H.C. Shah.

INTER-OFFICE MEMO **TENNECO CHEMICALS, INC.**

TO	Mr. C. G. Thompson	AT	Piscataway	DATE	September 25, 1974
FROM	Dr. P. A. Lobo	AT	Piscataway	COPY TO	Mr. J. Fisher Dr. R. T. Gottesman Mr. R. S. Miller Mr. W. Miringoff
SUBJECT	SCALEUP OF PILOT PLANT RESULTS - VCM STRIPPING OF SLURRY				

We have all been concerned about our inability to duplicate laboratory and pilot plant results on stripping of VCM from PVC slurry in the Burlington plant. The attached article points out the difficulty in translation of pilot results to the plant and reinforces our thinking that power input maybe too low in the plant strippers to afford good mass transfer. I understand that tests are underway in the pilot plant to determine the effect of horsepower on stripping efficiency.

P. A. Lobo

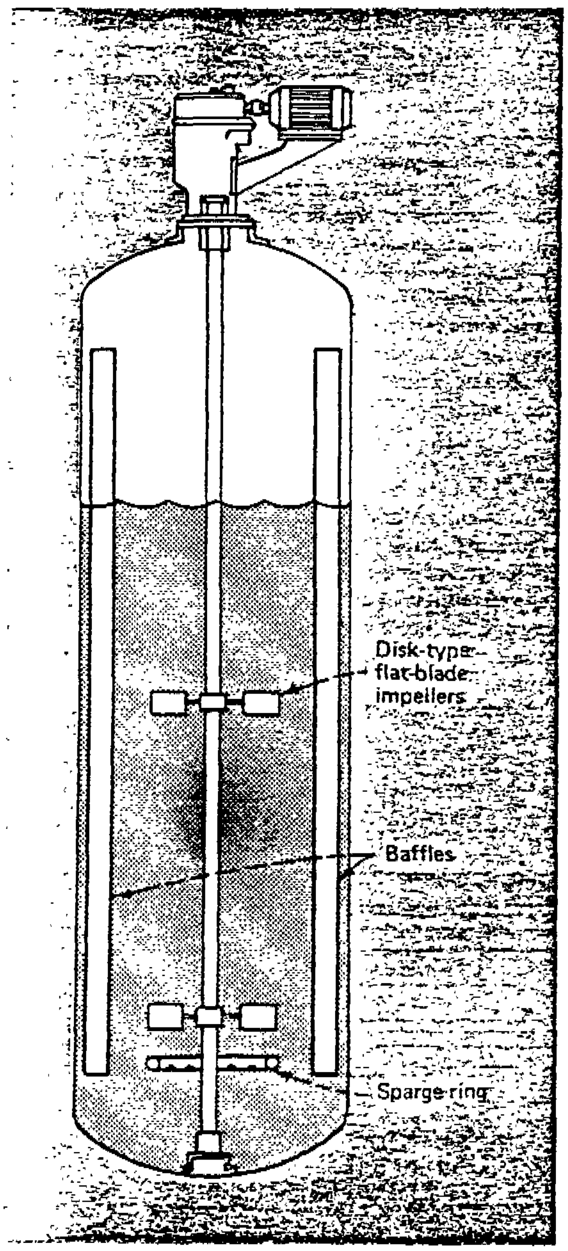
/jg
Attachment

COLORITE 017164

Scaleup of Gas-Dispersion Mixers

Gas-dispersion mixers are widely used in fermentation processes. Here are the problems encountered in scaling up laboratory data.

GEORGE H. LEAMY, Philadelphia Mixer Corp.



Dispersing gas into a liquid is an important industrial mixing application. The role of the mixer is to develop sufficient liquid flow to disperse the gas throughout the system, and to provide sufficient shear for optimum gas-interfacial area to give the greatest mass-transfer coefficients.

With the advent of new pharmaceuticals produced by fermentation (during and after World War II), the growth in gas-dispersion mixers became a prime factor in the industrial mixer field. Many of the early mixers were hit-or-miss devices: little thought was given to gas loading factors, optimum impeller-tip speeds, and total horsepower requirements.

After the initial flurry of activity died down, a period of evaluation and optimum-mixer selection followed. Most of these early evaluations were carried on independently by a number of pharmaceutical companies—this resulted in widely varying mixer selections being made by different users for what, many times, were identical applications.

During the last 5 to 10 years, a compilation of data from both external sources and laboratory work has been made available. These data, combined with practical information gathered from full-size mixers operating in the field, have led to meaningful mixer selections for gas dispersion.

The usual approach to production-size fermentation tanks has been to scale up from a laboratory test vessel. This has often proved to be a difficult step—not because correct scaleup is unpredictable, but because of a misunderstanding of the factors important to the proper transition from laboratory mixer tests to full-size production vessels.

Analysis of Scaleup Factors

Investigation into the scaleup of a gas-dispersion system must take a number of related variables into account. For a gas-dispersion system, shear stress, pumping rate and horsepower per unit volume are the variables most important on scaleup to full size.

$$Z \propto kN^3D^5 \quad (1)$$

$$Q \propto kND^3 \quad (2)$$

$$\dot{\gamma}^2 \propto Z/\mu \quad (3)$$

The value of k in the case of mixer horsepower (Z) must

Nomenclatur

C	Impeller off-bottom distance, ft.
D	Impeller dia., ft.
h	Tank height, ft.
k	Mixer constant varying with impeller configuration and overall liquid system. For gas-dispersion operations, values generally range from 2.5 to 6.5.
k_{La}	Overall liquid phase mass-transfer coefficient, lb. mole/(hr.) (sq. ft.) (lb. mole/cu.ft.)
N	Impeller speed, rev./min.
N_p	Impeller power number, kP_i/D^5N^3, dimensionless
N_{Re}	Reynolds number, dimensionless
P	Impeller power in ungassed condition, hp.
P_g	Impeller power in gassed condition, hp.
P_i	Impeller horsepower
Q	Mixer pumping rate, cu.ft./min.
T	Tank dia., ft.
Z	Mixer power, hp.
$\dot{\gamma}$	Mixer shear rate, sec.$^{-1}$
μ	Liquid viscosity, cp.
ρ	Liquid density, lb./cu.ft.

be evaluated for both the type of impeller and the overall liquid system. The same will hold true for k in Eq. (2). Since shear rate ($\dot{\gamma}$) in Eq. (3) is dependent on the viscosity and P_i, any changes in P_i will cause changes in the shear rate. Therefore, the maximum shear rate occurring at the impeller tip will increase in scaleup to higher values if impeller peripheral speeds are increased.

Experimentation[1] has shown that gas bubble size is a function of impeller speed. When the impeller operates in a speed range where it begins to approach a Reynolds number of 1,000 or greater, a decrease in bubble size will occur.

