER

Item	$/kW
1. Direct facility investment	
1.1 Hardware, installation labor[a]	2.37
1.2 Engineering, field expense contingencies (25% of 1.1)[b]	.59
1.3 Subtotal, direct facility investment	2.96
2. Additional items	
2.1 Sales tax (2% of 1.3)[c]	.06
2.2 Interest during construction, 15% of (1.3 + 2.1) for 1 month)	.04
2.3 Working capital, 4% of 1.3[d]	.12
3. Total capital investment	$3.18

[a]Ref. Table 4-3.

[b]Estimate

[c]Ref. 4-8.

[d]Based on Ref. 4-9.

TABLE 4-5. COMPARISON OF EXXON AND DETAILED COST ANALYSES

Facility Cost Factors	
Exxon Cost Estimate Items[a]	Detailed Cost Analysis Items[b]
Hardware requirements	Hardware requirements
Installation labor and supervision	Installation labor and supervision
Construction field expense	Construction field expense
Contractor's fee	Contractor's fee
Construction facilities	Construction facilities
Service facilities	Service facilities
Utility facilities	Utility facilities
	Engineering design and supervision
	Engineering fee
	Construction contingency
	Initial charges (license costs)
	Startup and performance tests

[a]Refs. 4-4 and 4-5.

[b]Ref. 4-5.

TABLE 4-6. TOTAL CAPITAL INVESTMENT -- THERMAL $DeNO_x$ (EARLY 1979 DOLLARS)

Item	$/kW_e	%[a]	%[b]
1. Direct plant facility investment			
1.1 Hardware, construction labor[c]	6.37[d]	57.3[c]	42.0
1.2 Engineering design, construction field expense, construction facilities, contingency and other costs (see Table 4-4)	4.74	42.7[c]	31.3
1.3 Subtotal, direct plant facility investment	11.11	100.0	
2. Additional items			
2.1 Licensing fee	3.03[e]	27.3	20.0
2.2 Sales tax	.22	2.0[f]	1.5
2.3 Interest during construction 15%[g] of (1.3 + 2.1 + 2.2) for 2 months[h]	.36	3.2	2.3
2.4 Working capital 4%[g] of 1.3	.44	4.0	2.9
3. Total capital investment (new installation)	15.16	136.5	100.0
4. Total capital investment (Retrofit-based on 10% of 1.3)	16.33		

[a]Percent of direct facility investment

[b]Percent of total capital investment.

[c]Ref. 4-5.

[d]Based on an estimate of $1,115,000 for a 175 MW boiler (Ref. 4-4); also, $6.00/$kW_e$ (Ref. 4-11).

[e]Derived from $20,000 for an oil field generator (Ref. 4-10).

[f]California sales tax on material, estimated as 2% of 1.3 (Ref. 4-8).

[g]Based on Ref. 4-9.

[h]Estimated construction time, based on Ref. 4-4.

refinery heaters at 65 MM Btu/hr), the licensing fee translates into $3.03/kW. As was indicated in Section 3.8, thermal $DeNO_x$ has a potential for meeting the NO_x removal criteria for this study when applied to the oil field steam generator and refinery heater.

When the various other factors such as sales taxes, interest during construction, and working capital are considered, a total capital investment of $15.16 kW_e was calculated for new thermal $DeNO_x$ installations. Retrofit costs are obviously site specific; however, for a relatively uncomplicated installation, an additional 10% of the plant facility investment is estimated (Ref. 4-8) as an additional retrofit factor. For a retrofit of this nature, a total capital investment of $16.33/$kW_e$ was computed. For increasing degrees of installation complexity as could be experienced in space-limited situations, higher retrofit factors would be applied. "Total project costs" are reported as $12/kW in Reference 4-11. This is in agreement with the $15.16/$kW_e$ (for new installations) computed in this study if consideration is given to the additional $3.03/$kW_e$ licensing fee.

For this study, $15.16/$kW_e$ and $16.33/$kW_e$ are used as appropriate in determining annual costs for thermal $DeNO_x$ installations on the oil field steam generators and oil refinery heaters, where it was determined that the potential exists to meet the NO_x reduction criteria defined for this study.

4.1.3 Selective Catalytic Reduction

Average battery limit costs for five SCR installations located in Japan with three levels of complexity were reported in Reference 4-1. These included Configuration A, a reactor installation including ammonia storage, ammonia vaporizer, and electrical instrumentation for SCR. Configuration B includes A plus a fan and ducting, and Configuration C includes all of B plus a gas-gas heat exchanger and gas heater. The emission sources involved in this study and corresponding equipment configurations are shown in Table 4-7.

Using the cost data from Reference 4-1 with a yen-to-dollar conversion ratio of 240 (typical of late 1977), the dollar equivalent battery limit capital costs are shown in Table 4-8; costs are a direct yen-to-dollar conversion from Japanese cost estimates. This cost is considered reasonable

TABLE 4-7. SELECTIVE CATALYTIC REDUCTION INSTALLATIONS CONSIDERED FOR THIS STUDY

NO_x EMISSION SOURCE	MW_e	% NO_x REDUCTION	FAN	REHEAT	CONFIGURATION[a]
SIMPLE TURBINE	121	90	NO	NO	A
COMBINED CYCLE TURBINE	236	90	YES	YES	C
I.C. ENGINE	1.6	90	YES	NO	B
OIL FIELD STEAM GENERATOR	6.6	70/55[b]	YES	YES	C
REFINERY HEATER	6.8	50	YES	NO	B
CEMENT KILN	18.9	90	YES	YES	C
GLASS FURNACE	15.4	90	YES	NO	B

[a]BASIS OF REFERENCE 4-1 AND PARA 4.1.3

[b]70% FOR NEW INSTALLATIONS, 55% FOR EXISTING

TABLE 4-8. CAPITAL COSTS FOR SELECTIVE CATALYTIC REDUCTION INSTALLATIONS (EARLY 1979 DOLLARS)[a]

Emission Source	Config-uration[b]	MW_e[c]	Battery Limits: SCR Only, New Installation, $[d]	Battery Limits: $/kW[d]	Battery Limits: $/kW[e]	PFI, $/$kW_e$[f]	Retrofit Factor, %[g]	Retrofit PFI, $/$kW_e$	TCI, $/$kW_e$[h]	TCI,$ (000,000)
Simple Turbine	A	121	1.52×10^6	12.56	14.62	25.53	10	28.08	33.28	4.03
Combined Cycle Turbine	C	236	12.30×10^6	52.12	60.68	105.90	25	132.38	162.30	38.30
IC Engine	B	1.4	1.02×10^5	72.50	84.40	147.26	10	161.99	183.05	0.26
Oil Field Steam Generator	C	$6.6N^i$	9.61×10^5	145.61	169.51	296.16	0	N/A	334.67	2.21
		$6.6E^j$	8.17×10^5	123.79	144.10	251.46	10	276.61	312.56	2.06
Refinery Heater	B	6.8	1.93×10^5	28.38	33.04	57.75	10	63.52	73.56	0.50
Cement Kiln	C	18.9	2.31×10^6	122.22	142.28	248.38	10	273.22	323.76	6.12
Glass Furnace	B	15.4	5.33×10^5	34.61	40.29	70.26	25	313.56	371.57	5.72

[a] Japanese Sources (Ref. 4-1).

[b] Section 4.1.3

[c] Table 4-1.

[d] 1977 yen-to-dollar conversion rate: 240 yen/dollar

[e] Corrected for cost escalation 1979 (6.8%) and 1978 (9.0%)

[f] Total plant facility investment (PFI). Battery limits are 0.57 x PFI (estimate from Ref. 4-5)

[g] Percentage of PFI, estimated by author.

[h] Total capital investment (TCI) includes organization and start, 6% of PFI, cash during construction 4%, and interest during construction 15% per annum

[i] New, 70% reduction

[j] Retrofit, 55% reduction

on the basis that, in Reference 4-12, Japanese costs would require a 10% increase to convert them to U.S. Thus, a one-to-one ratio may be slightly high; but on the other hand, Reference 4-13 indicates that the ultimate U.S. cost of an Americanized design is approximately 85% of a Japanese commercial design for the size of installations considered in this study. Therefore, a one-to-one conversion from Japanese costs was considered appropriate for the purposes of this study.

As can be seen in Table 4-8, the unit capital costs in dollars per megawatt equivalent vary considerably, depending on size and complexity. Factors were also applied to update the battery costs to early 1979, to correct battery limit costs to PFI costs for retrofitting, and finally to determine the total capital investment.

Battery limit estimates were provided by various sources (Refs. 4-13, 4-14, and 4-15). These were extrapolated from utility boiler installations in the 100 to 240 MW range (Ref. 4-13), or estimated specifically in response to the author's inquiry (Refs. 4-14 and 4-15), and are reported in Table 4-9. In all cases battery limit estimates were adjusted by a multiplier of 1.75 (1/0.573 = 1.75), based on Reference 4-5 to obtain the PFI, which in turn was multiplied by 1.28 (Ref. 4-9) to account for additional items such as interest during construction. These are reflected in Tables 4-8 and 4-9.

