June 28, 2007

Mr. Gerardo C. Rios
Chief, Permits Office
United States Environmental Protection Agency
Region IX
75 Hawthorne Street
San Francisco, CA 94105-3901

Dear Mr. Rios:

This letter amends the revised Prevention of Significant Deterioration (PSD) application submitted to your office on December 11, 2006 for the Big West of California, LLC Bakersfield Refinery Clean Fuels Project (CFP). There are several amendments to be made to the application, specifically:

- The refinery has committed to meeting a fuel gas sulfur content limit of 40 ppmv total sulfur (as H_2S) on a 4-hour average basis;
- Basic information is presented about the additional equipment that will be necessary to achieve this fuel gas sulfur level;
- The location of the existing Mild Hydrocracker and its two heaters 14-H1 and 14-H2 was incorrectly modeled;
- Additional supporting information is presented regarding the cost effectiveness of installing selective catalytic reduction (SCR) on the two CFP heaters that are less than 50 MMBtu/hr (VGO Feed Heater, 47 MMBtu/hr; and VGO-HDS Fractionator Feed Heater, 35 MMBtu/hr);
- An analysis of the proposed NSPS Subpart Ja and how it affects the project combustion units and FCCU, which indicates that a short-term NOx emissions limit will be required for the FCCU, additional CEMS or parametric monitoring will be required for PM emissions from the FCCU, and additional NOx CEMS monitoring will be required for the VGO-HDS heaters; *and*
- Additional clarifications on the design and operation of the ground flare that will serve CFP units.

Each of these amendments and additional ground flare data are described in detail below.

Fuel Gas Sulfur Limit

As you are aware, Big West has worked extensively with our process design engineers to achieve a fuel gas sulfur limit lower than 100 ppmv. As presented in the July 12, 2006 letter from J.

Scott Lewis of Linde BOC Process Plants LLC (provided as an attachment to an email dated July 24, 2006 from Ev Ashworth), the contractor for the amine treatment system was only willing to guarantee that 50 ppmv H_2S could be achieved. However, as the design has progressed, we were able to develop detailed fuel gas balance scenarios to ascertain the expected fuel gas sulfur content in normal and "worst-case" operating situations. Big West now anticipates that with the addition of a caustic scrubber to remove the potentially high level of sulfur in the fuel gas from Area 3 that may, in some circumstances, be introduced into the Area 4 fuel gas system, Big West can meet a limit of 40 ppmv total sulfur (as H_2S) averaged over a 4-hour period at all times.

This modification will not adversely affect the PSD permit application. Modeling will not be performed again, as the current modeling is conservative in its inclusion of 100 ppmv sulfur in the combusted fuel gas. To meet the 40 ppmv limit, the facility will need to treat fuel gas supplied to Area 2 from the Area 3 Delayed Coker Gas Amine Treater, to reduce non-H_2S sulfur compounds in the total fuel gas burned in Areas 2 and 4. The new treatment unit will use caustic to extract these sulfur compounds; the sulfur compounds will be converted to disulfides and returned to a hydrotreater for conversion to H_2S.

The addition of this caustic scrubbing unit does not directly affect the PSD permit application, as the only emissions from the unit will be volatile organic compounds (VOC), which is not within the scope of the PSD permit. The refinery will submit a revised application for a Permit to Construct with the San Joaquin Valley Air Pollution Control District to support construction of the revised Area 4 fuel gas treatment unit.

A revised Table 5-3, that reflects the lower SO_2 emissions resulting from combustion of fuel gas subject to a 40 ppmv sulfur content limit, is presented below.

Table 5-3: Big West Clean Fuels Project Source Emission Rates[a]

Source ID	Model ID	NO_x (g/s)	SO_2[b] (g/s)	CO (g/s)	PM_{10} (g/s)
VGO Feed Heater (47 MMbtu/hr)	vgohtr	0.1438	~~0.0833~~ 0.0333	0.2189	0.0441
VGO HDS Fractionator Feed Heater (35 MMBtu/hr)	vgofrhtr	0.1071	~~0.0620~~ 0.0248	0.1630	0.0329
Hydrogen Plant Reformer	h2reform	0.4904	~~1.1364~~ 0.4545	0.5971	0.6018
FCCU Regenerator (annual) [c]	fccuregen	1.0604	1.4765	1.9046	1.1647
FCCU Regenerator (1-hr, 3-hr, 8-hr, 24-hr) [c]	fccuregen	2.1208	3.6913	16.1407	1.1647
Existing MHC Feed Heaters (14-H1 & 14-H2)	mhc14h12	0.3748	~~0.1596~~ 0.0638	1.4904	0.0845
HF Alky Isostripper Reboiler	hfreboil	0.1645	~~0.3812~~ 0.1525	0.2003	0.2018
SWAATS Unit	swaats	0.0000	0.2322	~~4.3994~~ 0.3384	0.0000
Ground Flare[d]	gndflare	0.0279	~~0.0089~~ 0.0057	0.1519	0.0107
Diesel Firewater Pump Engines (annual)	firepump	0.0222	0.0000	0.0130	0.0007
Diesel Firewater Pump Engines (24-hr)	firepump	0.0809	0.0001	0.0474	0.0027
Diesel Firewater Pump Engines (8-hr)	firepump	0.2428	0.0003	0.1422	0.0082
Diesel Firewater Pump Engines (3-hr)	firepump	0.6475	0.0007	0.3792	0.0219
Diesel Firewater Pump Engines (1-hr)	firepump	1.9425	0.0022	1.1375	0.0656
Cooling Tower 1	coolt1	0.0000	0.0000	0.0000	0.0303
Cooling Tower 2	coolt2	0.0000	0.0000	0.0000	0.0303