Further work[2] indicated that the impeller blade width has no influence on bubble size at constant peripheral speeds. Variations in blade width will, however, give wide variations in the power drawn by these impellers. Increasing the blade width will give higher flowrates from the impeller and a compromise must be reached between acceptable P_i levels and turnover rates in the tank. This turnover rate must be sufficient to bring all of the tank contents to the high-shear zone near the impeller, at a rate greater than the coalescence rate of the dispersed gas.

The writer's experience in evaluating data derived from laboratory-scale gas-dispersion tests from many sources indicates that too often the test has been conducted with too small a vessel; e.g., 1 to 2 gal. Although impeller-tip speeds in these small vessels are kept within acceptable scaleup limits (800 to 2,000 ft./min.), the pumping rate of the impeller in relation to the vessel size will give turnover rates far in excess of any reasonable scaleup figure. This will produce several undesirable side effects that will lead to incorrect data and conclusions.

Small test vessels usually have a liquid height equal to the vessel diameter. As mixing intensity is increased to the point where N_{Re} approaches 20,000, surface aeration of the liquid will occur. At N_{Re} of 100,000 and above, severe induction of surface air will take place. As N_{Re} approaches 10^6, all sparge air can be turned off without any noticeable decrease of the gas interfacial area in the liquid.

This phenomenon is not experienced to any great extent in production-size fermenters since the height-to-diameter ratio is large enough to preclude any appreciable surface aeration.

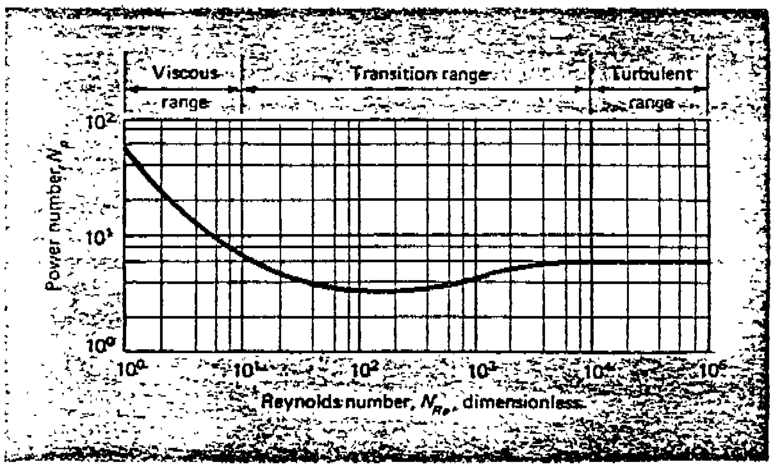
Small vessels also experience high turnover rates that often exceed 50 or more turnovers/min. This produces rapid circulation through the high-shear zone close to the impeller at a rate in excess of the dispersed gas coalescence rate. Since Q (or tank turnover) is proportional to the function of ND, the turbine bubble-size formula:

$$D_p \propto \left(\frac{1}{ND}\right)\frac{h}{T^{1/2}} \tag{4}$$

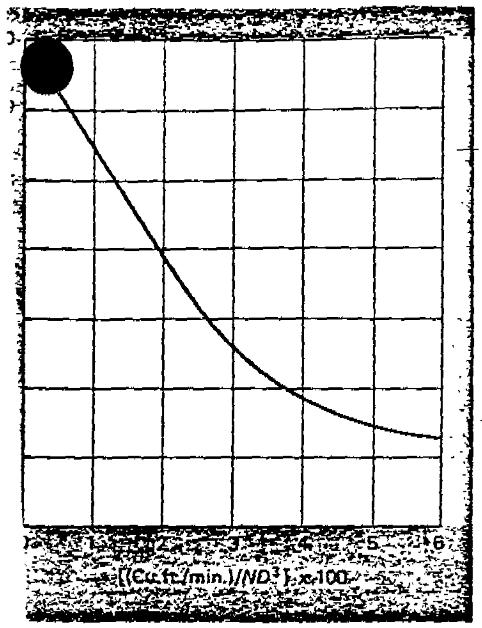
NB

illustrates that any increase in pumping rate will give a decrease in bubble size. Thus, any excessive pumping rate in a small-scale vessel that cannot be reasonably scaled up will result in a gas holdup rate and bubble size decrease beyond the point of scaleup. This results in high mass-transfer rates that will not be obtained on production-sized tanks.

Another factor to consider is the effect of proximity



POWER CURVE for standard baffled tanks is applicable to gas-dispersion operations—Fig. 1



ıion effect on impeller power—Fig. 2

.ler blade tip to the wall of a small test-vessel. ment of high velocity flow on the test vessel luce an added shear rate that will not relate ›sels on scaleup.

dispersion operations are normally done in tanks, the power curve for standard baffled l) ·plicable to this type of operation.

ıe ·e shows that as N_{Re} increases, the flow :s change from viscous to turbulent. Since

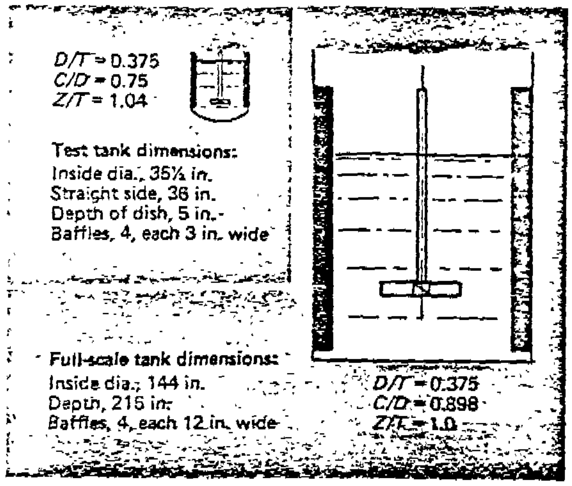
$$N_p = \frac{kP_s}{D^5N^3\rho} \qquad (5)$$

ɔwer number of the impeller will not be l $N_{Re} \geq 10^4$. Thus, when gas-dispersion lab- are conducted in a 1 to 2 gal. vessel, it is event surface induction of gas due to high :s, while at the same time N_{Re} below 10,000 ɔnsistent power levels with changes in im- , since N_{Re} remains in the transition zone.

sumpti n of Impellers

s dispersion in fermentation, industry has ıded, disk-type flat-blade turbine the most of impeller. It combines reasonably high th good pumping rates. Its most important · disk, which forces the gas to flow out to zone at the blades. Although uncommon, ers are sometimes used. Also, adjustable d 8-blade impellers are often specified to n the impeller when the product is varied onditions.

ı of data obtained from actual field units tests has confirmed to a great extent the ·bank[4]. Applied to both large and small cu· ·hown in Fig. 2 has stood the test lot· the correlation between gassed



SCALEUP by direct tank-volume relationship—Fig. 3

and ungassed power as a function of both gas rate and impeller speed and diameter.

Although some previous data have been presented concerning the effect of gassing on multiple-impeller installations, nothing definitive has been presented. The writer's experience, both in laboratory tests and actual installations, indicates a modifying factor of 0.70 to 0.85 of the P_g/P factor for a single impeller should be applied to all impellers located above the lowest impeller. This factor assumes a spacing of at least 1½ impeller diameters between impellers.