It should be noted that the estimates of Table 4-9 are in most instances higher than those of Table 4-8. This is expected because those of Reference 4-10 include a fan, heater, and heat exchanger, which are not required for all installations (Table 4-7). In some instances where they were required and a minimal extrapolation was made, the estimates from both sources (Refs. 4-1 and 4-13) are in reasonable agreement. This is the case for the combined cycle turbine generator, where the estimates are $\$130/kW_e$ vs $\$108/kW_e$, based on References 4-1 and 4-13, respectively. In most cases, battery limit estimates from both sources were extrapolated from costs at the 67 to 200 MW_e levels to 1.6 to 18.9 MW_e on the basis of the size ratio to the 0.6 power. For instance, the 1.6 MW_e unit is approximately 10.6% $[(1.6/67)^{0.6} = 0.106)]$ of the cost of a 67 MW installation.

TABLE 4-9. SELECTIVE CATALYTIC REDUCTION TOTAL CAPITAL INVESTMENT COSTS -- U.S. SOURCES[a] (EARLY 1979 DOLLARS)

Source	MW_e	% NO_x Reduction	TCI, $ (000)	TCI, $/$kW_e$
Simple Turbine[b]	121	90	16,700	138.10
Combined Cycle Turbine[b]	236	90	25,500	108.05
IC Engine[b]	1.6	90	809	505.62
Oil Field Steam Generator[b]	6.6	70	1,510	229.24
	6.6	55	1,550	234.80
Refinery Heater[b]	6.8	50	1,480	217.94
Cement Kiln[c]	18.9	90	13,750	727.50
Glass Furnace[d]	15.4	90	8,300[e]	538.96[e]

[a]Refs. 4-13, 4-14, and 4-15.

[b]Based on installation size extrapolations from data in Ref. 4-10 (U.S. Economics, Japanese Design) includes ID fan and flue gas heater, 25% retrofit factor on plant facility investment (PFI). Total capital investment (TCI) is 1.28 x PFI.

[c]Based on Ref. 4-14. Battery limit costs include reactor, fan, ducts, heater, and heat exchanger. excludes civil work and erection. A factor of 1/0.57 = 1.75 applied to obtain PFI, which includes civil work and erection. TCI is 1.26 x PFI.

[d]Based on Ref. 4-15. Battery limit costs include reactor fan, heater, heat exchanger; excludes civil work and erection. A factor of 1/1.75 = 1.75 applied to obtain PFI, which includes civil work and erection. TCI is 1.28 PFI. Electrostatic precipitator (ESP) estimated seperately and included in PFI.

[e]ESP estimated as $\$3.45 \times 10^6$ ($\$224/kW_e$).

Since the estimates in Table 4-7 are more nearly tailored to what is required and are based on an average of five different control system suppliers, these were used as representative facility investment costs for the SCR installations considered in this study.

4.2 ANNUAL COSTS

The annual costs of the $DeNO_x$ system include operation and maintenance (O&M) and annual charges. Unit costs are shown in Table 4-10 and were applied as appropriate to the various control system estimates. The annual capital charges for a 20-year life were computed as 20% of the total capital investment based on a 10% interest rate (Table 4-11). The 20% rate includes taxes and insurance, which were approximately 5% of the total investment.

A summary of the various costs is shown in Table 4-12. Estimates for equipment outage during NO_x control system installation were not included. In most cases it should not exceed 3 months* as construction can go on during normal operation with a limited time required to actually place the control equipment on the line. Because of diversity of emission source characteristics, generalizations comparing the various installations are inappropriate. However, where reheating of the gas is required to achieve $DeNO_x$ temperatures, the cost of fuel is a significant fraction of the total, approximately 10% for most cases and as high as 30% for the cement kiln application.

In many cases, the O&M costs were 25 to 30% of the total. Where the reheat fuel requirements for SCR systems were high or in the case of thermal $DeNO_x$ where the capital investment was relatively lower, the O&M costs were approximately 50%. Also in situations where thermal $DeNO_x$ was considered capable of meeting the NO_x reduction criteria, as in the oil field steam generators and refinery heaters, the annual cost was significantly less than for SCR (Table 4-12).

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*This is based on information from References 4-16 and 4-17 where SCR systems were retrofitted on utility boilers during two periods totalling 2 to 3 months.

TABLE 4-10. ESTIMATED UNIT COSTS (EARLY 1979 DOLLARS)

Item	Unit Cost
1. Ammonia	$200/ton
2. Electric Power	$.045/kwh[a]
3. Water	$0.55/1000 gal.
4. Steam	$1.80/million btu ($1.8/1000 lb)
5. No. 2 Fuel Oil (for reheat)	$0.50/gal
6. Catalyst	$200/ft^3[b]
7. Maintenance, Labor and Material (Annual)	3% of Plant Facility Costs
8. Operating Labor	$16/hr
9. Annual Fixed Charges, 10% Interest, 20 Years	20% of Total Capital Investment (TCI)
10. Taxes, Insurance, Interim Replacement	Approximately 6.5% TCI (Included in Item 9)

[a]Assumed customer rate, therefore no replacement cost for power penalty.
[b]Based on anticipated short run U.S. price (0 to 2 years) (Ref. 4-13).

TABLE 4-11. AVERAGE CAPITAL CHARGE RATES

Interest rate, percent	Percent of total capital investment[a, b]					
	Life of equipment (yr)					
	5	10	15	20	25	30
6	31	21	18	16	15	14
8	33	23	20	18	17	16
10	35	25	22	20	19	18

[a]Straightline is assumed with no salvage value

[b]Taxes estimated 5 percent of investment cost annually

Note: Suggested methodology for cost analysis. State of California Air Resources Board, Request for Proposals, "Assessment of Control Technology for Stationary Sources," January 1978

TABLE 4-12. SUMMARY OF NO_x REMOVAL COSTS FOR SELECTED STATIONARY SOURCES

Stationary Source[a]	MW_c[b]	HRS/YR[c]	Control Method	NO_x Reduction Criterion, %	Total Capital Investment, $/$kW_e$[d,e]	Control Costs[f]		
						Unit Costs	$/lb NO_x Removed	$/MMBtu Input
Utility Boiler	175	0.45[g]	LNB only[h]	27-55	3.18	0.19 mills/kWh	0.076-0.155	0.020
			LNB+exist CM[h,i]	7[j]	3.18	0.19 mills/kWh	0.60	0.020
Stationary Turbine	121	100	SCR	90	33	76 mills/kWh	17.28	5.03
Combined Cycle Turbine	236	1300	SCR	90	162	29.5 mills/kWh	57.78	5.24
Internal Combustion Engine	1.4	1800	SCR	90	183	$0.019/hp-hr	1.04	2.98
Oil Field Steam Generator	6.6	7900	Thermal	70 N[k]	15[l]	$0.29/bbl[m]	0.30	0.096
				55 E[k]	16[l]	$0.27/bbl[m]	0.36	0.091
			SCR	70 N	335	$4.03/bbl[m]	4.03	1.30
				55 E	313	$3.78/bbl[m]	3.78	1.22
Refinery Heater	6.8	7900	Thermal[n]	50	16[l]	--[o]	1.58	0.092
			SCR	50	74	--	5.19	0.30
Cement Kiln	18.9	7900	SCR	90	324	$6.42/ton	2.40	1.75
Glass Furnace	15.4	8760	SCR	90	372[p]	--[o]	0.91	1.38

[a] For specific units identified in Tables 1-1 and 1-5.

[b] MW equivalent to an electric power plant (based on thermal input to generate 1 MW of electricity).

[c] Equivalent hours at rated capacity.

[d] Rounded off to nearest dollar except for low NO_x burner.

[e] Includes retrofit cost on plant facility investment, see Tables 4-2, 4-6, and 4-8.

[f] Cost related to facility; downtime (if any) not included.

[g] Capacity factor.

[h] Burner replacement only; O_2 control costs not included.

[i] Existing combustion modifications.

[j] Incremental removal relative to unmodified levels (25 ppm, or 55/hr).

[k] N = new installation, E = existing installation.

[l] Includes Exxon licensing fee, $3.00/kW.

[m] Per net barrel of oil recovered (2 bbl net per bbl consumed).

[n] Assuming conditions in furnance are applicable for thermal $DENO_x$ (must be determined experimentally for the unit).

[o] Product rate not available.

[p] Includes $224/kW for an electrostatic precipitator (ESP).

4.2.1 Utility Boiler

The impact of installing LNB on 175 MW oil-fired utility boilers is shown in Table 4-13 for installations with existing CM and units that would utilize LNB exclusively. There is a relatively small projected impact in NO_x reduction, 55 lb/hr, when LNB is added to units with existing CM when compared to the use of LNB exclusively, 213-435 lb/hr.

With a total capital investment of the burner installation of $556,500, or $34,800 per burner, the annual cost for the LNB installation is $130,200 (Table 4-14). The annual costs for the various LNB alternatives are shown in Table 4-15. Since the thermal input and electrical output are constant regardless of the configuration, the annual NO_x control cost is 0.19 mills/kWh, or $0.20/MM Btu heat input. The cost benefit noted is on the basis of the cost per pound of NO_x removed. In the case of the LNB installation, the cost ranges from $0.076 to 0.155/lb of NO_x removed. In combination with other existing CM, the cost of control increases by a factor of 4 to 8 to $0.600/lb of NO_x removed.