[a] Strikeout values are from the December 2006 revised PSD application.

[b] Revised sulfur emission rates reflect combustion of refinery fuel gas at 40 ppmv total sulfur, expressed as H_2S.

[c] FCCU heater is a limited-use startup heater. FCCU regenerator emission rates are larger than those from the FCCU startup heater, so FCCU regenerator emissions were used in the modeling.

[d] Flare emission rates are annual averages that include process unit startups and shutdowns as well as continuous pilot flaring.

Mild Hydrocracker Location

It recently came to our attention that the location of the existing mild hydrocracker, which has been included as an affected unit in the PSD air quality impact analysis modeling for the Clean Fuels Project, was misrepresented in these dispersion modeling runs. The UTM coordinates should not have been 311795.2 Easting and 3917118.9 Northing (NAD27) as presented in Table 5-4 of the December 2006 revised application, but rather should be approximately 220 m SSE, at 311837.5 Easting and 3916901.5 Northing.

The dispersion modeling to compare maximum project impacts with Class II significance levels and monitoring significance levels has been performed again to take this change into account. The affected tables and figures from the December 2006 revised PSD application are included below. As noted above, this revised modeling analysis does not reflect the new and reduced fuel gas total sulfur content limit of 40 ppmv. However, this conservative approach demonstrates that the revised location of the Mild Hydrocracker does not result in any exceedances of relevant EPA PSD Significance Levels. The modeling to determine Air Quality Related Values (AQRV) impacts on Class I areas has not been revised, as the effect of a 200 m shift of one emission source would be imperceptible at the distance of the nearest Class I areas (~80 km).