An important point often overlooked in fermentation scaleup is bubble residence-time. Usually, scaleup is based on geometric similarity of the vessel and impellers. If, at the same time the superficial gas velocity is held constant, the bubble residence-time increases proportional to the tank size increase. Oxygen depletion of the air will result and has to be considered in scaleup.

The mass-transfer coefficient k_{La} is proportional to the gas volume used, and it is important on scaleup to retain the same or smaller bubble residence-time. This can be accomplished by using a direct tank-volume relationship. This relationship is illustrated in Fig. 3. ■

References

1. Westerterp, K. R., Thesis, Technische Hogeschool, Delft, Netherlands, 1962.
2. Uhl, Vincent and Gray, Joseph B., "Mixing Theory & Practice," Academic Press, New York, 1967
3. Moo-Young, M. B., Ph.D. Thesis, London University, 1961.
4. Calderbank, P. H., *Trans. Inst. Chem. Eng. (London)*, 36, p. 443 (1958).

Meet the Author

George H. Leamy is District Manager of Southeastern Sales for Philadelphia Mixer Corp., Schuylkill Expressway, King of Prussia, PA 19406, where his previous job was that of technical director. He was associated with Mixing Equipment Co. for a number of years as a process engineer and as head of the sample testing laboratory. He received his B.S. in chemical engineering from Clemson University.

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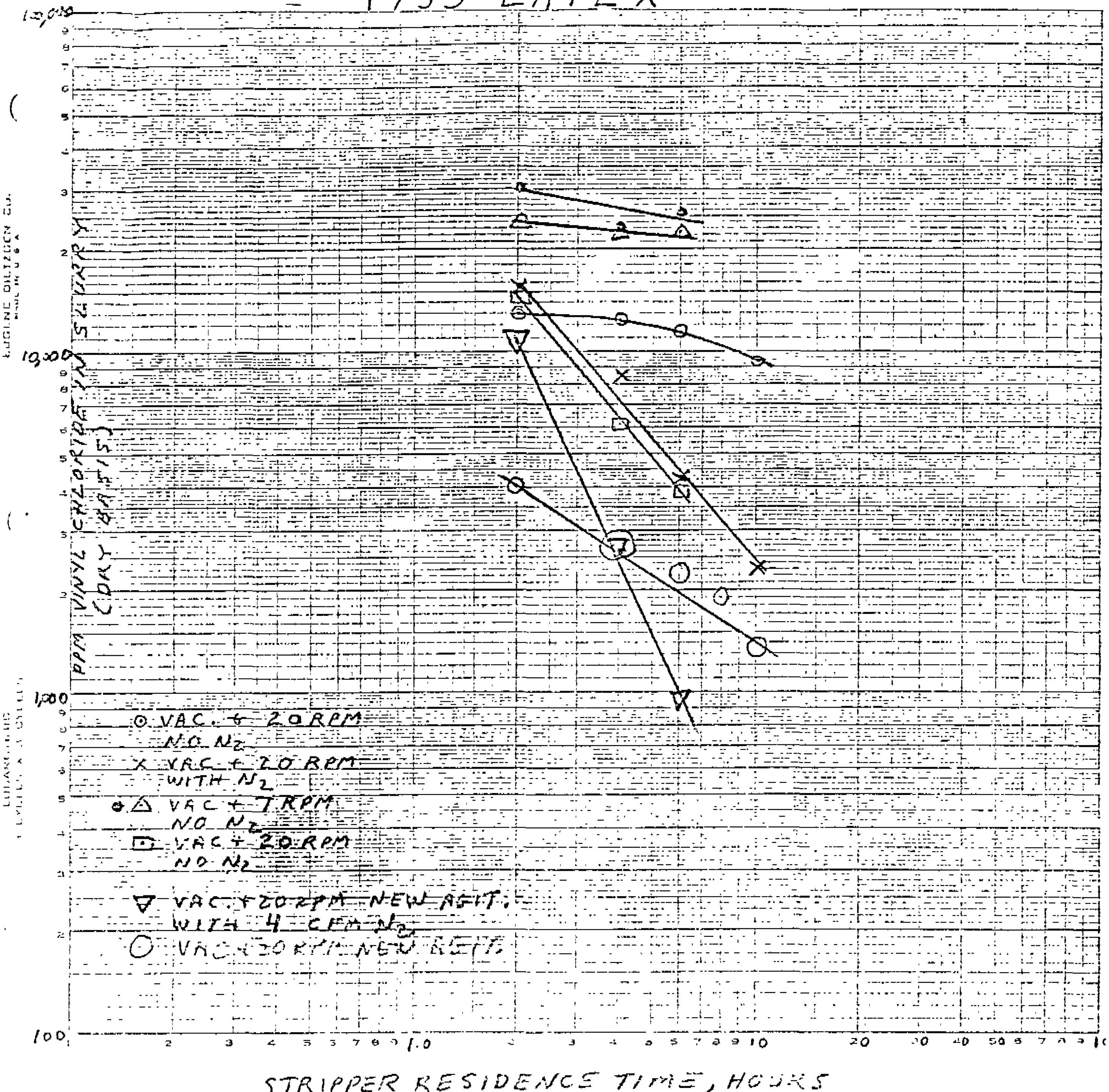
GRAPH III