4.2.2 Simple Turbine

Ninety percent NO_x control with the simple turbine is feasible with SCR. The annual control cost for the SCR installation on the eight turbines is $914,500 (Table 4-16). Operating and maintenance costs represent approximately 14% of the annual cost, with the annualized charge on capital being the remainder. Because the temperature of the exhaust out of the turbine is within the limits for SCR, no reheat is required. Because of the low usage of the turbine, catalyst lifetime is expected to be high, and one replacement during the 20-year life is estimated. Annual costs are estimated as $.076/kWh, $17.28/lb NO_x removed, and $5.03/MM Btu heat input.

4.2.3 Combined Cycle Turbine

For 90% NO_x control with the combined cycle turbine, the SCR reactor must be located downstream of the waste heat boiler. This requires approximately 41 bbl/hr, or 108%, of the fuel to heat the exhaust gas to the appropriate SCR reaction temperature. Locating the reactor upstream of the

TABLE 4-13. IMPACT OF ADDING LOW NO_x BURNERS TO 175 MW OIL-FIRED UTILITY BOILERS

	AVERAGE REDUCTION, %	EMISSIONS AT MAX LOAD				NO_X REDUCTION,
		FROM		TO		
		PPM	LB/HR	PPM	LB/HR	LB/HR[a]
LNB	55 (MAX)[b]	360	790	162	355	435
LNB	27[c]	360	790	263	577	213
TSC[d] + LNB	27	288[d]	632	263	577	55
OSC + LNB	41	238[e]	522	213	467	55

[a] As NO_2

[b] Japanese data, 2 units (Ref. 4-18).

[c] Japanese data, 80 units (Ref. 4-1).

[d] 20% for TSC only

[e] 34% for OSC only

TABLE 4-14. ANNUAL COSTS FOR LOW NO_x BURNER RETROFIT ON 175 MW OIL-FIRED UTILITY BOILER (EARLY 1979 DOLLARS)

ITEM	COST FACTOR	COST, $
ANNUAL CHARGE	20% x $556,500[a]	111,300
OPERATING COST	17%[b] x $111,300	18,900
		130,200

[a]TOTAL CAPITAL INVESTMENT: 175,000 kW x $3.18/kW, ANNUAL RATE: REF.: SECTION 4.2

[b]REF. 4-2.

TABLE 4-15. LOW NO_x BURNER RETROFIT COST FOR 175 MW OIL-FIRED UTILITY BOILER

COMBUSTION MOD NO_x CONTROL METHOD	% NO_x REDUCTION, AVERAGE	ADD LNB, % REDUCTION, AVERAGE	ANNUAL COST[a]		
			MILLS/KWH[d]	$/LB NO_x	$/$10^6$ BTU INPUT
NONE	--	55[b]	0.19	0.076	0.020
NONE	--	27[c]	0.19	0.155	0.020
TSC	20	27	0.19	0.600	0.020
OSC	34	41	0.19	0.600	0.020

a CAPITAL COST: $3.18/KW

b 2 UNITS, Ref. 4-18

c 80 UNITS, Ref. 4-1

d 45% CAPACITY FACTOR

e NO_x REMOVED, AS NO_2

TABLE 4-16. NO_x CONTROL COSTS: SIMPLE TURBINE[a] -- SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 588 lb/hr (as NO_2)
= 415 ppm (dry, 3% O_2)
MW_e = 121

SCFM = 508,000 (wet)
hr/yr = 100
Operation: Full Load

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	12 tons/year	2.4	
2. Electric Power, ($0.045/kwh)	46,000 kwh	2.1	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3 (Atomization)	0.1	
5. Fuel Oil, ($0.50/gal)	Not required	--	
6. Catalyst,[b] (200/ft^3)	5.1 ft^3	1.0	
7. Maintenance, Material, Labor (3%)	$3.40 x 10^6	101.9	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	1.0	
9. Annual Charges, (20%)	$4.03 x 10^6	806.0	
10. Total	--	914.5	$17.28/lb NO_x abated[c]

[a]For units identified in Tables 1-1 and 1-5.

[b]Equivalent annual replacement. Because of relatively low operational use, catalyst life may exceed 20 years.

[c]90% NO_x reduction.

boiler is impractical. Because of the NO_x generated in the reheat combustion process, a NO_x removal of greater than 100% would be required to achieve a 90% reduction based on current emissions.

The annual control cost for the combined cycle installation is $10,073,000 (Table 4-17). The fuel cost represents about 11% of the total. The O&M cost (including fuel) is 26% of the annual cost. The catalyst lifetime is estimated as at least 2 years because of the low use factor of the installation. The annual cost of catalyst replacement representing about 2.6% of the total annual cost is the average value for biannual replacement. The annual cost of $10,073,000 represents 29.5 mills/kWh, $57.78/lb NO_x removed, and $5.24/MM Btu input.

4.2.4 Internal Combustion Engine

Use of SCR was determined to be feasible with the 200 HP natural gas-fueled IC engine. The annual control cost is $67,700 (Table 4-18). Operating and maintenance costs represent about 24% of the annual cost. Catalyst lifetime is expected to be at least 2 years because of the relatively low usage of the equipment. The $67,700 represents $0.019/BHP-hr, $1.04/lb NO_x removed, and $2.98/MM Btu heat input.

4.2.5 Oil Field Steam Generator

The 70 and 55% NO_x reduction for new and existing units can be met by SCR or thermal $DeNO_x$.

Annual costs of SCR for new and existing units are $575,800 and $538,800/year (Tables 4-19 and 4-20). Reheating of the exhaust downstream of the stack exit temperature requires 3800 bbl/yr. The cost of this fuel (which is recovered crude) was estimated at $12/bbl (early 1979 dollars) and represents about 8% of the annual control cost. Operating and maintenance costs are about 24% of the total. Unit costs are shown in Table 4-21 and represent approximately 13 times the corresponding control cost for thermal $DeNO_x$. This is one instance where it was feasible for the reduction criteria to be met by a method other than SCR, and in such a case annual costs for thermal $DeNO_x$ are significantly less, $42,400 and $40,200, respectively (Tables 4-22 and 4-23).

TABLE 4-17. NO_x CONTROL COSTS: COMBINED CYCLE TURBINE[a] -- SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 140 lb/hr (as NO_2)
= 76 ppm (dry, 3% O_2)
MW_e = 236

SCFM = 524,000 (wet)
HR/YR = 1300
Operation : Full Load

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	40 tons/year	8.0	
2. Electric Power, ($0.045/kwh)	1.17×10^6 kwh	52.5	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	0.2	
5. Fuel Oil, ($0.50/gal)	41.4 bbl/hr	1,131.0	
6. Catalyst,[b] ($200/ft^3)	1300 ft^3	260.0	
7. Maintenance, Material, Labor (3%)	$\$31.24 \times 10^6$	937.2	
8. Operating Labor, ($16/hr)	5×10^{-3} men/MW	24.5	
9. Annual Charges, (20%)	$\$38.30 \times 10^6$	7,660.0	
10. Total	--	10,073.4	$57.78/lb NO abated[c]

[a]For unit identified in Tables 1-1 and 1-5.

[b]Equivalent annual replacement. Because of relatively low operational use, catalyst life is expected to be at least 2 years.

[c]90% NO_x reduction.

TABLE 4-18. NO_x CONTROL COSTS: INTERNAL COMBUSTION ENGINE[a] -- SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 40.2 lb/hr (as NO_2)
= 3365 ppm (dry, 3% O_2)
MW_e = 14

SCFM = 7340 (Wet)
hr/yr = 1800
Operation: Full Load

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	15 tons/year	3.0	
2. Electric Power, ($0.045/kwh)	11,000 kwh	0.5	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3 (Atomization)	0.1	
5. Fuel Oil, ($0.50/gal)	Not required	--	
6. Catalyst,[b] (200/ft^3)	24.5 ft^3	4.9	
7. Maintenance, Material, Labor (3%)	$ 2.28 x 10^5	6.8	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	1.1	
9. Annual Charges, (20%)	$ 2.56 x 10^{-5}	51.3	
10. Total	--	67.7	$1.04/lb NO_x Abated[c]

[a]For unit identified in Tables 1-1 and 1-5.

[b]Equivalent annual replacement. Because of relative low operational use catalyst life is expected to be at least 2 years.

[c]90% NO_x reduction.

TABLE 4-19. NO_x CONTROL COSTS: OIL FIELD STEAM GENERATOR, NEW INSTALLATION[a] -- SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 22.2 lb/hr (as NO_2)
= 335 ppm (dry, 3% O_2)
MW_e = 6.6

SCFM = 9950 (wet)
hrs/yr = 7,884
Operation: 6.5 GPM Fuel

	Item	Quantity	Annual Cost $(000)	
1.	Ammonia ($200/ton)	43.1 tons/year	8.6	
2.	Electric Power, ($0.045/kwh)	176,000 kwh	7.9	
3.	Water, (0.55/1000 gal)	--	--	
4.	Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	.2	
5.	Fuel Oil, ($0.50/gal)	3,800 bbl/yr	45.4	
6.	Catalyst, (200/ft^3)	48 ft^3	9.6	
7.	Maintenance, Material, Labor (3%)	$1.95 x 10^6	58.6	
8.	Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	3.7	
9.	Annual Charges, (20%)	$2.21 x 10^6	441.8	
10.	Total	--	575.8	$4.03/lb NO_x abated[b]

[a]For unit identified in Tables 1-1 and 1-5.