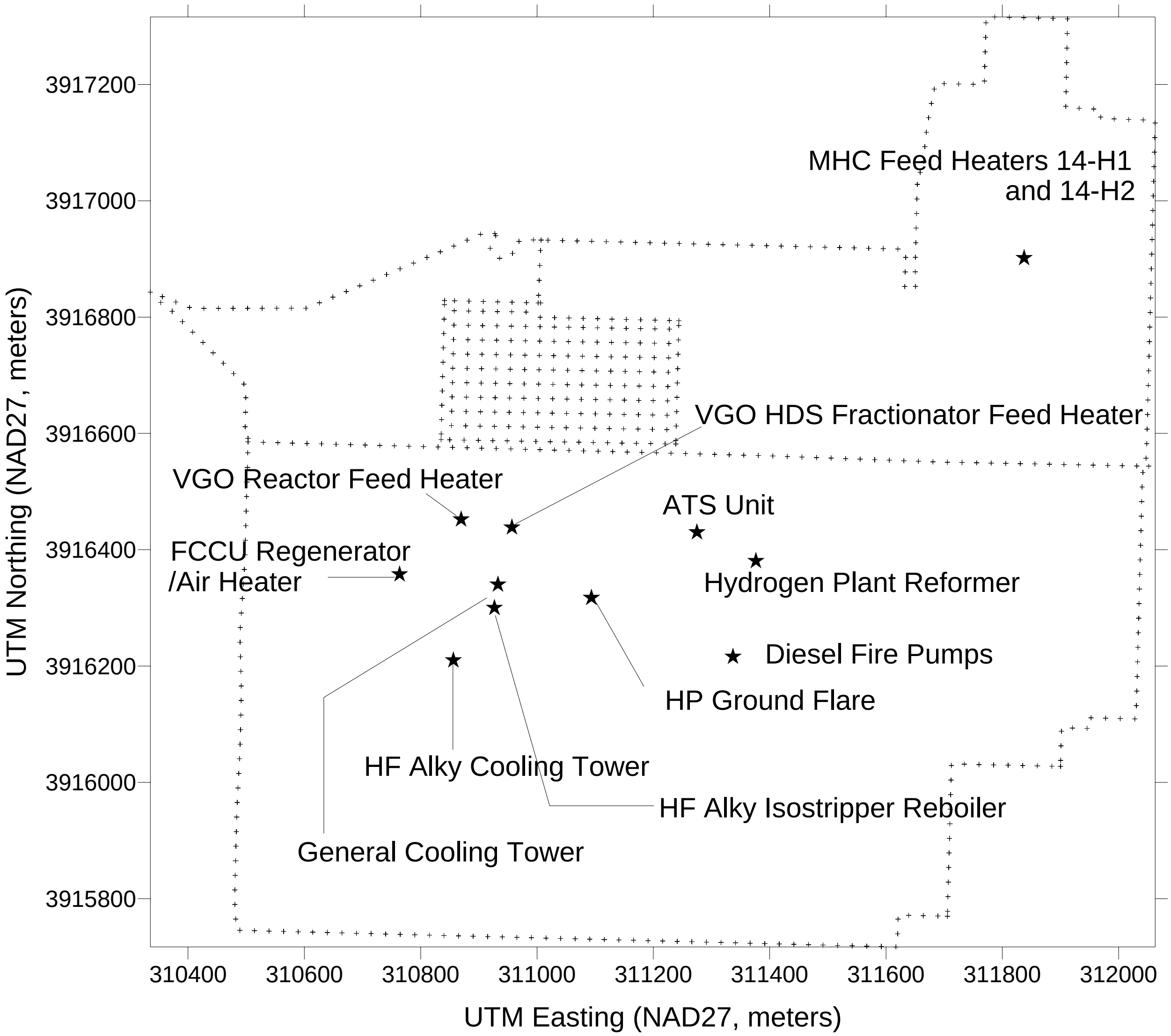

Figure 5-2: Big West – Boundary Receptors and Source Locations

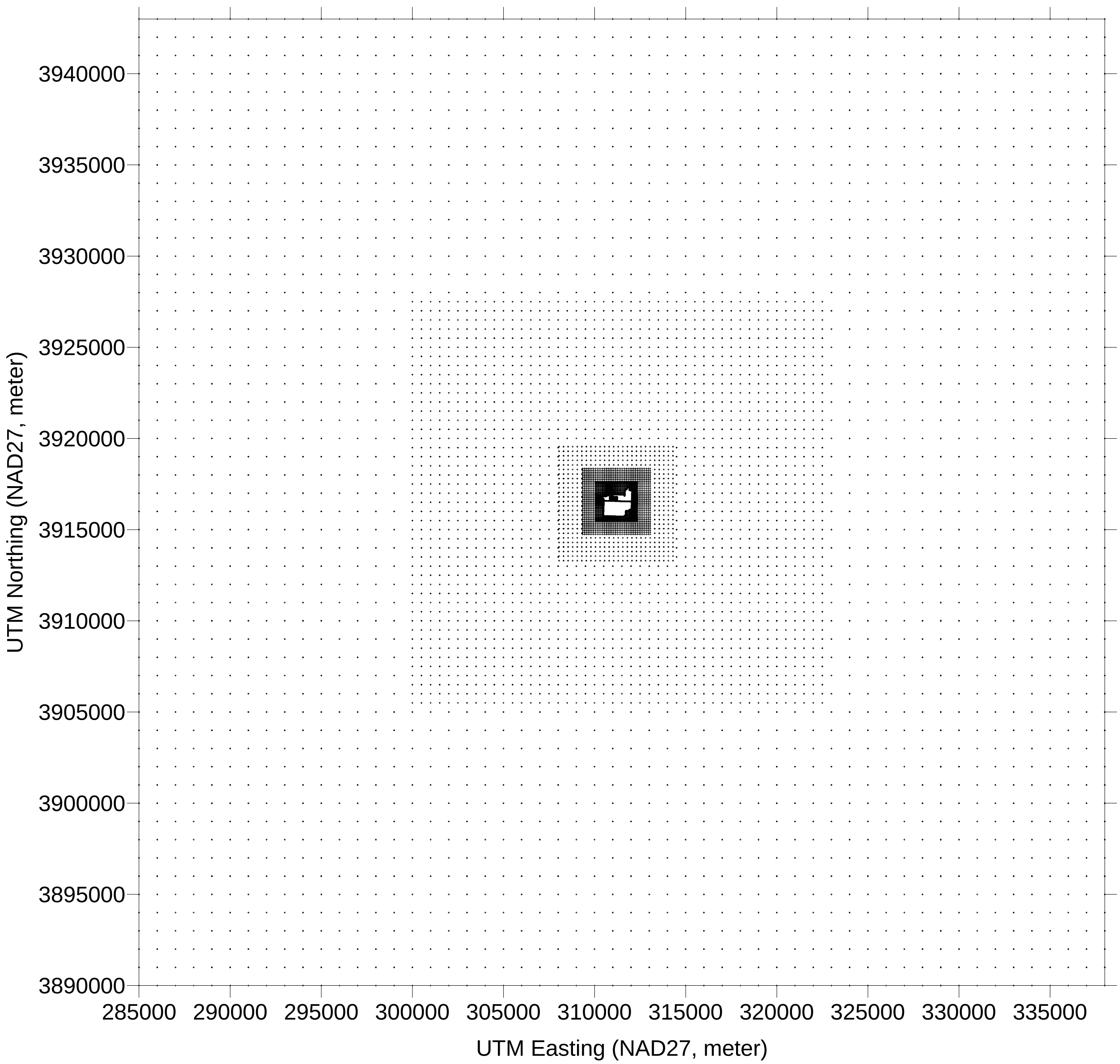

Figure 5-3: Big West Far Grid of Receptors

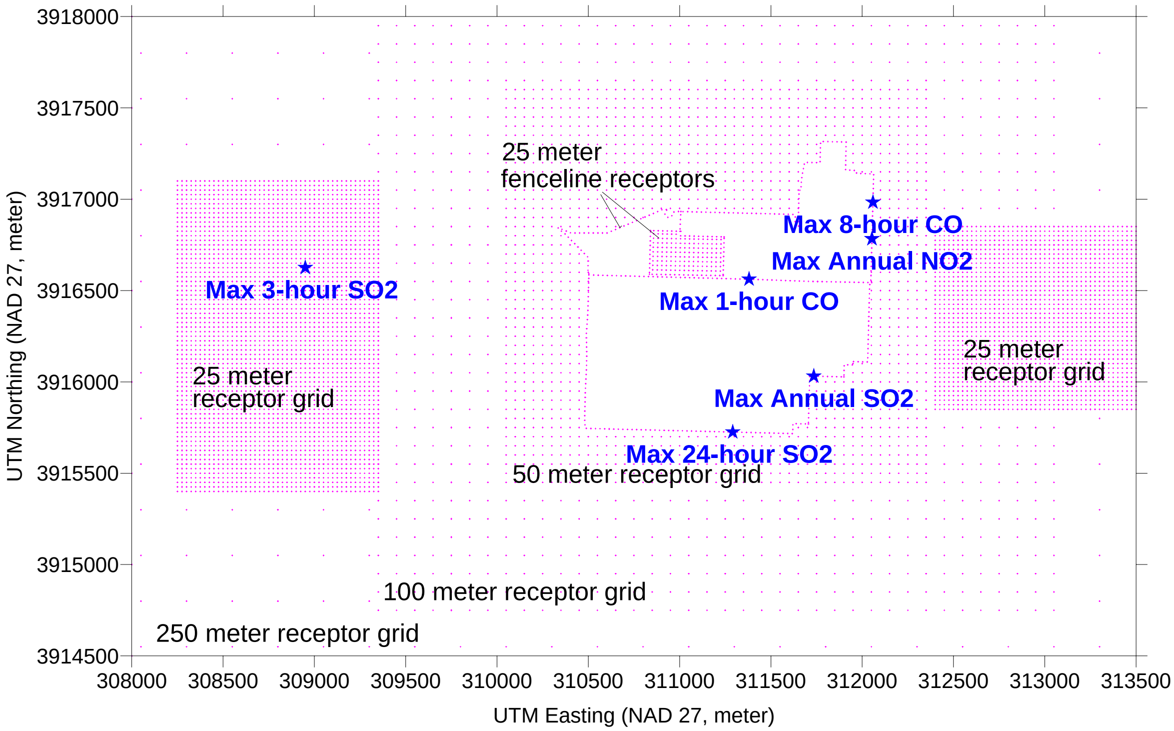

Figure 5-5: Maximum Impact Locations with Fine Grid for Maximum 3-hour SO2 Impact

Table 5-9: Maximum Project Impacts Compared with Class II Significance Levels and Monitoring Significance Levels[a]

Pollutant	Averaging Period	Maximum Predicted Impact (µg/m³)	Class II Significance Level (µg/m³)	Monitoring Significance Level (µg/m³)
NO_2	Annual	~~0.56~~[b] 0.68	1.0	14
SO_2	Annual	~~0.74~~ 0.83	1.0	NA
	3-hour	~~10.67~~ 10.71	25.0	NA
	24-hour	~~3.18~~ 3.37	5.0	13
CO	1-hour	~~181.31~~ 183.42	2,000	NA
	8-hour	~~44.08~~ 31.38	500	575

Notes:

[a] Strikeout values are from the December 2006 revised PSD application.

[b] EPA default Ambient Ratio Method factor of 0.75 applied.

NA = Not applicable/not defined

Cost Effectiveness of SCR on VGO-HDS Heaters

The BACT analysis presented in section 4.2.1 of the December 2006 revised PSD application concluded that BACT for NOx for refinery combustion units less than 50 MMBtu/hr is the installation of low NOx burners to achieve a NOx emission limit of 20 ppmv @ 3% O_2. This conclusion was reached with the following reasoning:

- The most stringent limit found to be achieved in practice or required by a state implementation plan (SIP) was 25 ppmv @ 3% O_2.
- The lowest vendor guarantee that the refinery was able to secure for state-of-the-art low NOx burners on a refinery heater of this size is 20 ppmv @ 3% O_2.
- While the addition of SCR would be technically feasible and could achieve lower NOx emissions, this is not achieved in practice on a small refinery heater and is not considered cost effective; the cost effectiveness, calculated at $13,766 and $12,779 per ton of NOx control for the smaller and larger heaters, respectively, was far above the SJVAPCD's cost effectiveness threshold for NOx control.