1755 LATEX



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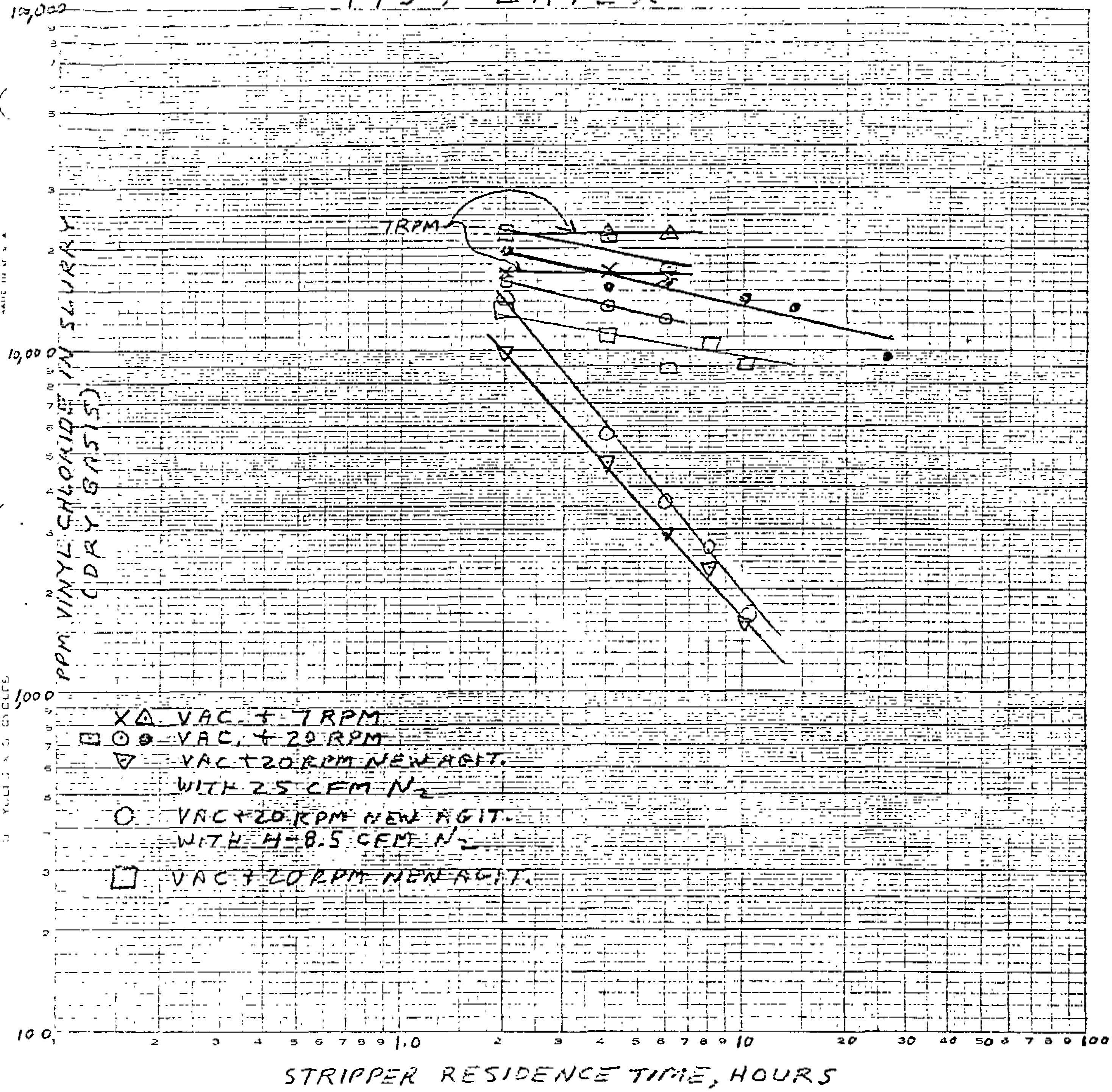
GRAPH IV

1757 LATEX



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(SPEC. D.B. 5.4 - 6.8)

PE DATA 1975

DATE	LOT/BATCH #	FORM. No.	SAMPLE TYPE	CONDITIONS	DRY BLEND TIME	VCM PPM SLURRY	VCM PPM RESIN
/9	275001	3725 17	DRY	55→210°, COOL 190°, +300 GAL	8.4	5.0, 0.5, 4.7 14.5, 35, 6.5	N.D.
/0	275002	" 17	"	"	8.9		0.7
/2	275003	" 17,18	"	"	9.1		N.D.
/5	275004	" 18	"	"	8.5		N.D.
/1	B-250	" 18	SLURRY	"	-	18.3	
/4	B-277	" 18	"	"	-	15.2	
4	B-288	" M18	"	"	-	11.0	
/8	275005	" M19	DRY	55→190° 1½+300 GAL	10.1		1.2
/28	275006	" M19	"	"	9.2		17
/29	275007	" M19	"	"	7.6		9
/30	275008	" M20	"	55→165°, HOLD 30 MIN	7.4		20
/6	275009	" 21	BLEND TK	55→180+200 gal	-	152	
/	275009	" 21	"	"	-	173	
/	B-425	" 21	SLURRY	"	-	384	
	275009	" 21	DRIER	" (1900 HR)	-		16
/	B-429	" 21	SLURRY	"	-	380	
/	B-430	" 21	"	"	-	278	
/	275009	" 21	DRIER	" (0100 HR)	-		
/	275009	" 21	BLEND TK	"	-	118	
/	275009	" 21	DRIER	" (0300 HR)	-		
/6	275009	" 21	"	" (0500 HR)	-		
/7	275009	" 21	DRY	"	6.9		18
/6	275011	" 21	DRY LOT	"	8.6		20
/9	275012	" 21	DRY LOT	"	7.0		46
/10	275013	" 21	COMB. DRY	"	8.0		49
/1	275015	" -22	DRY	55→170° (TRANSFER)	9.1 (HIGH FINES)		134
/2	275015	" 22	DRIER	" (1000 HR)	9.1	"	98
2	275015	" 22	"	" (1145 HR)	9.1	"	140

COLORITE 017170

DATE	LOT/BATCH #	FORM. NO.	SAMPLE TYPE	CONDITIONS	DRY BLEND	VCM PPM SLURRY	VCM PPM RESIN
2/17	275016	3125-23	DRY LOT	SS → 165°F + 200 GAL.	6.3		119
2/17	275017	" -23	DRY	"	6.0		152
2/18	275017	" 23	COMB. DRY	"	6.0		138
2/19	275018	" 23	COMB. DRY	"	6.3		90
2/20	275019	" 23	DRY	"	5.9		102
2/20	275020	" 23	DRY	"	6.2		158
2/21	275021	3125-23	COMB. DRY	"	6.2		107
2/21	275022	"	"	"	6.0		123
2/22	275023	"	"	"	6.5		

250 - 12S

2/21/75
(SPEC. O.B. 8.0 - 9.5)

DATE	LOT/BATCH #	FORM. No.	SAMPLE TYPE	CONDITIONS		DRY BLEND TIME	VCM PPM SLURRY	VCM PPM RESIN
	391628	3122-34	DRY LOT	55→210, HOLD 15 MIN		9.5		N.D.
5	395001	3122-34	"	"		10.6		5.7
5	B-48	3122-34	SLURRY	"		–	123	
5	B-49	3122-34	"	"		–	170	
5	B-38	3122-34	"	"		–	48	
5	B-407	3122-35	"	55→190° 1½	300 MM	–	119	
	395002	3122-35	COMB. LOT	"	"	7.4		28
	395003	3122-35	DRY	"	"	7.7		6.7
	395003	3122-35	COMB. LOT	"	"	7.7		5.9
	395003	3122-35	DRIER	"	"	7.7		19
	395003	3122-35	DRIER	"	"	–		21
	B-415	3122-35	SLURRY	"	"	–	100	
6	B-416	3122-35	"	"	"	–	293	
	395004	3122-36	COMB. LOT	"	500 MM	7.9		28
	395005	3122-36	DRIER	"	"	9.3		15

COLORITE 017172

21.4711

225/16R (SPEC D.8 6-8)