[b]70% NO_x reduction.

TABLE 4-20. NO_x CONTROL COSTS: OIL FIELD STEAM GENERATOR, EXISTING INSTALLATION[a] — SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 22.2 lb/hr (as NO_2) SCFM = 9950 (wet)

= 335 ppm (dry, 3% O_2) hr/yr = 7,884

MW_e = 6.6 Operation: 6.5 GPM Fuel

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	43.1 tons/year	8.6	
2. Electric Power, ($0.045/kwh)	176,000 kwh	7.9	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	.2	
5. Fuel Oil, ($0.50/gal)	3,800 bbl/yr	45.4	
6. Catalyst, (200/ft^3)	31 ft^3	6.2	
7. Maintenance, Material, Labor (3%)	$1.82 x 10^6	54.8	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	3.7	
9. Annual Charges, (20%)	$2.06 x 10^6	412.0	
10. Total	--	538.8	$3.78/lb NO_x abated[b]

[a]For unit identified in Tables 1-1 and 1-5.

[b]55% NO_x reduction.

TABLE 4-21. COST OF NO_x CONTROL FOR OIL FIELD GENERATORS (EARLY 1979 DOLLARS)

Control Measure	Installation	Annual Cost, $(000)	$/lb NO_x Removed	$/MMBTU Input	$/bbl Oil Produced[a]
SCR	New	575.8	4.08	1.30	4.03
	Existing	538.8	4.85	1.22	3.78
Thermal	New	42.4	0.30	0.096	0.29
	Existing	40.2	0.36	0.091	0.27

[a]Based on 2 barrels (net) of crude oil produced for 1 bbl burned.

TABLE 4-22. NO_x CONTROL COSTS: OIL FIELD STEAM GENERATOR, NEW INSTALLATION[a] -- THERMAL DENITRIFICATION (EARLY 1979 DOLLAR)

NO_x = 22.2 lb/hr NO_x (as NO_2) SCFM = 9950 (wet)

= 335 ppm (dry, 3% O_2) hr/yr = 7,884

MW_e = 6.6 Operation: 6.5 GPM Fuel

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	62.7 tons/year	12.5	
2. Electric Power, ($0.045/kwh)	70,000 kwh	3.2	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	0.3	
5. Fuel Oil, ($0.50/gal)	--	--	
6. Catalyst, (200/ft^3)	--	--	
7. Maintenance, Material, Labor (3%)	$73.3 x 10^3	2.2	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	4.2	
9. Annual Charges, (20%)	$100.0 x 10^3	20.0	
10. Total	--	42.4	$0.30/lb NO_x abated[b]

[a]For unit identified in Tables 1-1 and 1-5.

[b]70% NO_x reduction.

TABLE 4-23. NO_x CONTROL COSTS: OIL FIELD STEAM GENERATOR, EXISTING INSTALLATION[a] -- THERMAL DENITRIFICATION (EARLY 1979 DOLLARS)

NO_x = 22.2 lb/hr (as NO_2)
= 335 ppm (dry, 3% O_2)
MW_e = 6.6

SCFM = 9950 (wet)
hr/yr = 7884
Operation: 6.5 GPM Fuel

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	43.1 tons/year	8.6	
2. Electric Power, ($0.045/kwh)	70,000 kwh	3.2	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	0.2	
5. Fuel Oil, ($0.50/gal)	--	--	
6. Catalyst, (200/ft^3)	--	--	
7. Maintenance, Material, Labor (3%)	$80.6 x 10^3	2.4	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/KW	4.2	
9. Annual Charges, (20%)	$107.8 x 10^3	21.6	
10. Total	--	40.2	$0.36/lb NO_x abated[b]

[a]For unit identified in Tables 1-1 and 1-5.

[b]55% NO_x reduction.

No reheat fuel requirement is anticipated. Operating and maintenance costs are approximately 50% of the total. Unit costs are summarized in Table 4-21.

4.2.6 Refinery Heater

Either SCR or thermal $DeNO_x$ have the potential of meeting the 50% NO_x reduction criteria established for this study.

Total annual costs for SCR and thermal $DeNO_x$ are $137,200 and $40,400 (Tables 4-24 and 4-25), $5.19/lb NO_x removed and $0.30/MM Btu thermal input for SCR and $1.58/lb and $0.092/MM Btu input for thermal $DeNO_x$. Fuel to reheat the exhaust gas is not required for SCR, and none is anticipated for the thermal $DeNO_x$ process.

Operation and maintenance costs represent 27 and 45% of the total for the SCR and thermal systems, with the remainder being annual charge on capital.

4.2.7 Cement Kiln

SCR is capable of meeting the 90% NO_x removal criteria for the dry process cement kiln. Total annual cost is $2,480,000 (Table 4-26). Locating the reactor after particulate cleanup of the exhaust gas (downstream of the baghouse) requires reheating to bring the gas temperature to that required for SCR. The fuel oil represents approximately 30% of the annual cost. The annualized charges on capital and O&M costs are 49 and 51%, respectively. Catalyst replacement represents 10% of the annual costs. Unit costs are $2.40/lb NO_x removed, $1.75/MM Btu and $6.42/ton clinker (45 tons/hr).

4.2.8 Glass Melting Furnace

In order to utilize SCR for 90% removal, the system requires for particulate removal an electrostatic precipitator (ESP), which is included in the total capital investment (Table 4-27). Total annual cost is approximately 1.7 million dollars, with O&M costs representing 35%. The cost of the ESP represents approximately 60% of the total capital investment. The total annual cost represents $0.91/lb NO_x removed and $1.38/MM Btu heat input.

TABLE 4-24. NO_x CONTROL COSTS: OIL REFINERY HEATER[a] -- SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 6.7 lb/hr (as NO_2)
= 77 ppm (dry, 3% O_2)
MW_e = 6.8

SCFM = 11,700 (dry)
hr/yr = 7884
Operation: 100% Load

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	10.7 tons/yr	2.1	
2. Electric Power, ($0.045/kwh)	204,000 kwh	9.2	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	0.1	
5. Fuel Oil, ($0.50/gal)	Not Required	--	
6. Catalyst, (200/ft^3)	40 ft^3 [b]	8.0	
7. Maintenance, Material, Labor (3%)	$4.32 x 10^5	13.0	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	4.8	
9. Annual Charges, (20%)	$5.00 x 10^5	100.0	
10. Total		137.2	$5.19/lb NO_x abated[c]

[a]For unit identified in Tables 1-1 and 1-5.

[b]Estimated annual replacement.

[c]50% NO_x reduction.

TABLE 4-25. NO_x CONTROL COSTS: OIL REFINERY HEATER[a] -- THERMAL DENITRIFICATION (EARLY 1979 DOLLARS)

NO_x = 6.7 lb/hr (as NO_2)
= 77 ppm (dry, 3% O_2)
MW_e = 6.8

SCFM = 11,700 (dry)
hr/yr = 7884
Operation: 100% Load

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	10.7 tons/year	2.1	
2. Electric Power, ($0.045/kwh)	82,000 kwh	9.2	
3. Water, (0.55/1000 gal)	--	--	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	0.1	
5. Fuel Oil, ($0.50/gal)	--	--	
6. Catalyst, (200/ft^3)	--	--	
7. Maintenance, Material, Labor (3%)	$83.0 x 10^3	2.5	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	4.3	
9. Annual Charges, (20%)	$11.10 x 10^3	22.2	
10. Total	--	40.4	$1.58/lb NO_x abated[b]

[a]For unit identified in Tables 1-1 and 1-5.

[b]50% NO_x reduction.

TABLE 4-26. NO_x CONTROL COSTS: CEMENT KILN[a] -- SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 146 lb/hr (as NO_2)
= 274 ppm (dry, 3% O_2)
MW_e = 18.9

SCFM = 144,000
hr/yr = 7884

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	233 tons/year	46.5	
2. Electric Power, ($0.045/kwh)	567,000 kwh	25.5	
3. Water, (0.55/1000 gal)	11 GPM	2.9	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	1.3	
5. Fuel Oil, ($0.50/gal)	4.60 bbl/HR	762.5	
6. Catalyst, (200/ft^3)	1265 ft^3	253.2	
7. Maintenance, Material, Labor (3%)	$5.16 x 10^6	154.9	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	11.9	
9. Annual Charges, (20%)	$6.12 x 10^6	1,223.8	
10. Total	--	2,482.5	$2.40/lb NO_x abated[b]

[a]For unit identified in Tables 1-1 and 1-5.

[b]90% NO_x reduction.