You and your staff have indicated that EPA may not agree with the cost effectiveness thresholds as established by SJVAPCD and requested a more complete accounting of costs associated with the installation of SCR. Because the design of the project has progressed since the cost estimates were initially prepared over a year ago, and detailed cost estimates have been obtained for the other CFP heaters, a more complete cost estimate can now be provided. As we have explained to your staff, a more detailed cost analysis was not provided in the December 2006 revised PSD application because the estimated cost-effectiveness exceeded the SJVAPCD BACT cost-effectiveness thresholds. We note that the addition of SCR units on the VGO-HDS heaters this late in the project design would significantly increase these costs – unit redesign/placement, re-engineering, cancellation charges for parts already ordered, etc. – *none* of these schedule- and redesign-related costs have been included. Only incremental costs between installation of low NOx burners alone and installation of low NOx burners and SCR have been included in our revised cost-effectiveness analysis.

Revised Tables C-4 are attached, which provide the new cost estimates and cost effectiveness calculations. As before, guidance from the EPA OAQPS Cost Manual, 6^{th} Edition, Chapter 2 regarding cost estimates for SCR was followed, except where more specific data were available. Cost effectiveness estimates for the smaller and larger heaters were $45,170 and $39,450 per ton, respectively.

In summary, there are no existing refinery heaters or boilers <50 MMBtu/hr that are permitted to achieve NOx emission rates lower than proposed here, the proposed NOx emission limits for the VGO heaters are more stringent than any applicable SIP or proposed NSPS Subpart Ja requirements, and the cost effectiveness for SCR control is significantly more expensive than the BACT cost effectiveness thresholds required for refinery units or similar sources in California or elsewhere in the United States (under EPA, South Coast, Bay Area or San Joaquin Valley air district guidelines). We therefore conclude that SCR controls as applied to the VGO-HDS heaters are not representative of the lowest achievable NOx emission rate.

Proposed NSPS Subpart Ja

On May 17, 2007, EPA proposed amendments to the New Source Performance Standards for Petroleum Refineries (Subpart Ja and proposed modifications to Subpart J, see 72 FR 27178, 5/17/2007). NSPS requirements are effective based on the date of proposal; therefore, affected facilities in the Clean Fuels Project will have to comply with these requirements. We note that the proposed rule is subject to review and comment, and may be modified by EPA in light of these comments. Nevertheless, as demonstrated in the table below, the affected units under CFP can comply with Subpart Ja requirements with the following changes to the proposed project:

Unit	Proposed NSPS Ja Requirement	CFP Controls/Design
FCCU	PM: 0.5 lb/1,000 lb coke burn-off	Same
	PM Monitoring: Method 5 performance test; PM CEMS or control device operating parameter monitoring	Proposed continuous opacity monitoring; will incorporate Ja monitoring requirements (PM CEMS or parameter monitoring)
	NOx: 80 ppmv (dry, 0% O_2) 7-day rolling average	More stringent: 40 ppmv (dry, 0% O_2) daily average; 20 ppmv (dry, 0% O_2) 365-day rolling average
	NOx Monitoring: CEMS	Same
	SO_2: 50 ppmv (dry, 0% O_2) 7-day rolling average; 25 ppmv (dry, 0% O_2) 365-day rolling average	More stringent: 50 ppmv (dry, 0% O_2) daily average; 20 ppmv (dry, 0% O_2) 365-day rolling average
Claus Sulfur Recovery Plant	Provides new SO_2 and H_2S emissions limits	Not applicable to SWAATS unit
Process Heater and Other Fuel Gas Combustion Device	NOx: 80 ppmv (dry, 0% O_2) 24-hour rolling average	More stringent (<20 ppmv @ 3% O_2 15 minute average for CEMS)
	NOx Monitoring: CEMS	Proposed periodic sampling on VGO-HDS units to verify compliance; will install CEMS
	SO_2: 20 ppmv (dry, 0% O_2) 3-hour rolling average – or fuel gas limit of 160 ppmv H_2S 3-hour rolling average; 8 ppmv (dry, 0% O_2) 365-day rolling average – or fuel gas limit of 60 ppmv H_2S 365-day rolling average	More stringent: 40 ppmv total sulfur limit, expressed as H_2S on a 4-hour average

As summarized above, the current BACT emission limits for the CFP units are equal to or more stringent than the proposed Subpart Ja requirements; however, a new short-term NOx standard will be incorporated to address the proposed Ja standard for new FCCUs. Further, additional monitoring will be required to meet the proposed Subpart Ja monitoring requirements; specifically, installation of NOx CEMS for the VGO-HDS heaters and installation of a PM CEMS or parametric monitoring of PM emissions from the FCCU.