Date	LOT/BATCH #	FORM. No	SAMPLE TYPE	CONDITION	DRY BLEND TIME	VCM (PPM) SLURRY	VCM (PPM) RESIN
1/30	235001	3120-M-46	DRIER	55-190°F ↑↓ 500g	7.7	—	
1/31	235002	3120-M-47	"	55-180°F ↑↓ 400g	8.0	—	39
2/1	235003	"	"	"	8.0	—	
2/2	235004	"	"	"	8.4	—	
2/3	235005	"	"	"	7.7	—	45
2/4	235006	3120-M-48	"	"	7.3	—	43
2/4	B-381	"	SLURRY	"	—	469	—
"	382	"	"	"	—	517	—
"	383	"	"	"	—	355	—
"	384	"	"	"	—	228	—
"	391	"	"	"	—	881	—
"	392	"	"	"	—	764	—
"	395	"	"	"	—	453	—
"	397	"	"	"	—	1726	—
"	BLD. TK	"	"	"	—	660	—
2/5	235007	"	DRIER	"	7.0		

200-2

(SPEC. D.B. → 6-8)

DATE	LOT/BATCH #	FORM. NO.	SAMPLE	CONDITIONS	DRY BLEND TIME	VCM PPM SLURRY	VCM PPM RESIN
1/4	B-402	3439-M-15	SLURRY	55→170°	-	3469	
1/5	195001	"	DRIER	" (2200 HR)	8.4/8.5		753
1/5	195001	"	"	" (2300 HR)	8.4/8.5		807
1/5	195001	"	"	" (0100 HR)	8.4/8.5		864
1/5	B-405	"	SLURRY	"	-	2431	
1/12	195002	3439-M-16	COMB. DRY	"	7.2		1104
1/13	195002	"	"	"	7.2		661
1/13	195003	"	DRY LST	"	7.3		746
1/13	195004	"	"	"	8.0		1092
1/14	195005	3439-M-17	"	55→170° DEC. METH.	7.7		794
1/15	195006	"	COMB. DRY	"	7.1		800