TABLE 4-27. NO_x CONTROL COSTS; GLASS MELTING FURNACE[a] -- SELECTIVE CATALYTIC REDUCTION (EARLY 1979 DOLLARS)

NO_x = 220 lb/hr (as NO_2)
= 834 ppm (dry, 3% O_2)
MW_e = 15.4

SCFM = 95,000 (wet)
hr/yr = 8760
Capacity Factor = 100%

Item	Quantity	Annual Cost $(000)	
1. Ammonia ($200/ton)	392 tons /year	78.3	
2. Electric Power, ($0.045/kwh)	513,000 kwh	23.1	
3. Water, (0.55/1000 gal)	8.8 GPM	2.6	
4. Steam, ($1.80/1000 lb)	1.5 lb/lb NH_3	2.1	
5. Fuel Oil, ($0.50/gal)	Not required	--	
6. Catalyst,[b] (200/ft^3)	1560 ft^3	312.2	
7. Maintenance, Material, Labor (3%)	$6.18 x 10^6	185.4	
8. Operating Labor, ($16/hr)	5 x 10^{-3} men/MW	10.8	
9. Annual Charges,[b] (20%)	$5.72 x 10^6	1,144.4	
10. Total	--	1,758.9	$0.91/lb NO abated[c]

[a]For unit identified in Tables 1-1 and 1-5.

[b]Estimated catalyst life 1 year of operation.

[c]90% NO_x reduction.

4.3 REFERENCES

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4-17. Personal Communication: Chubu Electric Power Company to P. P. Leo, April 13, 1979.

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APPENDIX A

STATIONARY SOURCE CHARACTERISTICS

The various organizations operating the stationary sources of NO_x emissions provided operating data and other information basic to the conduct of this study. These were in the form of responses to questionaires prepared by The Aerospace Corporation. Copies are included as Tables A-1 through A-5 for the turbines at the Southern California Edison (SCE) Etiwanda and Cool Water Generating Stations; the internal combustion (IC) engine at the Southern California Gas Company Playa del Rey Station; the California Portland Cement Company kiln at Colton, California; and the PPG Industries glass furnace at Fresno, California.

TABLE A-1. SIMPLE TURBINE DATA

FACILITY DATA

1)	Facility Identification	Etiwanda Generating Station
2)	Installation Location	Etiwanda California
3)	Please attach process flow diagram, schematic, or provide reference (include gas temperature profiles, burner array schematic, overfire location if used, and FGR location)	
4)	Source Rating (quantity/day of product)	2904 megawatts/day
5)	Plant Equipment Layout (please provide plot plan)	Attached
6)	Unit Identification (No., etc.)	Unit No. 5
7)	Equipment Manufacturer	Turbo Power and Marine
8)	Year Placed in Service	1969
9)	Projected Lifetime (yrs.)	30
10)	Type of Operation (cyclic, peak, continuous, etc.)	Peaking
11)	Operating Cycle (% max. capacity vs. time) & Operating Range (% of max.)	1% full load

FACILITY DATA (Continued)

12)	Projected Operating Load Factors or Projected Changes in Operating Mode During Remainder of Projected Life-Time	1 to 2% full load
13)	Max. Continuous Generating Capacity (MW or Steam, lbs/hr)	121 mw
14)	Total Product Produced Annually (Kwh, tons, etc.)	12,100,000 KWh
15)	Annual Operation (hrs/yr)	50 to 150 hours
16)	Year Applicable	1969 to 1979
17)	Hours per Year at Max. Capacity	Normally zero at max. peak load
18)	Fuel: i.e., Oil/Gas, and Percent of Each Used Annually	Oil/Gas 70/30
19)	Fuel: Present and Anticipated (if any changes expected)	
	Composition (if gas)	Natural Gas
	Grade (if oil)	JP5
	C/H Ratio	
	% Sulfur	
	% Nitrogen	
	% Ash	
	GHV	
	Consumption/Day	
20)	Max. Continuous Heat Input (10^6 Btu /hr)	355x10 Btu^6(estimated)
21)	Unit Heat Rate (Btu/Kwh)	15000 Btu/KWh

FACILITY DATA (Continued)

22)	Max. and Average Fuel Consumption (bbl/hr., etc.)	332 BBL & 335 mcf &
23)	Percent Excess Air (avg. & max.)	15%
24)	Number of Chimneys	2
25)	Chimney Heights	50-1/2 feet approximately
26)	Average Flue Gas (Exhaust) Discharge Rate (ACFM or SCFM)	1045
27)	Max. Continuous Flue Gas Rate (ACFM or SCFM)	N.A.
28)	Stack Gas Exit Velocity	N.A.
29)	Flue Gas Temperature, ^{o}F (Max. and average)	900
30)	Exhaust Gas Reheat:	None
	Type of Reheater (Manufacturer, Capacity, etc.)	
	Fuel Type & Grade	
	Fuel Consumption	
	Temperature Rise (^{o}F)	
	Inlet/Outlet Gas Velocities	
	Fraction of Flue Gas Reheated (%)	
31)	Concentration in Exhaust, %, (Max. & Average):	
	SO_2	N.A.

FACILITY DATA (Continued)

31) (continued)

NO_x

Particulates

Oxygen

H_2O

32) Emission Controls Currently in Use & Control Characteristics (% removal SO_2, etc.):

None

SO_2

NO_x

Particulates

33) Burner Configuration, Type

N.A.

Number of Burners

Fraction of Burners Active

Fraction of Excess Air in Active Burners

Fraction of Combustion Air in Overfire Air

TABLE A-2. COMBINED CYCLE TURBINE DATA

FACILITY DATA

1)	Facility Identification	Cool Water Generating Station
2)	Installation Location	Daggett, CA.
3)	Please attach process flow diagram, schematic, or provide reference (include gas temperature profiles, burner array schematic, overfire location if used, and FGR location)	NA
4)	Source Rating (quantity/day of product)	11328 megawatts/day
5)	Plant Equipment Layout (please provide plot plan)	Attached
6)	Unit Identification (No., etc.)	Unit Nos. 3 and 4
7)	Equipment Manufacturer	Westinghouse
8)	Year Placed in Service	1978
9)	Projected Lifetime (yrs.)	30
10)	Type of Operation (cyclic, peak, continuous, etc.)	Cyclic
11)	Operating Cycle (% max. capacity vs. time) & Operating Range (% of max.)	100% max. 15% of time

FACILITY DATA (Continued)

12)	Projected Operating Load Factors or Projected Changes in Operating Mode During Remainder of Projected Life-Time	15 to 20% capacity factor 1979 to 1983, thereafter uninown
13)	Max. Continuous Generating Capacity (MW or Steam, lbs/hr)	472 MW
14)	Total Product Produced Annually (Kwh, tons, etc.)	NA
15)	Annual Operation (hrs/yr)	NA
16)	Year Applicable	NA
17)	Hours per Year at Max. Capacity	NA
18)	Fuel: i.e., Oil/Gas, and Percent of Each Used Annually	Oil - 100%
19)	Fuel: Present and Anticipated (if any changes expected)	
	Composition (if gas)	
	Grade (if oil)	No. 2
	C/H Ratio	-
	% Sulfur	0.08
	% Nitrogen	0.01
	% Ash	25 ppm
	GHV	19000-20000 Btu/lb.
	Consumption/Day	NA
20)	Max. Continuous Heat Input (10^6 Btu /hr)	4196 (Design) -
21)	Unit Heat Rate (Btu/Kwh)	8890 (Design)

FACILITY DATA (Continued)

22)	Max. and Average Fuel Consumption (bbl/hr., etc.)	NA -
23)	Percent Excess Air (avg. & max.)	15%
24)	Number of Chimneys	2
25)	Chimney Heights	250 ft.
26)	Average Flue Gas (Exhaust) Discharge Rate (ACFM or SCFM)	2505×10^3 lbs/hr. -
27)	Max. Continuous Flue Gas Rate (ACFM or SCFM)	2505×10^3 lbs/hr.
28)	Stack Gas Exit Velocity	NA
29)	Flue Gas Temperature, °F (Max. and average)	300°F -
30)	Exhaust Gas Reheat:	-
	Type of Reheater (Manufacturer, Capacity, etc.)	John Zink Duct Burners
	Fuel Type & Grade	Distillate No. 2
	Fuel Consumption	11600 lbs/hr.
	Temperature Rise (°F)	250°F
	Inlet/Outlet Gas Velocities	-
	Fraction of Flue Gas Reheated (%)	100%
31)	Concentration in Exhaust, %, (Max. & Average):	-
	SO_2	NA

FACILITY DATA (Continued)

31) (continued)

NO_x	NA (140 lbs/hr. max limit)
Particulates	NA (10 lbs/hr.
Oxygen	15%
H_2O	NA

32) Emission Controls Currently in Use & Control Characteristics (% removal SO_2, etc.):

SO_2	Clean Fuel
NO_x	Water Injection
Particulates	Clean Fuel

33) Burner Configuration, Type

Number of Burners

Fraction of Burners Active

Fraction of Excess Air in Active Burners

Fraction of Combustion Air in Overfire Air

TABLE A-3. INTERNAL COMBUSTION ENGINE DATA

FACILITY DATA

1) Plant

a) Facility Identification — Playa del Rey Station

b) Installation Location — 8141 Gulana Avenue, Playa del Rey, California 90291

~~c) Plant Layout~~

~~d) Process Diagram (please attach process flow diagram, schematic or provide reference -- include water injection and EGR, if used).~~

2) Typical Unit — Cooper Bessemer GMVH - 10

a) Unit Identification (include manufacturer)

b) Year Placed in Service — 1965

c) Project Lifetime (Years) — 50 - 100 years

d) Rated (Typical Operating) Conditions

Horsepower — 2000 H.P.