Other Issues

Separately, in recent email correspondence, Ms. Kathleen Stewart raised several issues regarding the ground flare, for which our responses are provided below:

- The flare is designed to handle only process upset gases, taken to include gases released during startup, shutdown and malfunctions; it therefore is not designed to comply with Subpart J or proposed Subpart Ja. Furthermore, it will not be permitted to handle releases subject to Subpart J or Ja under its federally enforceable operating permit;
- The minimum heat content of gases that will vent to the flare during process upset conditions, startups and shutdowns will be 300 Btu/scf. We do not anticipate any instance where the heat content of process upset gases will lower than 300 Btu/scf;
- No pressure relief devices will vent directly to the flare;
- The presence of a pilot flame on the ground flare will be monitored with thermocouples, which will record temperature, and hence, the presence of a pilot flame;
- The flow of gases released to the ground flare will be monitored with a GE Sensing ultrasonic flow meter (product brochure and technical data are attached). Please note that this unit does not require daily calibrations; Big West will calibrate the unit consistent with manufacturer’s recommendations; flow accuracy and repeatability data are provided in the brochure;
- The heat content and sulfur content of gases released to the flare during process upset conditions will be monitored by a sampling system that will consist of evacuated cylinders, which will sample gases during a release. We are working with vendors to define the specifications for this sampling system, which we understand is used by other facilities in California and required under the Motiva Consent Decree to measure heat content and sulfur concentrations in gases that are released to a flare. The sample gas obtained from the automated sampling system will then be analyzed for heat content (ASTM Method D2382-88; D3588-91 or D4891-89) and sulfur content (EPA Method 15/16 GC-FPD or equivalent); *and*

- Finally, we wish to confirm that the sulfur content of the torch oil used in the startup of the FCCU will be verified through testing.

A CD of the modeling input and output files will be sent to you under separate cover; copies will be provided directly to Ms. Carol Bohnenkamp as well. Please contact Mr. Everard Ashworth of Ashworth Leininger Group (805.370.1469) to discuss any concerns or questions raised by this letter. Thank you again for your continued assistance on this important project. Thank you again for your continued assistance on this important project.

Very truly yours,

Eugene Cotten
Vice President-Refining

Enclosure

cc: Carol Bohnenkamp, USEPA
Kathleen Steward, USEPA
Vince Memmott, P.E., Flying J Inc.
Bill Chadick, HSE Director, Big West
Everard Ashworth, ALG
Richard Karrs, SJVAPCD
Leonard Scandura, SJVAPCD
Perry Fontana, QEP, URS
Mike McCorison, US Forest Service

Table C-4: BACT Annual Cost Analysis – Refinery Combustion Units <50 MMBtu/hr

Cost Estimate for SCR

EPA OAQPS Cost Manual, 6th Ed., Chapter 2

Design/Operating Parameter	Identifier	Formula/Source	Value
Heater Name		47 MMBtu/hr VGO-HDS Feed Heater	
Heater Size (MMBtu/hr)	Q_B	Heater design	47
Uncontrolled NOx concentration (lb/MMBtu)	NOx_{in}	AP-42 Table 1.4-1, uncontrolled	0.098
NOx Removal Efficiency (%)	η_{NOx}	$=(NOx_{in}-NOx_{out})/NOx_{in}$	94%
Controlled NOx Emission Factor (lb/MMBtu)	NOx_{out}	5 ppmv NOx	0.006
Operating Time (hr/yr)	t_{op}	Full time operation	8760
Equipment Life (years)	n	EPA Guidance	20
NH_3 Cost ($/ton)	$Cost_{NH3}$	URS	$ 320
NH_3 Flow rate (lb/hr)	q_{NH3}	URS	1.0
NH_3 Storage Volume (gal)	Vol_{NH3}	URS	250.0
Anhydrous ammonia specific gravity	SG_{NH3}	IAG	0.620
Cost of Electricity ($/kWh)	$Cost_{elect.}$	PG&E	$ 0.10
Catalyst Operating Life (hours)	$h_{catalyst}$	Manufacturer guarantee of 5 years	43,800
Catalyst Volume (ft^3)	$Vol_{catalyst}$	URS	56.8
Catalyst Replacement Cost ($/$ft^3$)	$CC_{replace}$	URS	$ 271
Annual Interest Rate (%)	i	EPA Guidance	7.0%