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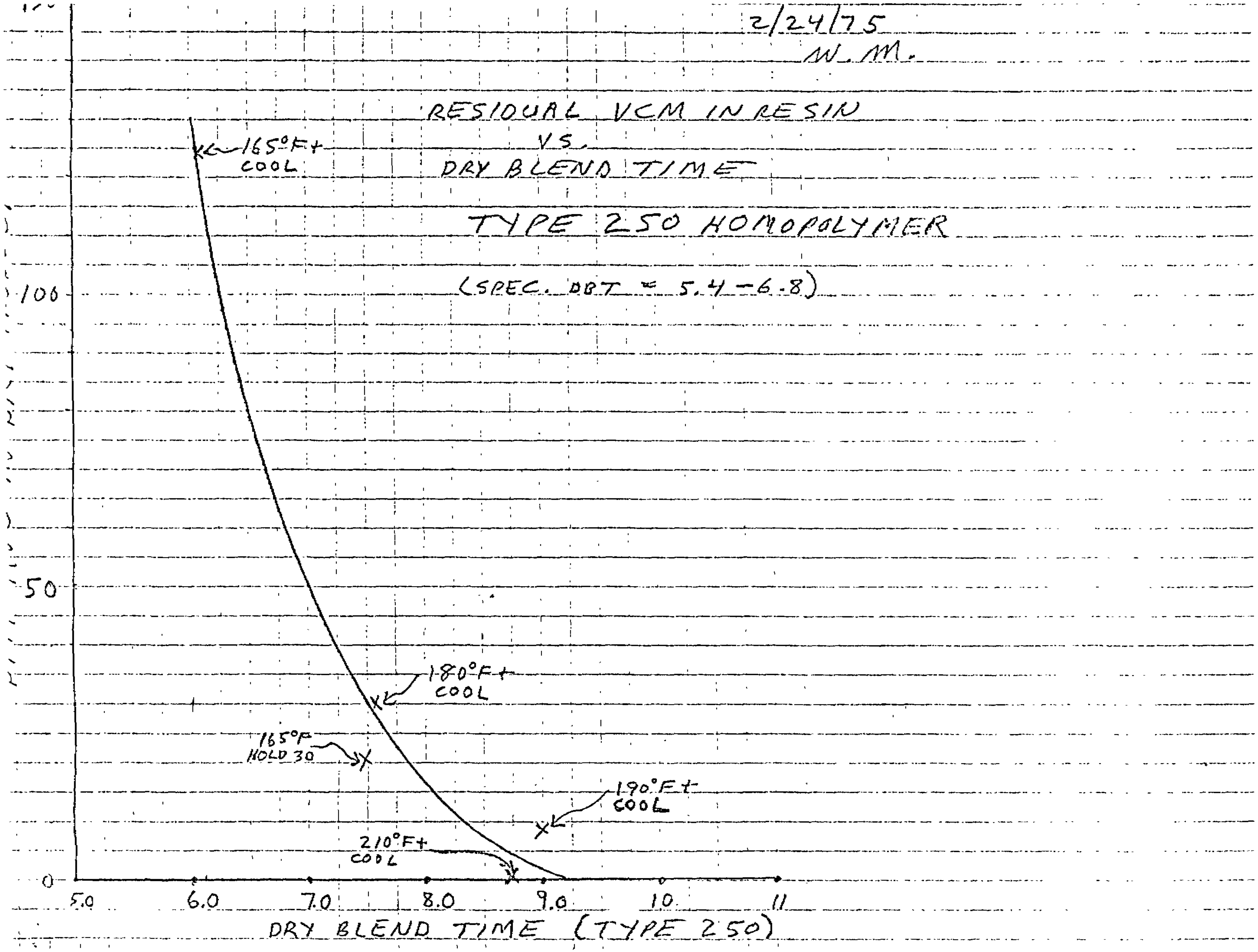
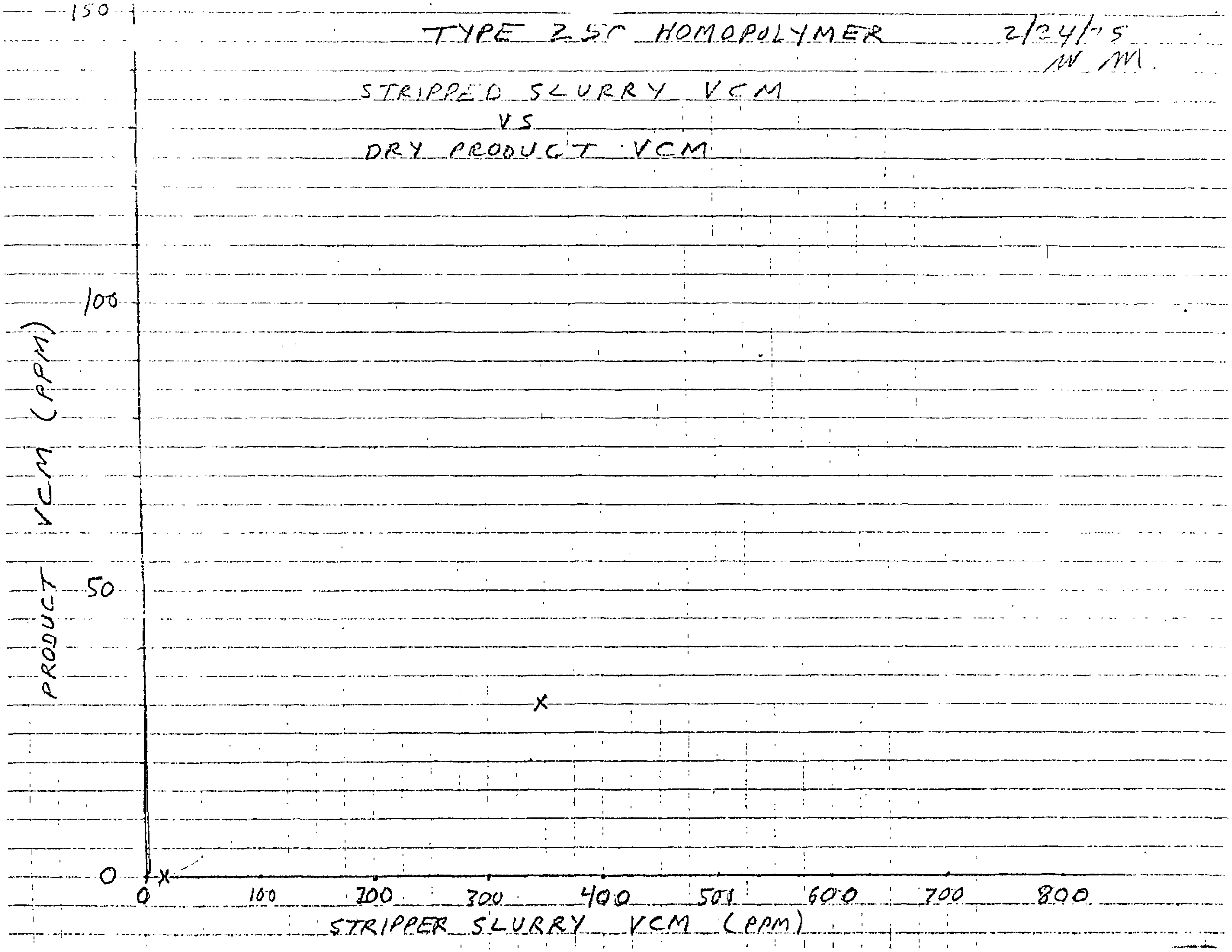
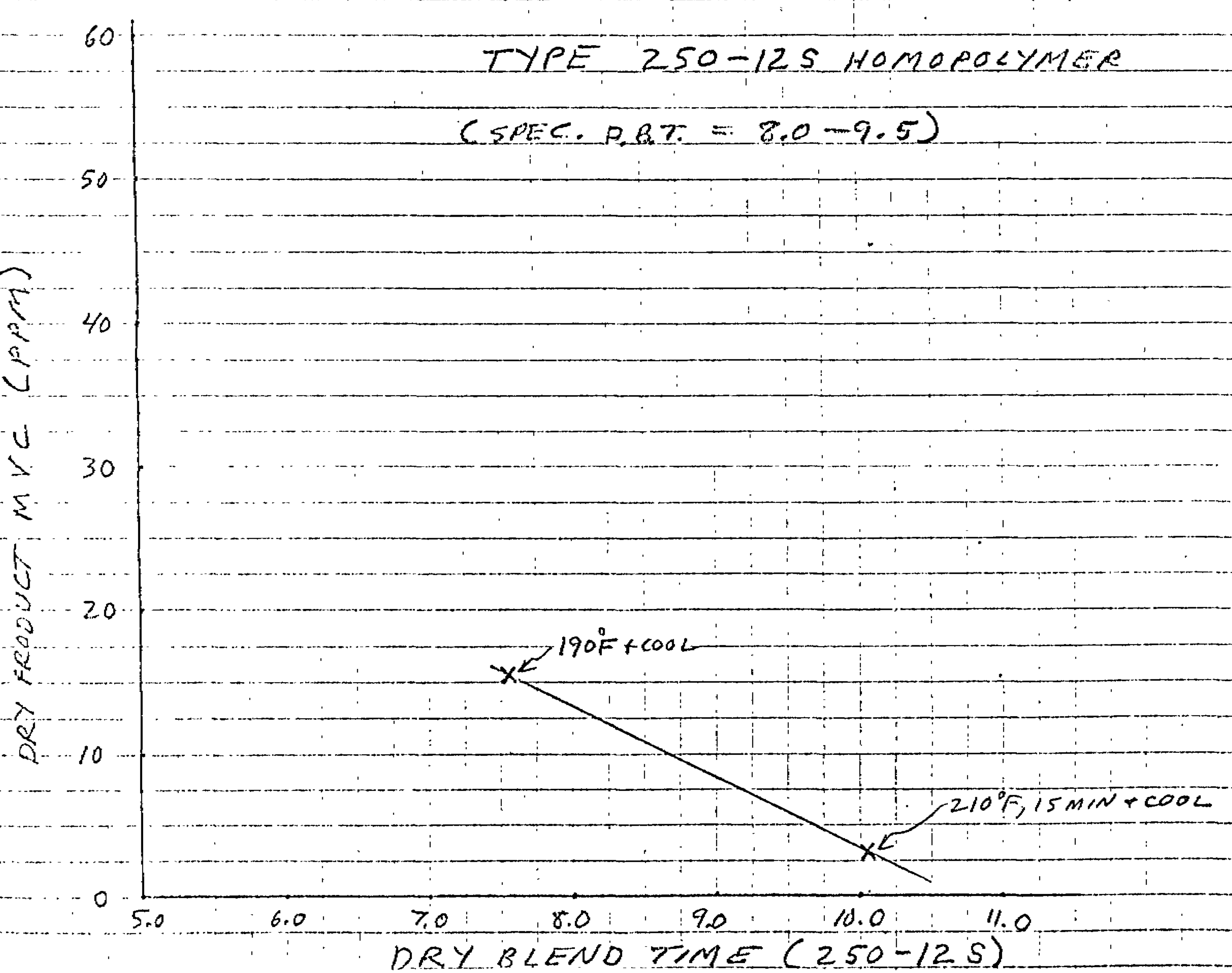

2/24/75
W. M.
RESIDUAL VCM IN RESIN
VS.
DRY BLEND TIME
TYPE 250 HOMOPOLYMER
(SPEC. DBT = 5.4 - 6.8)
165°F + COOL
180°F + COOL
165°F HOLD 30
190°F + COOL
210°F + COOL
100
50
0
5.0
6.0
7.0
8.0
9.0
10
11
DRY BLEND TIME (TYPE 250)



TYPE 25C HOMOPOLYMER
2/24/75
W.M.
STRIPPED SLURRY VCM
VS
DRY PRODUCT VCM
PRODUCT VCM (PPM)
150
100
50
0
0
100
200
300
400
501
600
700
800
STRIPPER SLURRY VCM (PPM)

2/24/75
W. M.