Fuel Consumption (SCFM — 12,581 CFH

Fuel/Air Ratio — 30 - 1

e) Maximum Continuous Horsepower-Special conditions for max. continuous (i.e., water injection, changed fuel/air ration, etc.) — Various H.P. - as required

f) SCFM Gas at Rated Conditions — 6800 BTU/BHP @ 80°

continued

g) Annual Operation

Hours/Year	1865
Hours/Year at Maximum Continuous Opertion	---
Projected Changes in Annual Operation	None

3) Multiple Unit Installations (if applicable)

a) Number of Units per Installation	Total of 9 engines - 3 GMVH-10
b) Number of Units per Stack	1
c) Total Rated Heat Input for Installation	---
d) Proportions of Fuel, if Different Units in the Installation	---
e) Rated Continuous Exhaust Gas Flow Rate per Stack (ACFM or SCFM)	35,000# per hour @ 80°

4) Fuels Used (Present and Anticipated — Natural Gas)

a) Gas Composition and Gross Heating Value (BTU/SCF)	1080 BTU .626 Grav. Low heat value 953
b) Oil	None
Grade	
~~Ultimate Analysis (Wt. %)~~	
~~Carbon~~	
Hydrogen	
Nitrogen	
Sulfur	
Oxygen	
~~Ash~~	
~~Gross Heating Value (BTU/lb)~~	
~~Density (lb/gal)~~	

5) Air Pollution Emissions and Control (Typical Unit

a) Exhaust Gas Temperatures, of (Rated Conditions)

650°

Combustion Air Temperature Manifold)

98° max.

Maximum Entering Exhaust Gas Manifold

650°

Maximum Entering Exhaust Stack ~~Temperature at Emission Control Devices (if applicable)~~

~~b) Exhaust Gas Analysis (Rated & Max Conditions)~~ --

CO_2 (%)

H_2O (%)

O_2 (%)

CO (ppm)

NO_x (ppm)

Hydrocarbons (ppm)

Particulates (Units?)

~~c)~~ Emission Controls Currently in Use and Control Characteristics

None

NO_x

Particulates

Hydrocarbons

EGR (% of combustion-air)

Other (special combustion chamber design, water injection, etc.)

6) Other Information

TABLE A-4. CEMENT KILN DATA

FACILITY DATA

1)	Facility Identification	CALIFORNIA PORTLAND CEMENT COMPANY
2)	Installation Location	Colton, California 92324
3)	Please attach process flow diagram, schematic, or provide reference (include gas temperature profiles, burner array schematic, overfire location if used, and FGR location)	
4)	Source Rating (quantity/day of product)	1100 ton clk/day each kiln
5)	Plant Equipment Layout (please provide plot plan)	See 3
6)	Unit Identification (No., etc.)	
7)	Equipment Manufacturer	Allis Chalmers Kiln
8)	Year Placed in Service	1963
9)	Projected Lifetime (yrs.)	Unknown
10)	Type of Operation (cyclic, peak, continuous, etc.)	Continuous
11)	Operating Cycle (% max. capacity vs. time) & Operating Range (% of max.)	Operates approximately 90% of time.

FACILITY DATA (Continued)

12)	Projected Operating Load Factors or Projected Changes in Operating Mode During Remainder of Projected Life-Time	Refer to 4	
13)	Max. Continuous Generating Capacity (MW or Steam, lbs/hr)	N/A	
14)	Total Product Produced Annually (Kwh, tons, etc.)	374,000 tons/yr. each kiln	
15)	Annual Operation (hrs/yr)	7796	
16)	Year Applicable	1978	
17)	Hours per Year at Max. Capacity	7796	
18)	Fuel: i.e., Oil/Gas, and Percent of Each Used Annually	81% Coal 19% Oil	
19)	Fuel: Present and Anticipated (if any changes expected)		
	Composition (if gas)		
	Grade (if oil)	Oil Bunker "C"	Coal
	C/H Ratio	Unknown	Unknown
	% Sulfur	1.20	0.35
	% Nitrogen	Unknown	Unknown
	% Ash	0.099	12.03
	GHV	150,000 BTU/gal.	22,804,000 BTU/ton
	Consumption/Day	6,866 gal/day	189 T/day
20)	Max. Continuous Heat Input (10^6 Btu /hr)	43	180
21)	Unit Heat Rate (Btu/Kwh)	N/A	

FACILITY DATA (Continued)

22)	Max. and Average Fuel Consumption (bbl/hr., etc.)	7.9T of coal/hr. and 286 gal. of oil/hr.
23)	Percent Excess Air (avg. & max.)	
24)	Number of Chimneys	1 each kiln
25)	Chimney Heights	100' approximately
26)	Average Flue Gas (Exhaust) Discharge Rate (ACFM or SCFM)	144,000 SCFM
27)	Max. Continuous Flue Gas Rate (ACFM or SCFM)	Refer to 26
28)	Stack Gas Exit Velocity	38 ft./sec.
29)	Flue Gas Temperature, ^{o}F (Max. and average)	350°F.
30)	Exhaust Gas Reheat:	N/A
	Type of Reheater (Manufacturer, Capacity, etc.)	
	Fuel Type & Grade	
	Fuel Consumption	
	Temperature Rise (^{o}F)	
	Inlet/Outlet Gas Velocities	
	Fraction of Flue Gas Reheated (%)	
31)	Concentration in Exhaust, %, (Max. & Average):	
	SO_2	0 to 50 ppm

FACILITY DATA (Continued)

31) (continued)

NO_x	125 to 300 ppm
Particulates	0.0100 grains/ft.^{3}
Oxygen	14.4%(@ kiln stack)
H_2O	5%

32) Emission Controls Currently in Use & Control Characteristics (% removal SO_2, etc.):

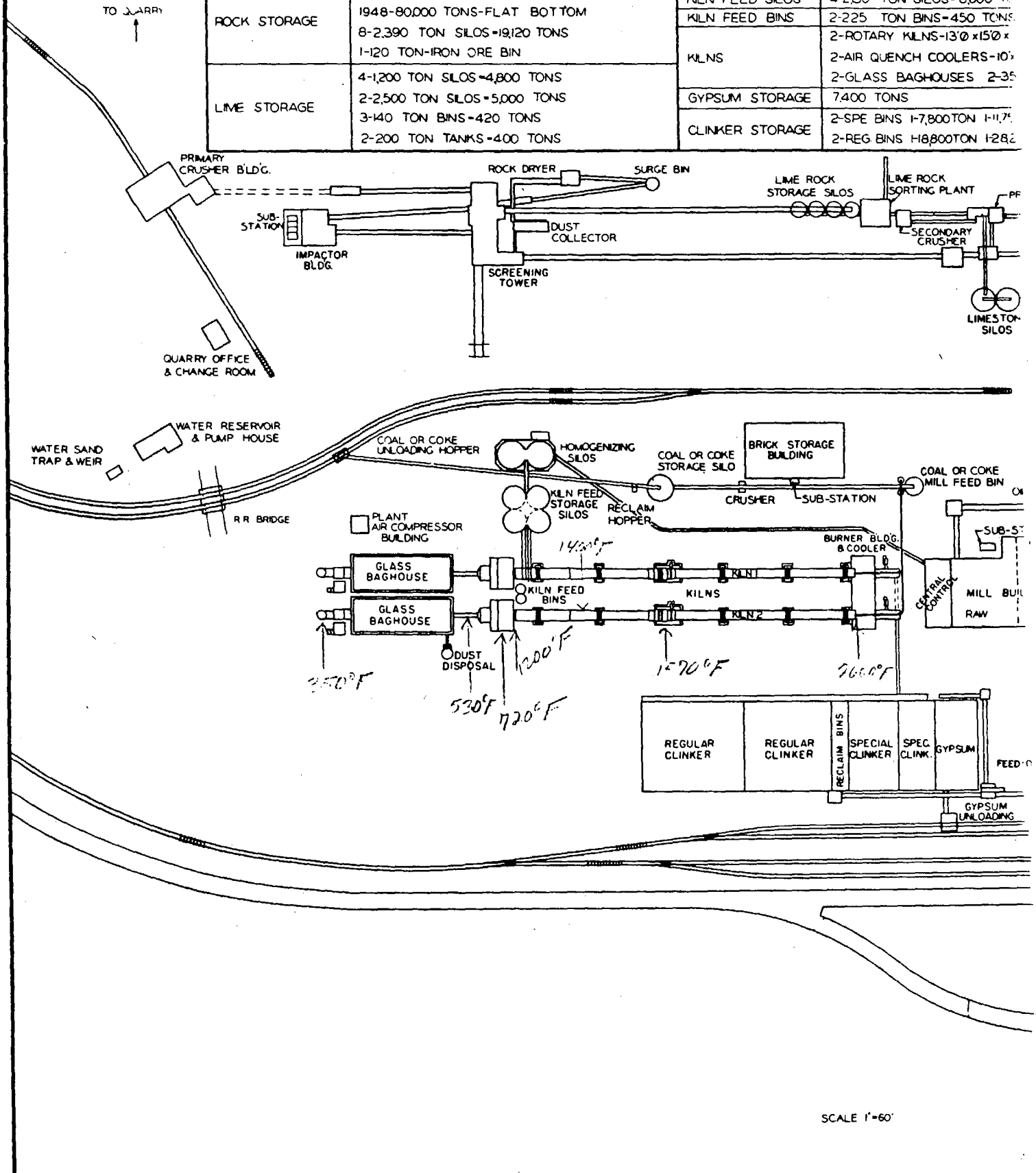
	None
SO_2	None
NO_x	None
Particulates	Glass baghouse and multiclones