Description	Identifier	Formula/Source	Value
Direct Capital Costs			
SCR Equipment Cost	SCR	Vendor (quote does not include foundation, piping, structural elements, etc as itemized below)	$ 388,720
Differential, Cylindrical vs. Box Heater	Cyl	Based on vendor quote, 4/13/06	$ 257,000
Modify Conv Sect for SCR	Mod	IAG estimate	$ 57,320
Air Preheat	Preheat	Based on previous purchase price	$ 553,390
Forced and Induced Draft Fans	Fan	IAG estimate	$ 47,748
CEMS Building	CEM	IAG estimate	$ 250,000
Estimated Equip. Escalation	Esc	IAG estimate/recent experience	$ 155,418
Total Equipment Costs	Equip	= SCR + Cyl + Mod + Preheat + Fan + CEM	**$ 1,709,596**
Concrete (Materials)	Install1	IAG estimate	$ 118,358
Structural steel (Materials)	Install2	IAG estimate	$ 172,157
Piping (Materials)	Install3	IAG estimate	$ 430,394
Electrical (Materials)	Install4	IAG estimate	$ 161,398
Control Systems (Materials)	Install5	IAG estimate (connection to DCS)	$ 118,358
Paint and Insulation (Materials)	Install6	IAG estimate	$ 53,799
Construction/Installation Labor	Install7	IAG estimate	$ 1,616,089
Total Installation Costs	Install	=Install1 + Install2 + Install3 + Install4 + Install5 + Install6 + Install7	**$ 2,670,553**
Total Direct Capital Cost	A	=Equip + Install	$ 4,380,149
Indirect Capital Costs			
Demo	Demo	IAG cost estimate	$ 0
Site work and civil	Site	IAG cost estimate	$ 88,857
Engineering costs	Eng	IAG cost estimate	$ 734,000
Construction/Installation Labor (Indirect)	IndInstall	Included in Direct Construction/Installation Labor costs above	$ 0
Total Indirect Installation Costs	B	= Demo + Site + Eng + IndInstall	**$ 822,857**
Project Contingency	C	= (A + B) x 0.15	$ 780,451
Total Plant Cost	D	= A + B + C	**$ 5,983,457**
Allowance for Funds During Construction	E	Assumed 0 for SCR	$ 0
Royalty Allowance	F	Assumed 0 for SCR	$ 0
Inventory Capital (ammonia stored at site, i.e., first fill of reagent tanks)	G	= Vol_{NH3} x ($Cost_{NH3}$/2000 lb/ton) x SG_{NH3} x 8.345 lb/gal	$ 207
Initial Catalyst and Chemicals	H	Assumed 0 for SCR	$ 0
Total Capital Cost			
Total Capital Investment	TCI	= D + E + F + G + H	**$ 5,983,664**
Direct Annual Costs			
Operating and Supervisory Labor	L	IAG estimate	$ 15,234
Maintenance	M	= 0.015 x TCI	$ 89,755
Reagent Consumption	RC	= q_{NH3} x ($Cost_{NH3}$/2000 lb/ton) x t_{op}	$ 1,373
Utilities	U	= P x $Cost_{elect.}$ x t_{op}	$ 74,753
- Power Needed (kW)	P	Vendor estimate	85.33
Annual Catalyst Replacement Cost	ACR	= $Vol_{catalyst}$ x $CC_{replace}$ x FWF	$ 2,675
- Catalyst Replacement Term (years)	Y	$=h_{catalyst}/t_{op}$	5
- Future Worth Factor	FWF	$= i/((1+i)^Y - 1)$	0.1739
Total Direct Annual Cost	DAC	= L + M + RC + U + ACR	**$ 183,790**
Indirect Annual Costs			
Property Taxes	PT	Assumed 0 for SCR	$ 0
Insurance Costs	IC	Assumed insignificant for SCR	$ 0
Administrative Charges	AC	Assumed insignificant for SCR	$ 0
Overhead	OH	Assumed insignificant for SCR	$ 0
Capital Recovery Costs	CRC	= CRF x TCI	$ 564,816
- Capital Recovery Factor	CRF	$= i/(1-(1+i)^{-n})$	0.0944
Total Indirect Annual Cost	IAC	= PT + IC + AC + OH + CRC	**$ 564,816**
Total Annual Cost			
Total Annual Cost	TAC	=DAC + IAC	**$ 748,606**

Total NOx Removed (tpy)	ΔE	$=(1/(1-\eta_{NOx}) - 1)$ x E	18.9
Cost Effectiveness of NOx Removal ($/ton)		= TAC/ΔE	**$ 39,541**

Table C-4: BACT Annual Cost Analysis – Refinery Combustion Units <50 MMBtu/hr (cont.)

Cost Estimate for SCR

EPA OAQPS Cost Manual, 6th Ed., Chapter 2

Design/Operating Parameter	Identifier	Formula/Source	Value
Heater Name		35 MMBtu/hr VGO-HDS Fractionator Feed Heater	
Heater Size (MMBtu/hr)	Q_B	Heater design	35
Uncontrolled NOx concentration (lb/MMBtu)	NOx_{in}	AP-42 Table 1.4-1, uncontrolled	0.098
NOx Removal Efficiency (%)	η_{NOx}	$=(NOx_{in}-NOx_{out})/NOx_{in}$	94%
Controlled NOx Emission Factor (lb/MMBtu)	NOx_{out}	5 ppmv NOx	0.006
Operating Time (hr/yr)	t_{op}	Full time operation	8760
Equipment Life (years)	n	EPA Guidance	20
NH_3 Cost ($/ton)	$Cost_{NH3}$	URS	$ 320
NH_3 Flow rate (lb/hr)	q_{NH3}	URS	0.8
NH_3 Storage Volume (gal)	Vol_{NH3}	URS	250.0
Anhydrous ammonia specific gravity	SG_{NH3}	IAG	0.620
Cost of Electricity ($/kWh)	$Cost_{elect.}$	PG&E	$ 0.10
Catalyst Operating Life (hours)	$h_{catalyst}$	Manufacturer guarantee of 5 years	43,800
Catalyst Volume (ft^3)	$Vol_{catalyst}$	URS	42.3
Catalyst Replacement Cost ($/$ft^3$)	$CC_{replace}$	URS	$ 271
Annual Interest Rate (%)	i	EPA Guidance	7.0%