RESIDUAL VCM IN RESIN
VS
DRY BLEND TIME



2/24/75

SUMMARY OF RESIDUAL VCM IN DISPERSION RESINS (FRESH PRODUCTION)

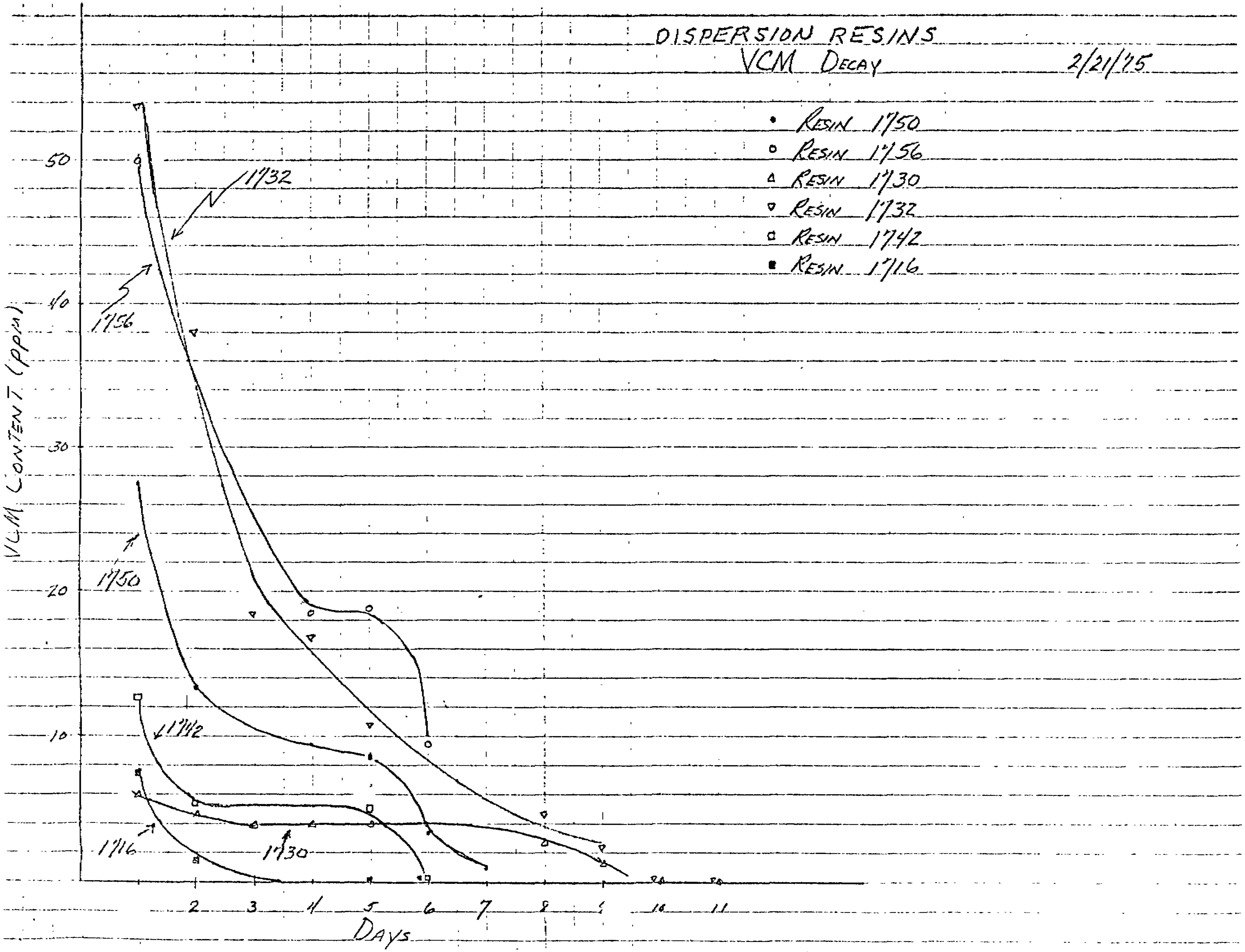
PRODUCT TYPE	NUMBER OF DATA POINTS	AVERAGE VCM (PPM)	σ	RANGE (PPM)
1716	78	4.8	6.6	ND → 29
1730	28	4.7	6.1	ND → 32
1732	17	18.8	16.9	ND → 66
1742	26	9.4	14.1	ND → 59
1755	22	13.0	16.9	ND → 72
0565	8	0.8	0.9	ND → 3.0

AGING STUDY
DISPERSION RESINS

2/21/75

RESIDUAL VCM (PPM)

DAY \ TYPE	1750	1750	1756	1756	1730	1730	1732	1732	1742	1742	1716	1716	1755
1		27.6	40	59	1.9	5.1	66	33	18	7.5	7.8	6.9	72
2	15	12	—	—	6.9	3.9	42	34	6.9	3.8	2.7	1.1	
3	—	—	—	—	3.7	4.1	22	15	—	—	—	—	
4	—	—	20	17	3.9	4.3		17	—	—	—	—	
5	7.9	9.5	15	24	N.D.	4.1	15	7.9	7.0	4.5	N.D.	N.D.	
6	5.0	2.1	8.4	11	—	—	—	—	N.D.	N.D.	N.D.	N.D.	
7	1.3	1.1			—	—	—	—					3.3
8					2.7	3.2	7.3	5.5					6.7
9					1.7	2.0	5.2	8.1					4.6
10	—	—			1.1	N.D.	5.7	4.1					N.D.
11	—	—			1.3	N.D.							