33) Burner Configuration, Type

Number of Burners	1 Coal and 3 Oil per kiln
Fraction of Burners Active	Varies
Fraction of Excess Air in Active Burners	20% in coal burner
Fraction of Combustion Air in Overfire Air	NA

DEPT	EQPT SCHEDULE & STORAGES
CRUSHER	1-60" GYRATORY 2-PENN IMPACTORS (C15-50)
ROCK STORAGE	1914-17,200 TONS (6 BINS) 1948-80,000 TONS-FLAT BOTTOM 8-2,390 TON SILOS=19,120 TONS 1-120 TON-IRON ORE BIN
LIME STORAGE	4-1,200 TON SILOS=4,800 TONS 2-2,500 TON SILOS=5,000 TONS 3-140 TON BINS=420 TONS 2-200 TON TANKS=400 TONS

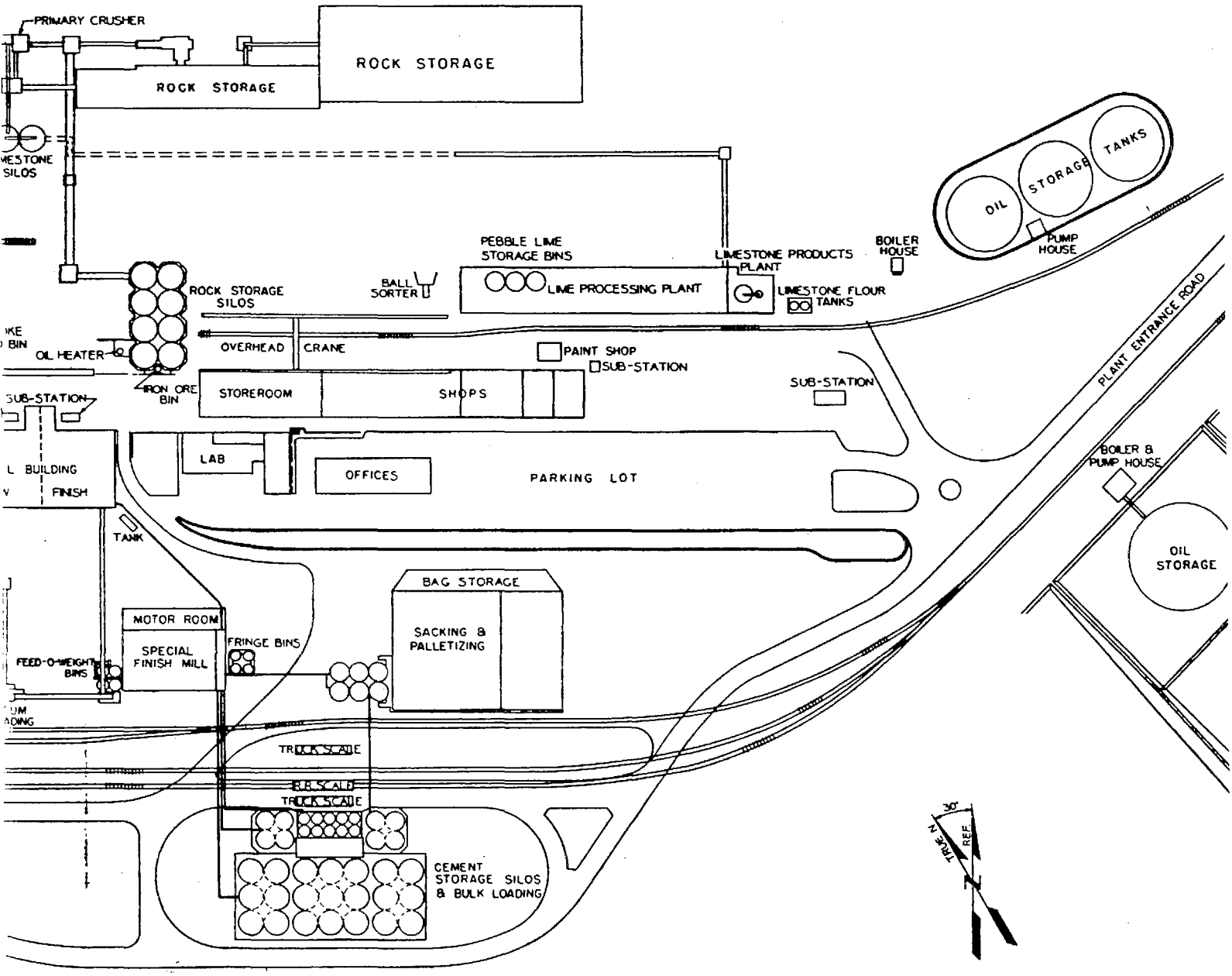
DEPT	EQPT SCHEDULE
RAW MILL	2-BALL MILLS-13'Ø x17' LONG 4-SEPARATORS-16'Ø (2/MILL
HOMOGENIZING SILOS	2-1,800 TON SILOS=3,600 T
KILN FEED SILOS	4-2,150 TON SILOS=8,600 T
KILN FEED BINS	2-225 TON BINS=450 TONS
KILNS	2-ROTARY KILNS-13'Ø x15'Ø x 2-AIR QUENCH COOLERS-10' 2-GLASS BAGHOUSES 2-35
GYPSUM STORAGE	7,400 TONS
CLINKER STORAGE	2-SPE BINS 1-7,800 TON 1-11,7 2-REG BINS 1-8,800 TON 1-28,2

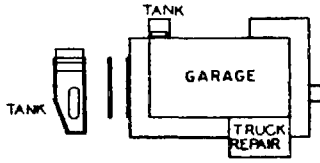


SCALE 1"=60'

NO.	REVISION	BY	DATE
1	REDRAWN	T.M.I.	3/15/77

E & STORAGES	DEPT	EQPT. SCHEDULE & STORAGES.	DEPT.	EQPT. SCHEDULE & STORAGES
LONG /MILL-CLOSED CIRC.)	FINISH MILL	2-BALL MILLS-13'0 x 21' LONG 4-SEPARATORS-16'0 (2/MILL-CLOSED CIRC.)	CEMENT SILOS (CONT'D)	10-195 TON BINS=1,950 TONS TOTAL STORAGE=79,638.8 TONS
00 TONS 00 TONS	SPECIAL GRINDING	4-BALL MILLS-7'0 x 24' LONG 4-BALL MILLS-7'0 x 26' LONG	OIL STORAGE	3-CONCRETE TANKS-20,000 BBLS=60,000 BBL 1-STEEL TANK-100,000 BBLS.
TONS	SPE. FINISH STG BINS	4-165 TON BINS = 660 TONS	COAL STORAGE	1-2,500 TON SILO 1-250 TON MILL FEED BIN
15'0 x 490' LONG S-10' x 55' 2-350,000 CFM EA.	FEED-O-WEIGHT BINS	4-210 TON BINS = 840 TONS	ROCK DRYER	9'0 x 80' LG. ROTARY DRYER
1-11,750 TON 1-28,200 TON } =66,550 TONS	CEMENT SILOS	21-2820 TON SILOS=59,220 TONS 8-1504 TON SILOS=12,032 TONS 6-9976 TON SILOS=5,985.6 TONS 2-2256 TON SILOS=451.2 TONS		





CALIFORNIA PORTLAND CEMENT COMPANY

COLTON , CALIFORNIA

5649

TABLE A-5. GLASS MELTING FURNACE DATA

FACILITY DATA

1)	Facility Identification	PPG Industries Inc. Wks. No 15, Fresno, Ca
2)	Installation Location	3333 South Peach Fresno, Ca.
3)	Please attach process flow diagram, schematic, or provide reference (include gas temperature profiles, burner array schematic, overfire location if used, and FGR location)	See Attached
4)	Source Rating (quantity/day of product)	Conf.
5)	Plant Equipment Layout (please provide plot plan)	N/A
6)	Unit Identification (No., etc.)	N/A
7)	Equipment Manufacturer	PPG
8)	Year Placed in Service	1976
9)	Projected Lifetime (yrs.)	20 yrs.
10)	Type of Operation (cyclic, peak, continuous, etc.)	Continuous
11)	Operating Cycle (% max. capacity vs. time) & Operating Range (% of max.)	N/A

FACILITY DATA (Continued)

12) Projected Operating Load Factors or Projected Changes in Operating Mode During Remainder of Projected Life-Time — N/A

13) Max. Continuous Generating Capacity (MW or Steam, lbs/hr) — N/A

14) Total Product Produced Annually (Kwh, tons, etc.) — Confidential

15) Annual Operation (hrs/yr) — Continuous

16) Year Applicable — ?