Description	Identifier	Formula/Source	Value
		Direct Capital Costs	
SCR Equipment Cost	SCR	Vendor (quote does not include foundation, piping, structural elements, etc as itemized below)	$ 320,935
Differential, Cylindrical vs. Box Heater	Cyl	Based on vendor quote, 4/13/06	$ 212,000
Modify Conv Sect for SCR	Mod	IAG estimate	$ 42,680
Air Preheat	Preheat	Based on previous purchase price	$ 456,891
Forced and Induced Draft Fans	Fan	IAG estimate	$ 39,422
CEMS Building	CEM	IAG estimate	$ 250,000
Estimated Equip. Escalation	Esc	IAG estimate/recent experience	$ 132,193
Total Equipment Costs	Equip	= SCR + Cyl + Mod + Preheat + Fan + CEM	**$ 1,454,121**
Concrete (Materials)	Install1	IAG estimate	$ 100,680
Structural steel (Materials)	Install2	IAG estimate	$ 146,444
Piping (Materials)	Install3	IAG estimate	$ 366,111
Electrical (Materials)	Install4	IAG estimate	$ 137,292
Control Systems (Materials)	Install5	IAG estimate (connection to DCS)	$ 100,680
Paint and Insulation (Materials)	Install6	IAG estimate	$ 45,764
Construction/Installation Labor	Install7	IAG estimate	$ 1,374,712
Total Installation Costs	Install	=Install1 + Install2 + Install3 + Install4 + Install5 + Install6 + Install7	$ 2,271,683
Total Direct Capital Cost	A	=Equip + Install	$ 3,725,804
		Indirect Capital Costs	
Demo work	Demo	IAG cost estimate	$ 0
Site work and civil	Site	IAG cost estimate	$ 75,586
Engineering costs	Eng	IAG cost estimate	$ 624,000
Construction/Installation Labor (Indirect)	IndInstall	Included in Direct Construction/Installation Labor costs above	$ 0
Total Indirect Installation Costs	B	= Demo + Site + Eng + IndInstall	**$ 699,586**
Project Contingency	C	= (A + B) x 0.15	$ 663,808
Total Plant Cost	D	= A + B + C	**$ 5,089,198**
Allowance for Funds During Construction	E	Assumed 0 for SCR	$ 0
Royalty Allowance	F	Assumed 0 for SCR	$ 0
Inventory Capital (ammonia stored at site, i.e., first fill of reagent tanks)	G	$= Vol_{NH3}$ x $(Cost_{NH3}/2000$ lb/ton) x SG_{NH3} x 8.345 lb/gal	$ 207
Initial Catalyst and Chemicals	H	Assumed 0 for SCR	$ 0
		Total Capital Cost	
Total Capital Investment	TCI	= D + E + F + G + H	**$ 5,089,405**
		Direct Annual Costs	
Operating and Supervisory Labor	L	IAG estimate	$ 15,234
Maintenance	M	= 0.015 x TCI	$ 76,341
Reagent Consumption	RC	$= q_{NH3}$ x $(Cost_{NH3}/2000$ lb/ton) x t_{op}	$ 1,133
Utilities	U	= P x $Cost_{elect.}$ x t_{op}	$ 61,718
- Power Needed (kW)	P	Vendor estimate	70.45
Annual Catalyst Replacement Cost	ACR	$= Vol_{catalyst}$ x $CC_{replace}$ x FWF	$ 1,992
- Catalyst Replacement Term (years)	Y	$=h_{catalyst}/t_{op}$	5
- Future Worth Factor	FWF	$= i/((1+i)^Y - 1)$	0.1739
Total Direct Annual Cost	DAC	= L + M + RC + U + ACR	**$ 156,419**
		Indirect Annual Costs	
Property Taxes	PT	Assumed 0 for SCR	$ 0
Insurance Costs	IC	Assumed insignificant for SCR	$ 0
Administrative Charges	AC	Assumed insignificant for SCR	$ 0
Overhead	OH	Assumed insignificant for SCR	$ 0
Capital Recovery Costs	CRC	= CRF x TCI	$ 480,404
- Capital Recovery Factor	CRF	$= i/(1-(1+i)^{-n})$	0.0944
Total Indirect Annual Cost	IAC	= PT + IC + AC + OH + CRC	**$ 480,404**
		Total Annual Cost	
Total Annual Cost	TAC	=DAC + IAC	$ 636,823

Total NOx Removed (tpy)	ΔE	$=(1/(1-\eta_{NOx}) - 1)$ x E	14.1
Cost Effectiveness of NOx Removal ($/ton)		= TAC/ΔE	**$ 45,169**