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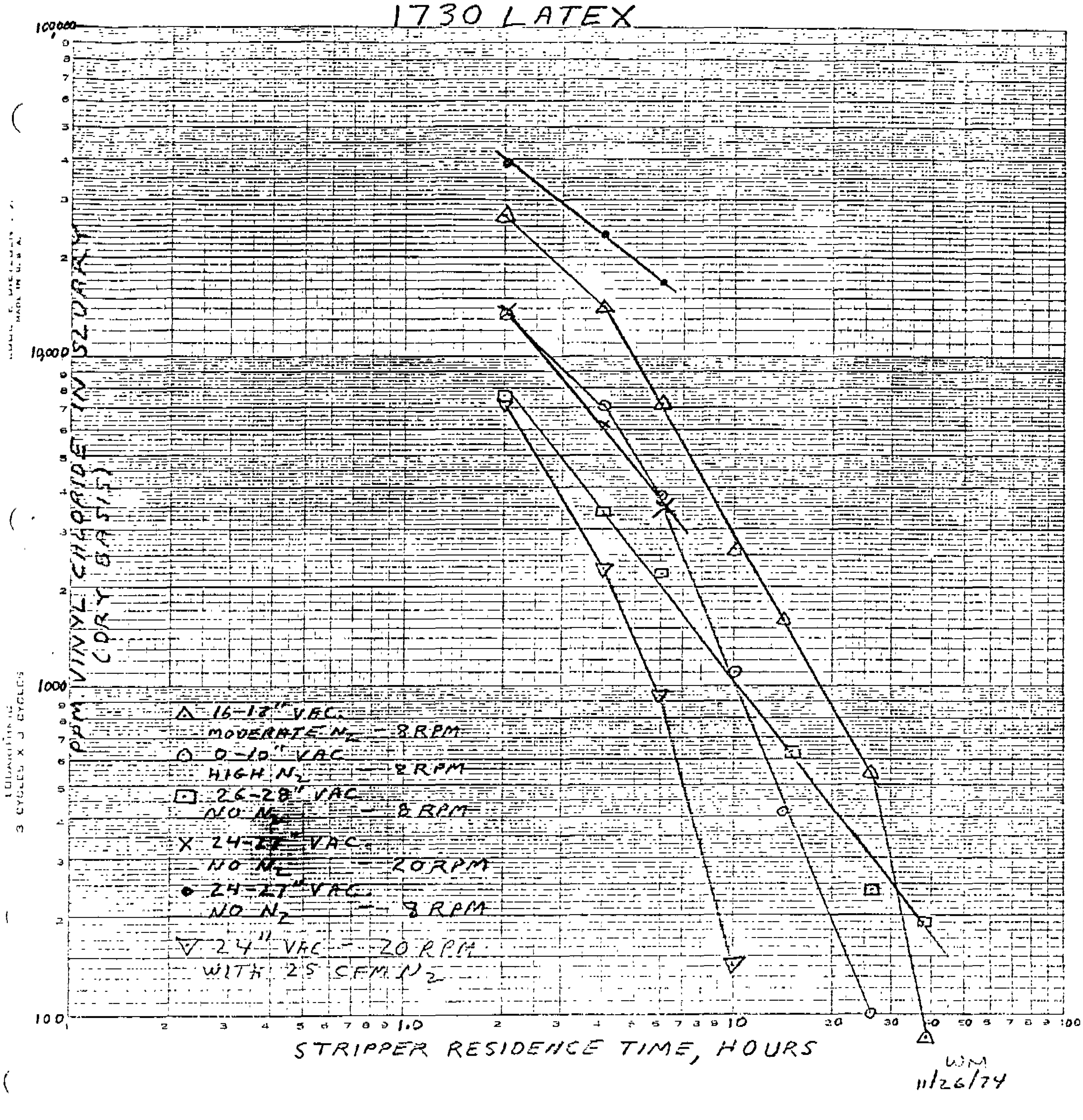
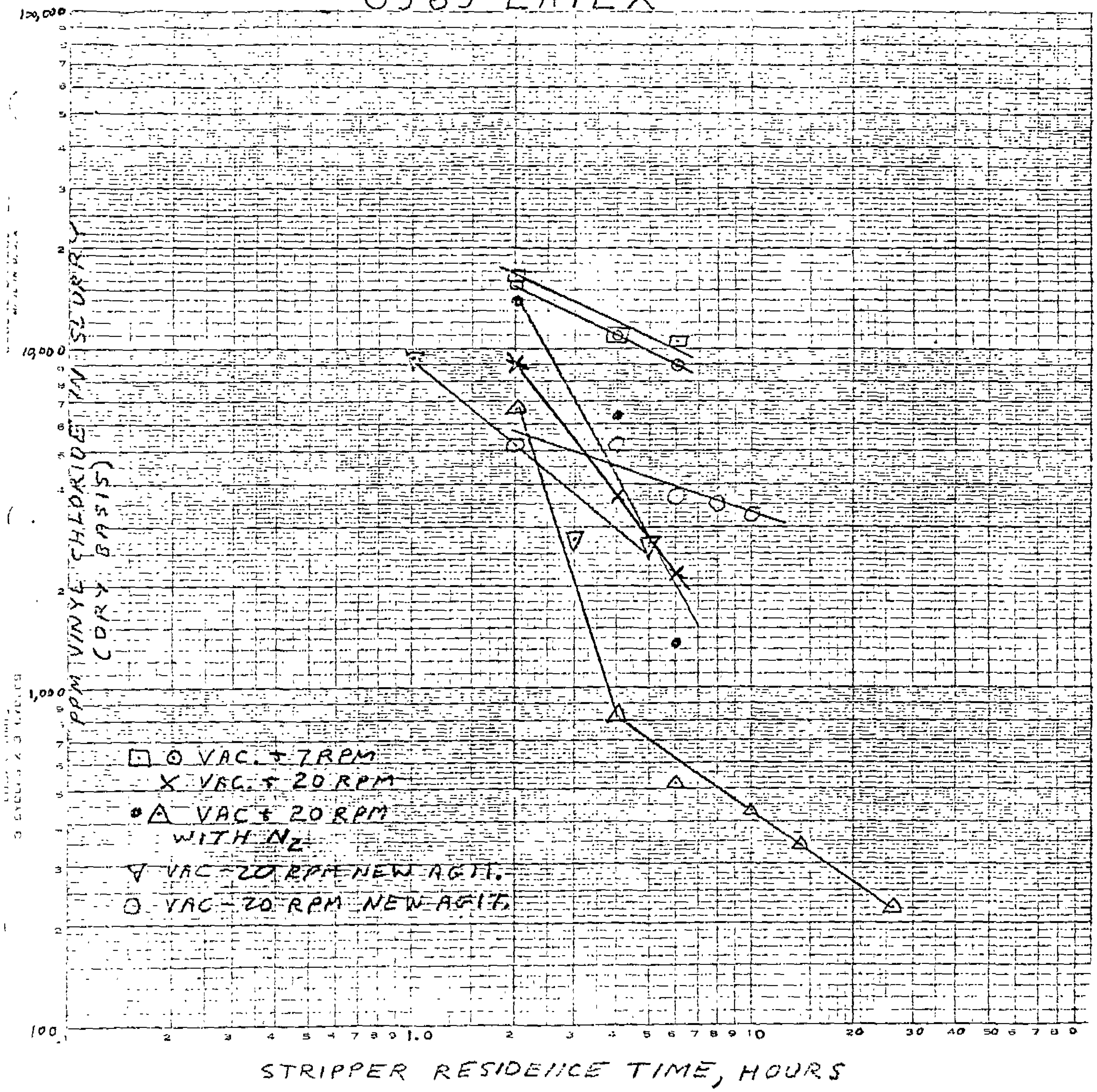
GRAPH I


1730 LATEX
PPM VINYL CHLORIDE IN SLURRY (DRY BASIS)
STRIPPER RESIDENCE TIME, HOURS
△ 16-18" VAC. MODERATE N2 — 8 RPM
○ 0-10" VAC HIGH N2 — 8 RPM
⊡ 26-28" VAC NO N2 — 8 RPM
X 24-27" VAC. NO N2 — 20 RPM
• 24-27" VAC. NO N2 — 8 RPM
▽ 24" VAC — 20 RPM WITH 25 CFM N2
WM 11/26/74

GRAPH II

0565 LATEX



WH
11/26/74

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