17) Hours per Year at Max. Capacity — N/A

18) Fuel: i.e., Oil/Gas, and Percent of Each Used Annually — ~~Oil~~ Gas - 100%

19) Fuel: Present and Anticipated (if any changes expected) — Gas
Composition (if gas)
Grade (if oil)
C/H Ratio
% Sulfur
% Nitrogen
% Ash
GHV
Consumption/Day

20) Max. Continuous Heat Input (10^6 Btu/hr) — Conf.

21) Unit Heat Rate (Btu/Kwh) — N/A

FACILITY DATA (Continued)

22)	Max. and Average Fuel Consumption (bbl/hr. , etc.)	Conf.
23)	Percent Excess Air (avg. & max.)	10% max 5% avg.
24)	Number of Chimneys	1
25)	Chimney Heights	84 ft
26)	Average Flue Gas (Exhaust) Discharge Rate (ACFM or SCFM)	95,000 scfm + 1/3 prod of comp 1/2 ambient air from ejector
27)	Max. Continuous Flue Gas Rate (ACFM or SCFM)	same
28)	Stack Gas Exit Velocity	25.4 ft/sec.
29)	Flue Gas Temperature, °F (Max. and average)	450° F top of stack
30)	Exhaust Gas Reheat:	N/A
	Type of Reheater (Manufacturer, Capacity, etc.)	
	Fuel Type & Grade	
	Fuel Consumption	
	Temperature Rise (°F)	
	Inlet/Outlet Gas Velocities	
	Fraction of Flue Gas Reheated (%)	
31)	Concentration in Exhaust, % (Max. & Average):	
	SO_2	

FACILITY DATA (Continued)

31) (continued)

NO_x	220 lbs/hr
Particulates	19 lbs/hr.
Oxygen	
H_2O	

32) Emission Controls Currently in Use & Control Characteristics (% removal SO_2, etc.):

	N/A
SO_2	
NO_x	
Particulates	

33) Burner Configuration, Type — Confidential

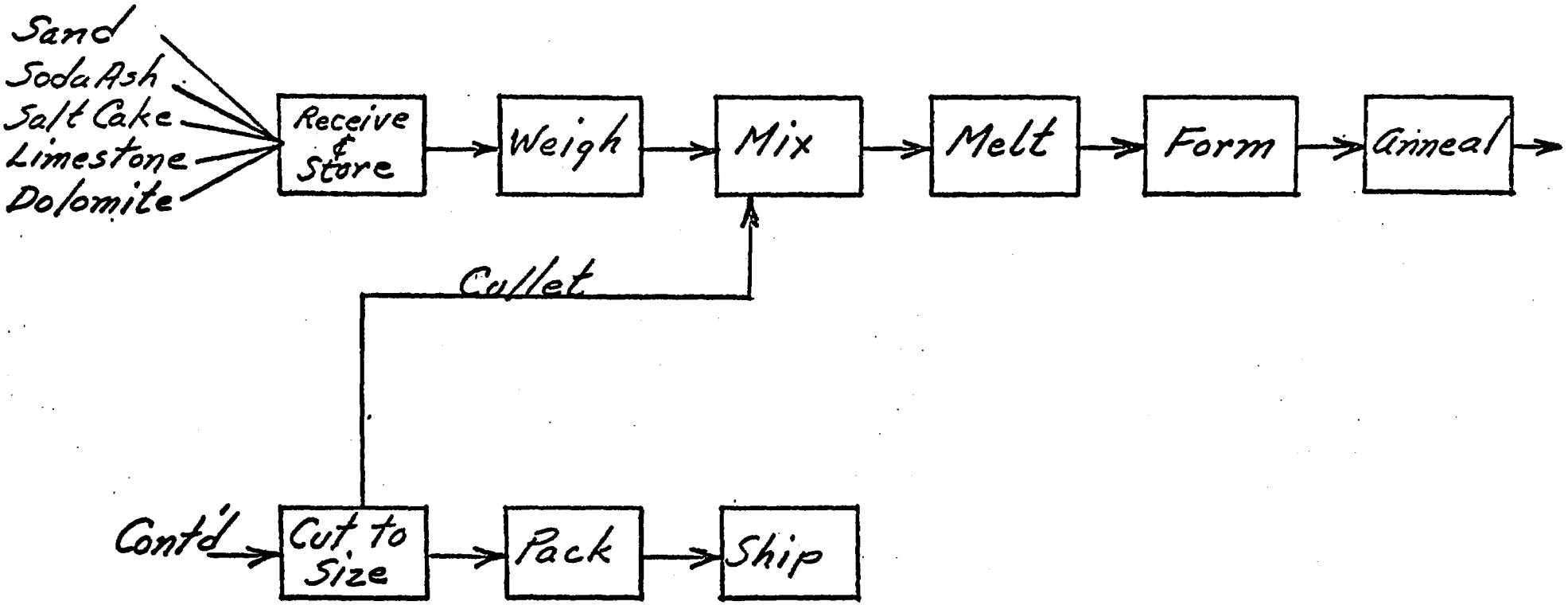
Number of Burners

Fraction of Burners Active

Fraction of Excess Air in Active Burners

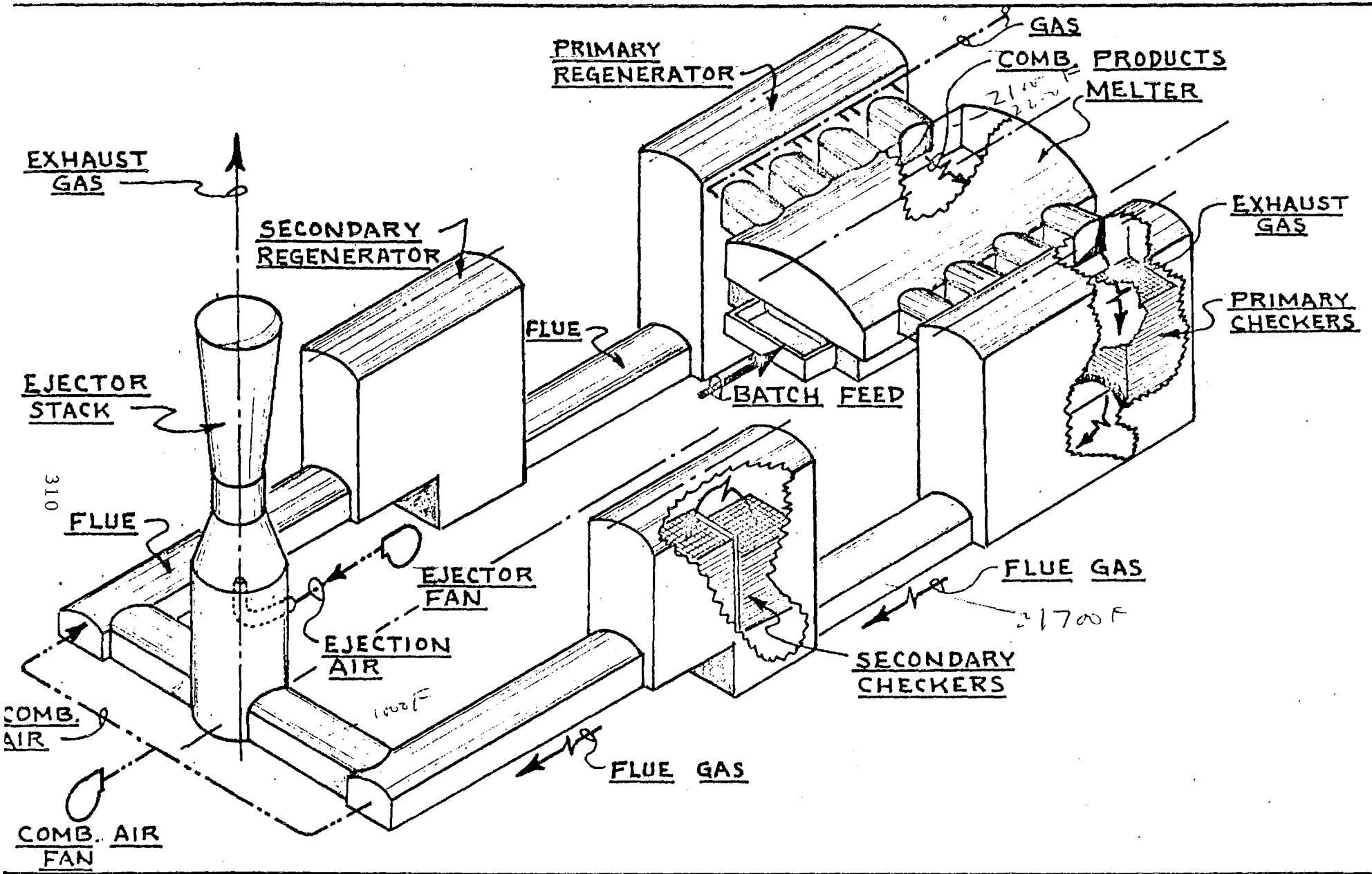
Fraction of Combustion Air in Overfire Air

FLOW DIAGRAM - FLOAT GLASS



309

HF c/7/78



REVISIONS		PPG INDUSTRIES, Inc. WORKS NO. 14	AUTH.	REF.
		PROPOSED FLOAT GLASS	DATE:	BY:
		MELTING FURNACE	"S" NO.	
		FLOW DIAGRAM		